Solid Wastes and Residues

Conversion by Advanced Thermal Processes

Jerry L. Jones and
Shirley B. Radding, EDITORS
SRI International

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FOREWORD

The ACS SYMPOSIUM SERIES was founded in 1974 to provide a medium for publishing symposia quickly in book form. The format of the SERIES parallels that of the continuing ADVANCES IN CHEMISTRY SERIES except that in order to save time the papers are not typeset but are reproduced as they are submitted by the authors in camera-ready form. As a further means of saving time, the papers are not edited or reviewed except by the symposium chairman, who becomes editor of the book. Papers published in the ACS SYMPOSIUM SERIES are original contributions not published elsewhere in whole or major part and include reports of research as well as reviews since symposia may embrace both types of presentation.

PREFACE

Although the word "advanced" was used in the title of the symposium, the reader should be aware that the practice of producing fuel gases from wastes and residues is not new. During the early part of the 20th century, Crossley Brothers Limited of Manchester, England, was selling gasifiers worldwide that produced a low-Btu fuel gas from agricultural or forestry residues. During World War II, many vehicles in Sweden were retrofitted with low-Btu gas producers fueled by wood or charcoal. Therefore, some individuals may consider some of the research and development work described here to be refinements of previous art in the gasification field. In other cases, though, the processes have been conceived only within the last decade and are concerned with concepts new to the processing of wastes and residues. (However, some of the processes may have been applied previously to the conversion of fossil fuels such as coal or oil shale.)

The first two chapters present overview discussions of the feedstocks, technologies, and economics for producing fuels from wastes and residues. Chapter One briefly describes alternative technologies, including combustion and biochemical conversion processes, and also presents waste and residue quantities and a method for categorizing pyrolysis, thermal gasification, and liquefaction processes. Chapter Two discussed the production of low-Btu and medium-Btu fuel gases and the relative advantages to the product, which has a higher heating value.

The next section of the book contains six chapters that describe verticle flow packed-bed reactor (or fixed-bed reactor) processes. The feedstocks included in these chapters are municipal solid waste, municipal wastewater treatment sludge, scrap tires, agricultural residues (such as peanut shells, corn cobs, cotton gin trash, walnut shells), and wood residues. Two of the process reactors are designed to operate with oxygen, whereas the other four are air-blown. The reactor products described include low-Btu gas (which is immediately combusted in a secondary combustion chamber), a medium-Btu gas, pyrolytic oil, and char.

The third section contains three papers describing processes with multiple-hearth-type staged reactors. Although this type of reactor has been used for many years for ore roasting and sludge combustion, a great deal of development work has been done recently to lower the fuel requirements for processing high-moisture feedstocks, such as sludge or manure. The methods developed include operating in a starved-air combustion mode and changing the conventional gas flow between stages.

Processes with horizontal or inclined-flow rotary reactors are described in the fourth section. Municipal sludge and solid waste, wood, and scrap tires are the respective feedstocks for the three processes discussed. The products include char, electric power, carbon black, and a pyrolytic oil. All of the rotary reactors described use indirect heat transfer.

The last section of the book describes a flash pyrolysis process based on the use of a transport reactor, a new concept in fluidized-bed design, and research work being funded by the Environmental Protection Agency on topics such as a molten salt pyrolysis reactor that will accept whole (unshredded) scrap tires and a process for producing polymer gasoline from wastes. The last chapters also discuss a multiple reactor system that can produce a high methane content gas with a heating value of 370 Btu/scf, a high-pressure catalytic liquefaction process that is currently being developed to produce oil from wood, and the processing of char (from municipal solid waste) to produce an adsorbent for wastewater treatment.

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SRI International
Menlo Park, CA 94025
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JERRY L. JONES
SHIRLEY B. RADDING

CONTRIBUTORS

Alderstein, Joseph K., Syngas Inc., Suite 1260 East, First National Center, Oklahoma City, OK 73102

Bailie, Richard, Chemical Engineering Dept., West Virginia University, Morgantown, WV 26505

Beck, Steven R., Department of Chemical Engineering, Texas Tech University, Lubbock, TX 79409

Bowen, M. D., Tech–Air Corp., Atlanta, GA 30341

Brunner, Paul H., EAWAG, Swiss Federal Institutes of Technology, CH 8600, Dubendorf, Switzerland

Coffman, J. A., Wright–Malta Corp., Malta Test Station, Plains Rd., Ballston Spa, NY 12020

Cone, Eugene J., Occidental Research Corp., 1855 Carrion Rd., La Verne, CA 91750

Davidson, Paul E., Andco–Torrax Division, Andco Inc., 25 Anderson Rd., Buffalo, NY 14225

Farrell, J. B., U.S. E.P.A., MERL, Cincinnati, OH 45268

Feldman, Herman F., Battelle–Columbus Laboratories, 505 King Ave., Columbus, OH 43201

Funk, Harald F., 68 Elm St., Murray Hill, NJ 07974

Garrett, Donald E., Garrett Energy Research and Engineering Co., Inc., 911 Bryant Place, Ojai, CA 93023

Goss, J. R., University of California, Davis, Agricultural Engineering Dept., Davis, CA 95616

Hoang, Dinh Co, Garrett Energy Research and Engineering Co., Inc., 911 Bryant Place, Ojai, CA 93023

Hooverman, R. H., Wright–Malta Corp., Malta Test Station, Plains Rd., Ballston Spa, NY 12020

Huffman, George L., U.S. Environmental Protection Agency, Industrial Environmental Research Laboratory, Cincinnati, OH 45268

Huffman, William J., Department of Chemical Engineering, Texas Tech University, Lubbock, TX 79409

Jenkins, B., University of California, Davis, Agricultural Engineering Dept., Davis, CA 95616

Jones, Jerry L., SRI International, 333 Ravenswood Ave., Menlo Park, CA 94025

Knight, J. A., Engineering Experiment Station, Georgia Institute of Technology, Atlanta, GA 30332

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ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

Leckie, James O., Department of Civil Engineering, Stanford University, Stanford, CA 94305

Liberick, Walter W., Jr., U.S. Environmental Protection Agency, Industrial Environmental Research Laboratory, Cincinnati, OH 45268

Lindemuth, T. E., Research and Engineering, Bechtel National, Inc., 50 Beale St., San Francisco, CA 94119

Lucas, Theodore W., Jr., Andco–Torrax Division, Andco Inc., 25 Anderson Rd., Buffalo, NY 14225

Mann, Uzi, Department of Chemical Engineering, Texas Tech University, Lubbock, TX 79409

Mehlschau, J. J., University of California, Davis, Agricultural Engineering Dept., Davis, CA 95616

Mikesell, Ritchie D., Garrett Energy Research and Engineering Co., Inc., 911 Bryant Place, Ojai, CA 93023

Moses, C. T., Union Carbide Corp., Linde Division, Tonawanda, NY 14150

Mudge, L. K., Battelle-Northwest, P.O. Box 999, Richland, WA 99352

Negra, John S., Nichols Engineering & Research Corp., Homestead and Willow Rds., Belle Mead, NJ 08502

Purdy, K. R., Engineering Experiment Station, Georgia Institute of Technology, Atlanta, GA 30332

Ramming, J., University of California, Davis, Agricultural Engineering Dept., Davis, CA 95616

Richmond, C. A., Wheelabrator Incineration, Inc., 600 Grant St., Pittsburgh, PA 15219

Roberts, Paul V., Department of Civil Engineering, Stanford University, Stanford, CA 94305

Rohrmann ,C. A., Battelle-Northwest, P.O. Box 999, Richland, WA 99352

Schulman, B. L., Tosco Corp., 18200 West Highway 72, Golden, CO 80401

Shelton, Robert D., BSP/Envirotech, One Davis Dr., Belmont, CA 94002

Sigler, John E., 2839 Tabago Place, Costa Mesa, CA 92626

Smyly, E. D., Tech–Air Corp., Atlanta, GA 30341

Stern, G., U.S. E.P.A., MERL, Cincinnati, OH 45268

von Dreusche, Charles, Nichols Engineering & Research Corp., Homestead and Willow Rds., Belle Mead, NJ 08502

White, P. A., Tosco Corp., 10100 Santa Monica Blvd., Los Angeles, CA 90067

Williams, R. O., University of California, Davis, Agricultural Engineering Dept., Davis, CA 95616

Wong, Ho–Der, Department of Chemical Engineering, Texas Tech University, Lubbock, TX 79409

Young, K. W., Union Carbide Corp., Linde Division, Tonawanda, NY 14150

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Overview of Solid Waste and Residue Generation, Disposition, and Conversion Technologies

JERRY L. JONES

SRI International, 333 Ravenswood Avenue, Menlo Park, CA 94025

This discussion of solid wastes and residues is restricted to the materials that are predominantly organic in nature. Included are municipal refuse, scrap tires, agricultural residues, and forestry residues as well as organic sludges (mainly from wastewater treatment). The distinction between the terms "solid waste" and "residue" may be made precise by definition. Residues are defined here as solid by-products that have some positive value or represent no cost for disposal. These materials that represent a disposal cost are defined as solid wastes. These definitions, however, do not allow simple classification. For many types of materials, local market conditions are quite variable and the material may fall into both categories depending on the specific site, the season of the year, or the state of the economy.

The conversion processes to be discussed at this symposium are those that may be described as pyrolysis, thermal gasification, or liquefaction processes. The products from these so-called advanced thermal processes may be gaseous or liquid fuels, a synthesis gas for use as a chemical feedstock, or a carbon char. In some cases, a gaseous fuel generated from partial combustion containing condensible tars may be immediately combusted in a second stage. In others, a fuel product may be transported off-site for use. The process types included in the advanced processes category are not all necessarily new, but their application to the conversion of solid wastes and residues is new or has been little used in the past.

This paper provides some background on the types, quantities, and current disposition of wastes and residues as well as some information on other existing and developing processing options that are not included in this symposium. The method used to classify the processes to be discussed by the symposium speakers will also be described.

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SOLID WASTE AND RESIDUE QUANTITIES GENERATED AND DISPOSITION

More than 500 million tons (dry basis) of organic solid wastes and residues are currently generated yearly in the United States. These materials represent a potential energy resource equivalent to more than 5 quads.* Estimates of the quantities of these materials by type are summarized in Table I. The figures have been presented on a dry basis to allow comparison in consistent units. In fact, these materials vary greatly in moisture content and other physical and chemical properties (bulk density, particle size, particle size distribution, inorganic ash content).

The use of a specific material as a feedstock to an energy recovery process depends on many factors related to the cost to collect and deliver the material to the site of the conversion reactor, and the physical and chemical properties of the material that affect the conversion cost. The homogeneity of the material, for instance, often determines the extent of preprocessing required. A major cost in certain municipal refuse processing systems is the front end processing to reduce the particle size and to separate out metals and glass before the thermal conversion step. By-product wood bark, a pulp and paper mills residue, on the other hand is often quite homogeneous and requires minimal preprocessing. Such a difference in preprocessing cost may be partially offset, however, by the dumping fee paid to the conversion plant for the refuse or by sale of the recovered materials such as ferrous metals and aluminum.

The moisture content of materials may also affect the costs of the preprocessing (dewatering and drying) or fuel product separation (dehydration of a fuel gas). Where a reactor product gas contains condensible organics and requires dehydration, the dehydration operation can represent a major cost because of possible tar fog removal and water pollution problems. The moisture contents of some common solid wastes and residues are shown in Table II. It may be preferable to predry rather than to attempt to process a wet material in the reactor. (See reference 8 for a more extensive discussion of this topic.)

The disposition of the several categories of materials listed in Table I has been estimated in Tables III and IV. The estimates shown in Table I are based on an exhaustive study by SRI International, whereas those shown in Table IV are merely crude approximations. One thing is obvious from all these estimates: A large

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*A quad is 10^{15} Btu. Current yearly U.S. energy consumption is more than 75 quads.

Table I

APPROXIMATE TONNAGES OF SELECTED ORGANIC RESIDUES
AND SOLID WASTES IN THE UNITED STATES IN 1975
(Dry Weight Basis)

Type of Material	Quantity Generated (millions of dry tons)	Data Sources (references)
Agricultural residues		
crop	≥ 278*	1
manures	≥ 26*	1
Forestry residues	≥125*	1
Municipal refuse	~100	2
Industrial wastewater treatment sludges	~9	5,6
(Food processing, pulp and paper, plastics and synthetics, organic chemicals, textiles and petroleum refining)		
Municipal wastewater treatment sludges	~5.3	3,4
Scrap tires	~3	7
	≥ 546	

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*Quantity collected during normal operations or realistically collectible.

Table II

MOISTURE CONTENT OF SELECTED SOLID WASTES AND RESIDUES

Solid Waste or Residue	Moisture Content (wt %)
Wood residues	≥45
Fresh manure (with urine)	85
Rice hulls	5-10
Bagasse	~50
Municipal refuse	20-35
Wastewater treatment sludges	
Gravity thickened	95
Centrifuged	80-85
Vacuum filtered	75-80
Filter pressed (plate and frame filter)	60-70
Scrap tires	Negligible

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Table III

CROP, FORESTRY, AND MANURE RESIDUE DISPOSITION--COTERMINOUS UNITED STATES (1975)

	Tons--Dry Weight						
	Sold	Fed	Fuel	Returned*	Wasted	Total	Percent of Total
Crop	11,266,626	52,142,254	1,741,990	205,599,417	6,810,630	277,560,917	64.7%
Forestry	37,449,165	--	19,301,797	--	67,909,684	124,660,646	29.1
Manure	3,920,283	247,064	16,436	17,845,698	4,429,867	26,459,348	6.2
Total	52,636,074	52,389,318	21,060,223	223,445,115	79,150,181	428,680,911	100.0%
Percent of total	12.3	12.2	4.9	52.1	18.5	100.0%	

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*To soil

Source: Reference 1

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Table IV

ORGANIC SLUDGE AND MUNICIPAL REFUSE
DISPOSITION IN THE UNITED STATES

	Percent of Total		
Disposal Technique	Municipal Sludges	Industrial Sludges	Refuse
Landfill	25-30	≥90	85
Land application	25-30		--
Onsite lagoons	small		--
Thermal conversion*	25-30	small	10
Ocean Dumping	15	small	
Recycled	--	--	4-5
Quantity (millions of dry tons/year)	5-6	9	100

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*Small number of units with energy recovery.
Sources: References 2, 3, 4, 5.

amount of solid waste and residue material is still potentially available (at some price) as feedstock to energy recovery processes. In the case of sludge disposal, a primary consideration is not to recover energy, but rather, to dispose of the material for the lowest cost possible with minimal energy use.

ENERGY RECOVERY OPTIONS

The major process options for energy recovery from solid wastes and residues have been described previously (see Figure 1).(8) The most widely used processes for energy recovery from wastes and residues have been direct combustion in a furnace or cofiring of the materials with a fossil fuel for steam generation. A recent use of solid wastes and residues is as fuel for cement kilns. It has been predicted that by 1980, 90% of the kilns in the United States will be able to burn coal.(9) In many of these units, refuse-derived fuel (RDF) and agricultural or forestry residues can be substituted for the coal.

The more advanced thermal processes of pyrolysis, thermal gasification, and liquefaction (PTGL)* may produce liquid, gaseous, or solid char products. In some instances, the gases leaving the reactor are immediately combusted for steam generation. One advantage of this approach over conventional incineration is that the solid waste or residue can be burned with less excess air, which means lower gas volumes for cleaning (and perhaps lower particulate loading in the combustor flue gas) and a higher thermal efficiency (less heat loss out of the stack).

Biochemical conversion processes include methane fermentation (anaerobic digestion) for production of a fuel gas (up to 70% CH_4 and the remainder mostly CO_2) and fermentation of sugars to ethyl alcohol. The fermentable sugars may be produced by acid or enzymatic hydrolysis of cellulose.

This paper focuses on the advanced thermal processes; however, to provide a perspective on the applications for the newer processes, it also presents some information on the use of combustion and biochemical conversion technologies.

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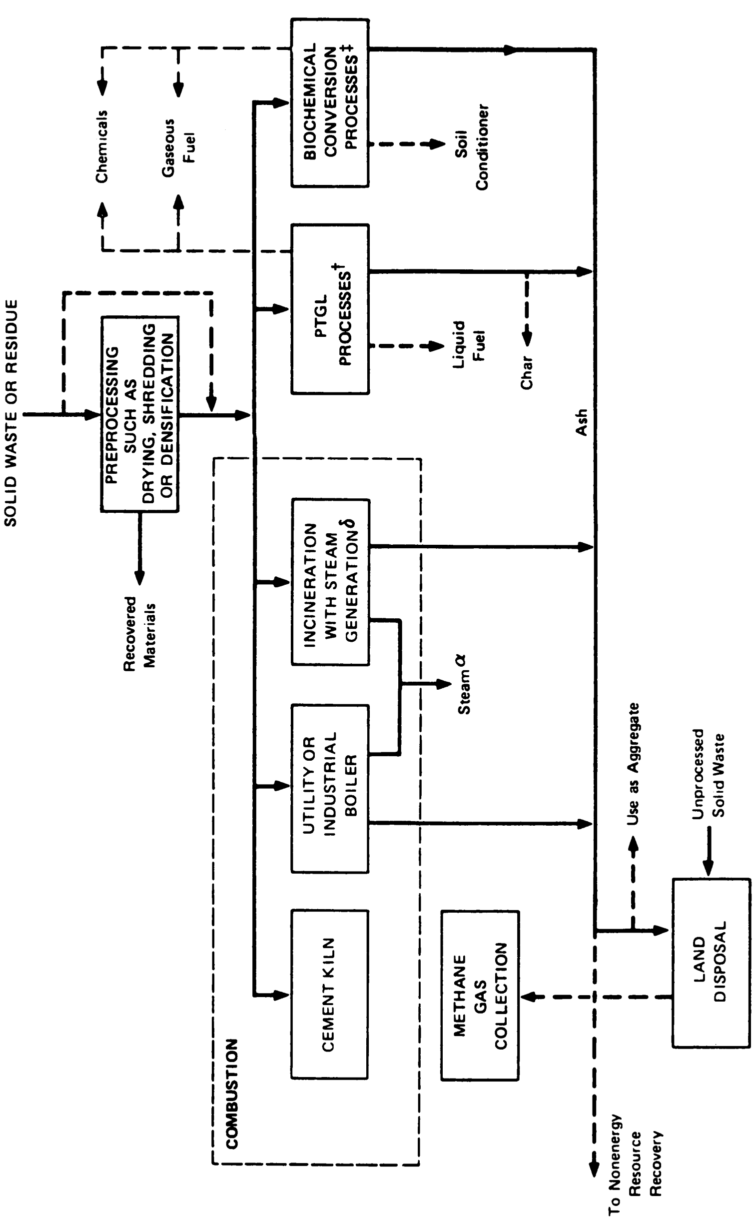
*For definition of terms see reference 8.



†PTGL: Pyrolysis, thermal gasification, and liquefaction processes

‡May be preceded by hydrolysis in case of cellulose fermentation

δRDF may be used as a supplemental fuel for sludge incineration

αFor district heating, industrial process heating, electric power production, water desalting

Figure 1. Energy recovery processes

USE OF DIRECT COMBUSTION AND BIOCHEMICAL CONVERSION

Direct Combustion

For many decades, energy has been recovered from solid wastes and residues by direct combustion. As early as 1900, numerous municipal refuse incinerators in the United States were generating steam for electric power production. These incineration units were of the refractory-walled design, which was the basic type used in this country until about 1970. Energy recovery, however, was not common with the refractory-walled incineration installations. Only about 20% of the more than 200 refractory-walled incinerator systems installed before 1965 in the United States included heat recovery boilers for steam generation.(12) Refractory-walled incinerators have become steadily less popular, and the water-wall furnace is now preferred. While capacity figures vary depending on the source, the more than 200 large refuse incinerators (more than 50 tons per day) in existence in the United States during the late 1960s(13, 14) had a combined process capacity of less than 10% of the total municipal refuse generated. By the mid-1970s, only about 160 of these refuse incinerators were operating in the United States.(15) Since 1970 at least six water-wall furnaces for refuse have been installed in the United States (16) for energy recovery.* The combined capacity of these units is approximately 1 million tons/year. These water wall units were initially developed in Europe and require only 50 to 100% excess air for combustion compared with 150 to 200% for refractory wall furnaces. This factor makes a substantial difference in flue gas volume and improves thermal efficiency. Whereas the older incinerators received a great deal of adverse publicity concerning air pollution problems, the modern units, which are equipped with high energy scrubbers, electrostatic precipitators, or fabric filters, have been able to meet stringent particulate standards.(17, 18)

Even though refuse combustion will probably continue to play a role in the solid waste management field, production of RDF for cofiring in boilers is currently receiving a great deal of attention in regions where coal is fired in utility boilers. Of the 47 electric utility feasibility studies (involving refuse use) under way in the United States during 1976, 29 (62%) were concerned with the use of shredded waste; 4 (8%), the combustion of raw waste in incinerators; 3 (6%), the use of a pelletized fuel; 6 (13%), the use of a powdered fuel; and 5 (11%), the use of a pyrolysis fuel.(19)

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*A number of projects based solely on materials recovery (not fuel) from municipal refuse are now being planned. Nevertheless, most activity in the future will focus on systems that include energy recovery. Energy sales account for 50% of the revenues for a system that also recovers aluminum and ferrous metals. The disposal fee may represent about 25% of the unit's revenue.

Another solid waste problem for many cities is the disposal of sludges from wastewater treatment plants. Sludges from wastewater treatment plants in the United States have been combusted in multiple hearth furnaces since the 1930s and in fluidized bed units since the 1960s. Currently, about 25 to 30 percent or more of the sludge generated is incinerated.(3, 4) Energy recovery from the incinerator flue gases by use of waste heat boilers has only recently become common practice, however. Co-disposal of municipal refuse and sludge in thermal conversion processes is currently a topic of great interest to municipalities.

An EPA contractor recently reported that co-incineration of sludge and refuse "will have lower overall cost than separate incineration of sludge and refuse" (20); although this approach has met with varied degrees of technical and economic success in the past, it may represent a future solution for certain sites where refuse disposal by thermal conversion processes can be economically justified. It is probably not going to be widely used in the near term, however, unless some form of government subsidies are offered to local government agencies. For a further discussion of this topic, see reference 21.

As mentioned earlier, cement kilns offer another opportunity for use of RDF because many are near large population centers. Government-supported projects are under way in the United States, Canada and the United Kingdom to determine the effect of using RDF on cement quality.(22, 23). Agricultural residues also represent a fuel source for these kilns. At Stuttgart, Arkansas, rice hulls, which have a high silica content, are now being used as a fuel for a cement kiln.(24) While use of organic residues as a fuel by a nonproducer of the residues may be new, the use of residues as fuel is not.

Combustion of wood wastes and bagasse for steam production currently represents a major energy source for the pulp and paper industry, lumber mills, and sugar cane processors.(25) The combined consumption of these residues for fuels represents close to one quad of energy for these industries.

Direct combustion of whole scrap tires for energy recovery has not been widely practiced in the United States but use of a Lucas Cyclonic Furnace for this purpose was reported in 1974 for a Goodyear Tire and Rubber Company Plant in Jackson, Michigan.(26, 27) Use of shredded scrap tires as a supplemental fuel in coal fired industrial boilers has been practiced at several sites.(27)

Biochemical Conversion

Anaerobic digestion of sludges from wastewater treatment plants has been practiced in the United States for over 50 years. According to the EPA, (28) there are currently approximately 6000 anaerobic digestion facilities at treatment plants in the United States. In the past, a portion of the fuel gas produced by these units has been flared and the main purpose of the process was to stabilize the putrescible solids. Now, more and more plants are using the product gas as a fuel for internal combustion engines to drive pumps and compressors. Such digesters, however, do not produce large quantities of fuel gas. The amount of energy in the product gas ranges from 300 to 600 Btu/day per person served by the treatment plant (approximately 2000 Btu/lb of dry solids processed).

Since the early 1970s, research has been under way to develop an anaerobic digestion process to produce methane from a mixture of municipal refuse and sludge.(29) The first demonstration plant will process about 100 tons per day of refuse and sludge and is scheduled to begin operation during 1978 in Pompano Beach, Florida, under Department of Energy sponsorship.(30) The projected net gas yield will be about 1700 Btu/lb of dry solids processed or more than 5000 Btu/day per person served by the plant.

Projects to tap the methane produced from the in situ anaerobic digestion at landfill sites have also been under way during this decade. Reserve Synthetic Fuels Company and the Los Angeles County Sanitation District are recovering methane at the Palos Verdes landfill, and the Pacific Gas and Electric Company is recovering gas from the Mountain View, California, landfill.(31)

Numerous projects are also under way to produce methane from cattle manure by anaerobic digestion. Net gas yields are equivalent to as high as 2000 Btu/lb of dry solids processed. A large plant (processing 1000 dry tons of manure solids/day) can produce approximately 4 million scf of methane per day.(32) A process development program partially funded by the United States Department of Energy is currently under way at Kaplan Industries, Incorporated, of Bartow, Florida, under the direction of Hamilton Standard (Windsor Locks, Connecticut). The process development unit will anaerobically digest the manure from an environmental feedlot with 10,000 head of cattle. Other active Department of Energy-sponsored anaerobic digestion programs are under way at Cornell University, the University of Illinois, the U.S. Department of Agriculture, and Stanford University.

Carbohydrate containing materials can also be used as a feedstock to fermentation for the production of ethanol for use as a transportation fuel. Ethanol production from grain is being actively investigated by the State of Nebraska.(33) One of the economic problems with the use of cellulosic wastes and residues as a feedstock for ethanol production is the cost for hydrolysis of the cellulose to fermentable sugars.

The conversion of carbohydrates to sugars by hydrolysis is a very active area of research, which includes enzymatic hydrolysis. Enzymatic hydrolysis has been studied actively by the U.S. Army Laboratory in Natick, Massachusetts, Louisiana State University, the University of California at Berkeley, Rutgers University, the University of Pennsylvania, the General Electric Company, and the Massachusetts Institute of Technology. Unlike enzymatic hydrolysis technology, hydrolysis of cellulose in dilute acid has been known for well over a hundred years and has been practiced commercially. Some believe that acid hydrolysis offers a more economical approach than does enzymatic hydrolysis.(34) Acid hydrolysis is currently under study at Dartmouth College, New York University, and Purdue University.

ADVANCED THERMAL CONVERSION PROCESSES

Reasons for Development

Based on the previous discussion of available conversion technologies, the first question one might ask about the advanced thermal processes (PTGL processes) is--why do we need them? Numerous answers are given but the most common ones are:

(1) PTGL processes can produce gaseous or liquid fuel products that can be burned more efficiently (low excess air) in secondary combustion chambers or used as fuel for existing combustion equipment designed to burn fuel oil or natural gas with only minor modifications required.

(2) Unlike steam generated from a combustion process, liquid or char fuel products are storable--a medium Btu gas may also be storable at a reasonable cost.

This second answer is important because at many sites, no market exists for steam, and electric power cannot be generated economically because of the small size of the electrical generator. In other cases, it is uneconomical to ship the solid waste or residue and the area where the material is located offers no market for the product fuels. Thus, the only way to utilize the material is to produce a high value shippable fuel.

The products also may also find nonfuel uses. The char material produced from some processes can be used as feedstock to produce a low cost adsorbent for wastewater treatment. This may be attractive to both municipalities and industry. The reactor product gas (which is mainly CO and H_2 from an oxygen blow reactor) can be used as synthesis gas to produce storable high value products like methanol or ammonia. Product oils may represent a source of chemicals, but because of the highly oxygenated nature of the compounds (organic acids), the oils may not be suitable for processing in conventional refining units.

Another reason often cited for developing PTGL technologies is that they may require lower investments for environmental control equipment. This statement may not be valid in all cases, as was discussed previously by this author in reference 8.

Categorization of Processes

Numerous ways are available to categorize PTGL processes, such as by:

- Feedstock Type
  - Refuse, tires, sludge, agricultural and forestry residues, etc.
- Product Distribution
  - Maximized char yield
  - Maximized fuel gas yield
  - Maximized liquid organics yield
- Product Use
  - Direct combustion of gases from reactor (i.e., furnaces operating in a starved air or partial combustion mode to supply gas to a boiler)
  - Clean fuel gas for industrial uses
  - Synthesis gas for methanol or ammonia
  - Oil for offsite use
- Specific Process Operating Characteristics
  - Slagging versus nonslagging
  - Air blown versus oxygen blown
- Fundamental Process Reactor Characteristics
  - Solids flow direction
  - Conditions of solids in reactor
  - Type of reactor vessel
  - Heat transfer method
  - Relative gas/solid flow direction

The last option appears to be the most logical approach to analyze conversion processes and is the approach used in previous work.(8, 35) The basic process categories are listed in Table 5. (Reference 8 includes drawings showing the different types of reactors).

SYMPOSIUM PAPERS

The other 22 speakers scheduled to participate in this symposium will describe a wide variety of thermal conversion process development and demonstration programs with

- Equipment sizes ranging from bench scale to commercial scale.
- Many types and qualities of feedstocks.
- Numerous product types, qualities and uses, and
- More than a dozen reactor types.

Most of the papers will include not only technical details concerning process design and testing, but also estimates of the investment and operating costs.

The papers have been organized for presentation by process reactor type as previously described. The excellent response of process developers to participate in this symposium is gratifying.

Table V

PTGL PROCESS CATEGORIES[*]

I	VERTICAL FLOW REACTORS	
	Direct Heat Transfer	Indirect Heat Transfer
	• Moving packed bed[†] (shaft furnaces) • Moving staged stirred bed (multiple hearth furnaces) • Entrained bed (transport reactors)	• Moving packed bed (shaft furnaces) • Entrained bed (recirculating heat carrier)
II	FLUIDIZED BED REACTORS	
	Direct Heat Transfer	Indirect Heat Transfer (recirculating heat carrier)
III	HORIZONTAL OR INCLINED FLOW REACTORS	
	Direct Heat Transfer	Indirect Heat Transfer
	• Tumbling solids bed (rotary kilns) • Agitated solids bed (on conveyer)	• Tumbling solids bed - Rotary calciners - Rotary vessels (recirculating heat carrier) • Agitated solids bed (on conveyer) • Static solids bed (on conveyer)
IV	MOLTEN METAL OR SALT BATH REACTORS Numerous flow and mixing options	
V	MULTIPLE REACTOR SYSTEMS Numerous flow and mixing options	
VI	BACK-MIX FLOW REACTORS For slurries and melts	

[*] Some reactors may be designed with numerous solids and gas flow regimes (countercurrent, cocurrent, split flow, crossflow).

[†] Also known as fixed bed reactors.

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

With few exceptions, the entire range of the technology is represented. Every type of process reactor listed in Table V (except the two that are checkmarked) will be discussed. The rotary kiln reactor (directly fired) is the type being used at the demonstration plant in Baltimore, Maryland. Information on that process may be obtained from many sources, such as references 36 and 37. The horizontal flow, static solids bed reactor is now being proposed only for tire pyrolysis and has not been described in the open literature.

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2

Economics Associated with Waste or Biomass Pyrolysis Systems

RICHARD BAILIE

Chemical Engineering Department, West Virginia University, Morgantown, WV 26505

C. A. RICHMOND

Wheelabrator Incineration, Inc., 600 Grant Street, Pittsburgh, PA 15219

In the early dats of the industrial revolution, wood was used as the primary fuel. In the highly developed industrial nations wood was soon replaced by fossil fuels. Wood could not compete economically with these fossil fuels. Until recent times, these fossil fuels were low in cost and available in what seemed to be an endless supply. Recently, costs have risen and are expected to rise even further in the future. It is becoming more apparent that this supply of fuel is limited. Wood remains an alternative fuel. Unlike fossil fuel, wood is not found in large amounts in concentrated areas. It is available in limited amounts spread over a wide area and is renewable.

Consideration is being given to once again making more use of wood as a fuel source. In addition to wood, all plant life (biomass) can be used for its fuel value. In the discussion to follow, wood represents but one type of biomass that may soon see use as a fuel.

During the time that has elapsed since wood was widely used, several important changes have occurred. Two of these changes are:

1. Energy consumption (both total and per capita) has increased dramatically.

2. Size of energy conversion systems have increased. These large systems have achieved "economies of scale."

In terms of electric power generation, a 50 Mw power station is extremely small and cannot achieve the thermal efficiencies and economics that are attributed to larger systems. Even this size utility would require 1,000 tons/day of biomass or 365,000 tons per year on a continuing basis. The Department of Energy has a study underway (1) to determine if this amount of fuel can be obtained on a continuing basis in a local area in the Northeast.

Wood may be burned directly to provide for generation of steam or electricity or it may be converted to a more attractive fuel form such as gas or liquid that may be transported to a central

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site where generation of steam or electricity occurs. The final judgement as to what form biomass can most effectively be used as a fuel must rest on economics. It is not likely that there is a single conversion process that will be preferred for all biomass fuels in all situations. The most common process for conversion is direct combustion of biomass. Burning of wood and bagasse to produce steam has been an established practice.

There are many situations where direct combustion may not be the most attractive alternative. This paper describes one alternative to direct combustion of biomass. Biomass may be converted to a medium energy fuel gas that is later burned. The paper makes the case for consideration of generating medium energy fuel gas, describes a system that can accomplish the conversion of gas at a moderate scale, discusses the economics of this system, and provides some examples as to how the medium energy gas compares to direct combustion.

Reasons to Consider Medium Energy Gas

If a power plant operator were given a choice of natural gas, petroleum, or coal as a fuel (all at the same cost in $\$/10^6$ Btu); natural gas would be preferred. Some of the reasons for this preference are:

1. Natural gas can be burned in an environmentally acceptable manner without air pollution control equipment.

2. The size of the furnace is smaller. Bechtel in a report for the electric power industry (2) gave the ratio of 1.0/1.35/1.85 for gas, oil and coal fuel systems.

3. There is a lower maintenance for the gas fired systems.

Figure 1 is a schematic that compares solid fuel and gas fuel boilers. The area of the furnace block represents the size. The natural gas furnace requires no pollution control equipment. Both of these factors relate directly to costs. It is cheaper to convert natural gas than solid fuel to steam or electricity in a utility boiler. In addition to these factors, favoring gas fired systems, the hog fired boilers and bagasse boilers have a lower thermal efficiency. The size of these biomass boiler systems are much larger than these for fossil fuels of the same capacity. Based on these facts, it can be seen that natural gas is much more valuable (on a Btu basis) than is biomass fuel.

Biomass may be converted to a medium Btu fuel gas. Bechtel Corporation in their report to the Electric Power Industry (2) provided curves of the thermal efficiency, the air required and the combustion products produced for gases of varying fuel values. They are shown in Figures 2 and 3. These curves show that a 300 Btu gas has a higher efficiency, lower air requirements and produce less combustion product. This indicates that a 300 Btu gas could be burned as effectively as natural gas. If a 300 Btu gas can be exchanged for a natural gas, then the statement regarding the desirability of natural gas can apply to this fuel gas. It follows that a 300 Btu gas would be preferred over biomass fuels in their natural state.

Figure 4 provides a representation of the fuel costs and all other yearly costs to obtain steam or electricity from wood, coal, or a 300 Btu gas.

The costs of operating a wood fired plant is higher than a coal fired plant. To produce electricity at the same cost, the basic fuel cost must be less as shown in Figure 4. The yearly cost of operating a gas plant is considerably less (about 1/2) than for a coal plant. If the conversion of wood to gas is equal to Δ in Figure 4 then wood to gas to electricity would produce electricity at the same price as coal.

The substitution of biomass for fossil fuel in an existing facility is seldom possible. The size of a furnace is larger for systems using a biomass fuel than for fossil fuels. Figure 5 shows a hierarchy of substitutions that are possible.

Gas may be substituted for petroleum, coal, and wood; petroleum may be substituted for coal and wood; and coal may be substituted for wood. Substitutions cannot normally be made in the reverse direction without major retrofit and/or derating of the boiler.

Natural gas is predicted to be the first form of fossil fuel to become depleted (estimates predict as little as eleven years of natural gas remain). Industry has already felt curtailment and it may be safely predicted that it will be industry that will be first to feel any effects of dwindling supplies. Experience over the last two years shows industry is the first to feel the effects of curtailment. What will happen to industrial and utility boilers built to fire gas? Many of these are relatively new and many were installed to meet environmental regulations. They are highly efficient and are in excellent operating condition, but cannot be switched to oil or coal. In such a situation a customer will be willing to pay a premium price for a gas that may be substituted for natural gas and use this facility. The alternative would be to retire the present system and commit the capital to build a new (and expensive) furnace. In such a situation, a 300 Btu gas becomes an attractive fuel.

Economics of Medium Energy Gas Production

Two processes are available that can produce a 300-400 Btu/ft^3 fuel gas from biomass that have been demonstrated in pilot plant size operations.

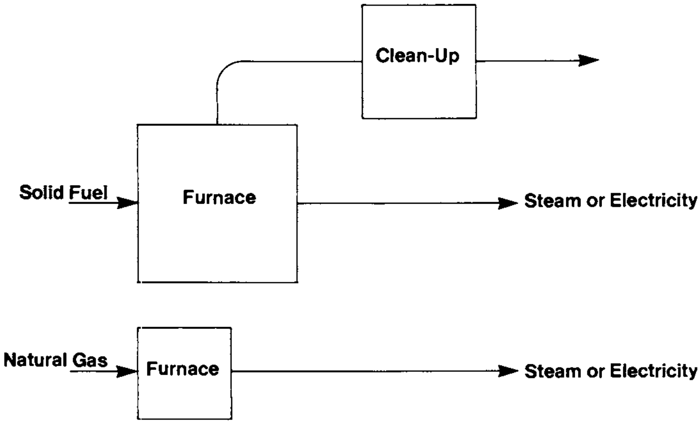

Figure 1. Comparison of solid and gas fired

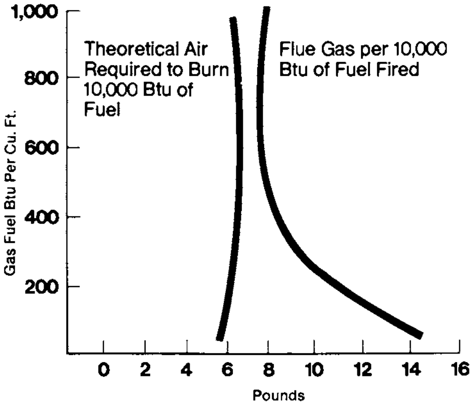

Figure 2. Theoretical air and flue gas

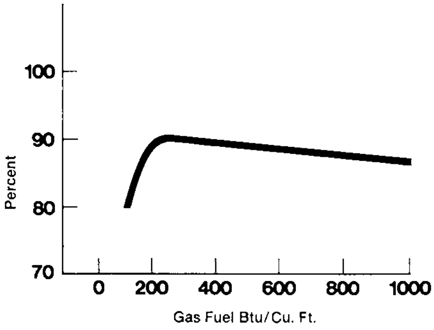

Figure 3. Unit efficiency

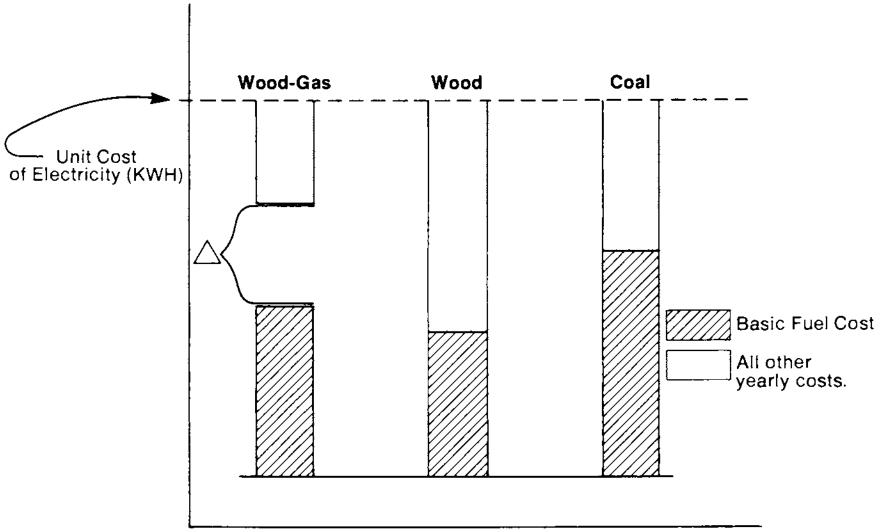

Figure 4. Cost breakdown for electricity from coal, wood, or wood–gas

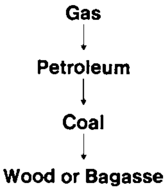

Figure 5. Fuel substitution hierarchy

1. Purox - Union Carbide has operated a 200 ton/day pilot facility in Institute, West Virginia (3). The major system components are a shaft kiln and an Oxygen plant.

2. Pyrox - Kikai Kunii has operated a 50 ton/day pilot facility in Japan. Some results have been reported by Kunii (4).

Both of these systems have used municipal waste to produce an intermediate Btu gas and the Pyrox unit has been operated on biomass. Since the organic portion of municipal waste is primarily cellulose, comparable performance should be expected from solid waste or biomass.

Both processes pyrolyze the organic portion present in the solid fuel. To produce an intermediate Btu gas, it is necessary that the gas is not diluted with Nitrogen. In the Purox Process this is accomplished by separating the Oxygen from the Nitrogen in an air separation plant. In the single shaft kiln reactor, both a combustion reaction and pyrolysis occur. In the Pyrox Process the combustion reaction and the pyrolysis take place in separate fluidized bed reactors. Solids circulate between the two beds to provide the heat needed for the pyrolysis reaction.

In the United States market, the pyrolysis process developed by Bailie and available through Wheelabrator Incineration, Inc., is comparable to the Pyrox system. This process was the one selected for this paper because the authors are more familiar with the process economics and it is more compatible with the modest sized facility needed for most biomass conversion sites. In a recent report by Battelle in a study of sugar cane as a fuel crop (5) the following comment was made:

> Commercialization of relatively small synthesis gas plants needs a process such as Professor R. C. Bailie (1972) has suggested so that the expense of an oxygen facility can be avoided. . . .The process is admittedly speculative but it strikes directly at a major drawback of the processes discussed above--the need for an energy-intensive oxygen plant.

Figure 6 is the basic schematic of the two fluidized bed process. The solid fuel is fed into a pyrolysis reactor at about 1500°F. This reactor is composed of hot inert sand. Heat transfer to the solid feed is rapid and pyrolysis occurs. The rapid heat transfer leads to high gas yields (6). The fuel gas goes to a cyclone where the char is removed. The char is fed along with air to a second fluid bed held at about 1800°F. Combustion takes place and provides the energy needed for the pyrolysis reaction. This energy is transferred to the pyrolysis bed by a sand circulation system between the two reactors.

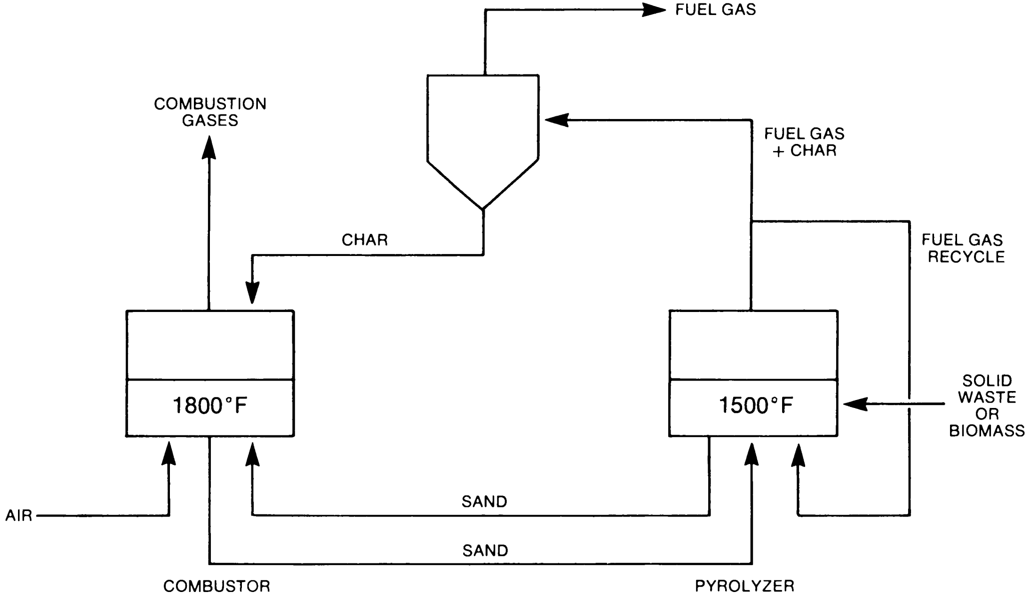

Figure 6. Schematic flow diagram of two-reactor system: temperature—1800°F, 1500°F; feed—char —unsorted waste or biomass; products—fuel gas + char + solid waste

Stanford Research Institute received a contract from West Virginia University to make an economic evaluation of the two fluid bed system to produce an intermediate Btu fuel gas from solid waste (7). Bechtel Corporation working for the City of Charleston, West Virginia, in preparing a grant application to EPA provided a second economic evaluation of the two fluid bed system (8). They confirmed the economics and process flow sheet given in the SRI study.

These studies were made in 1972 at a C. E. cost index of 135. For this paper, the costs are updated to a C. E. index of 210. The process flow sheet provided by Stanford Research Institute is divided into three sections: fuel preparation, pyrolysis, and gas clean-up and a detailed equipment list. The total capital investment in 1977 dollars is estimated to be 18.2 x 10^6 as shown in Table 1.

The annual operating costs were taken as 1.5 times the costs in 1972 or \$3.0 x 10^6/year. Using these figures, the cost of gas produced from municipal solid waste is estimated in Table 2.

The product gas composition developed by S.R.I. is given in Table 3 and the material balance in Table 4. The energy yield is defined as:

$$\text{Energy Value} = \frac{\text{Higher Heating Value of Gas}}{\text{Higher Heating Value of Fuel}}$$

was 77%.

Chiang, Cobb and Klinzing (11) have just completed a study for Resources of the Future where they estimate the cost of fuel gas from refuse using municipal waste. They based their values on the same S.R.I. study previously cited (7). These authors assumed that between 1972 and 1977, the capital costs escalated by a factor of 4 and the operating costs doubled. Under these assumptions and using economic criteria similar to "low debt economics" they calculated that the gas would cost \$6.14/10^6 Btu with no drop charge. The S.R.I. study cost estimate was made based upon sizing and costing all of the major equipment items. It would seem that the C. E. cost index between 1972 and 1977 (210/135 = 1.56) would be a more appropriate procedure to update the capital cost. If a factor of 1.56 were used in place of 4.00, the fuel gas cost becomes \$2.60/10^6 Btu which is in substantial agreement with the cost of \$2.28/10^6 Btu shown in Table 2.

In a more recent cost estimate by Battelle (5) the cost of gas producedin a two fluid bed system is given in Table 5.

In converting total capital costs to annualized capital charges the ratio of annualized capital costs to capital costs obtained from the costs of the two bed system for the pyrolysis of bagasse obtained by Battelle were used. The ratios used were: (1) 0.167 for low debt economics and (2) 0.09 for high debt economics.

Table I. Estimated Capital Investment
(1000 ton/day Dry Ash Free Municipal Waste)
Dollar Amounts in Millions

Feed Preparation	4.8
Pyrolysis	6.1
Product Recovery	2.4
Total	13.4
Utilities	2.2
General Facilities	0.8
Total	3.0
Land	.2
Start-up	.8
Working Capital	.8
Total	1.8
Total Working Capital	18.2

Table II. Annual Operating Costs and Gas Costs

	Low Debt* Economics		High Debt* Economics	
Operating Costs	$\$3.0\mathrm{x}10^6$	(1.18)	3.0	(1.18)
Annualized Capital Charge	2.8	(1.10)	1.6	(0.63)
Total	5.8	(2.28)	4.6	(1.81)

( ) \$ per 10^6 Btu

* These are the economics developed and used by Battelle in their recent study for D.O.E. and given in Appendix A.

Table III. Composition of Product Gas from Two-Reactor System (In Mole Percent)

	Dry Gas	Dry, CO_2-Free
CO	27.1%	31.7%
CO_2	14.7	0.0
H_2	41.7	48.9
CH_4	7.7	9.0
C_2 unsaturates	7.1	8.3
C_2H_6	0.7	0.9
C_3 unsaturates	0.6	0.7
C_3H_8	0.4	0.5
Total	100.0%	100.0%
Gross heating value (Btu/scf)	443	529
Yield of gas, scf/lb dry refuse	9.3	8.0

Table A-I. Values for Economic Parameters

Term	Symbol	Low Debt Economics		High Debt Economics	
		Steam Plant	Other	Steam Plant	Other
Debt fraction	D/T	0.30	0.30	0.60	0.60
Equity fraction	E/T	0.70	0.70	0.40	0.40
Interest rate	i	0.085	0.085	0.0875	0.0875
Return on equity	I	0.15	0.15	0.14	0.14
Tax rate	R	0.50	0.50	0.50	0.50
Project life, years	N	25.0	20.0	25.0	20.0
Depreciation life, years	n	25.0	11.0	25.0	11.0
Investment tax credit	P	0.24	0.24	0.24	0.24
Years for tax credit	Y	3.0	3.0	3.0	3.0
Constant for (T-W)	J	0.0142	0.0261	0.0191	0.0329
Constant for (T)	K	0.2355	0.2355	0.1645	0.1645
Constant for (F)[a]	L	0.0683	0.1052	0.0670	0.1029
Investment Tax Credit	Depreciation Schedule				
12	SL	0.0541	0.0906	0.0535	0.0889
24	SL	0.0683	0.1052	0.0670	0.1029
12	SYD	0.0729	0.1077	0.0715	0.1047
24	SYD	0.0871	0.1224	0.0850	0.1187
12	DDBSL	0.0688	0.1046	0.0681	0.1046
24	DDBSL	0.0829	0.1192	0.0823	0.1173

[a] A straight line depreciation (SL) schedule was assumed for the values presented in the upper portion of Table A-I. If a sum-of-the-year's digits (SYD) depreciation schedule and/or a 12 percent investment tax credit had been assumed the impact on the "Constant for (F)" bearing the symbol "L" would be tabulated as in the lower portion of Table A-I. Similarly, values are shown for computations using a double declining balance with a shift to straight line (in the n/2 year) depreciation schedule DDBSL in the lower portion of Table A-I. (Constants for (T-W) and (T) are not subject to change.)

Table IV. Yields from Pyrolysis of Refuse
(Dry Basis - Weight Percent)

	C	H	O	N	S	ASH	Total
Feed	30.85 wt%	3.84 wt%	22.32 wt%	(0.4) wt%	(0.1) wt%	42.49 wt%	100.00wt%
CO	8.01	--	10.68	--	--	--	18.69
CO_2	4.32	--	11.52	--	--	--	15.84
H	--	2.05	--	--	--	--	2.05
CH_4	2.25	0.76	--	--	--	--	3.01
C_2H_2	3.22	0.27	--	--	--	--	3.49
C_2H_4	0.95	0.16	--	--	--	--	1.11
C_2H_6	0.43	0.11	--	--	--	--	0.54
C_3H_6	(0.52)	(0.09)	--	--	--	--	0.61
C_3H_8	(0.35)	(0.08)	--	--	--	--	0.43
Liquids	3.45	(0.32)	(0.12)	(0.1)	--	--	3.99
Ash	--	--	--	--	--	42.49	42.49
Char	7.35	--	--	(0.3)	(0.1)	--	7.75
Total	30.85 wt%	3.84 wt%	22.32 wt%	(0.4) wt%	(0.1) wt%	42.49 wt%	100.00wt%

Note: Parentheses indicate estimated values.

Table V. Product Cost for a Plant of 1530 Tons/day

	Low Debt Economics		High Debt Economics	
Raw Material Costs (\$1.00/10^6 Btu)	7.9	(1.2)	7.9	(1.2)
Operating Costs	3.9	(0.6)	3.9	(0.6)
Annualized Capital Charge	3.6	(0.6)	2.0	(0.3)
Total (\$/10^6 Btu)	15.4	(2.4)	13.8	(2.1)

These same ratios were used throughout this paper to convert total capital cost to annualized capital costs. The basis Battelle used for these two cases is given in Appendix A.

The values provided by Battelle (5) were modified to reflect an energy yield of 80% rather than the 62% used in that study. Wheelabrator Incineration, Inc. (9) recently provided an estimate of \$2.00/10^6 Btu for producing a gas from wood waste in a plant of 250 tons capacity. In preparing the cost estimate, Battelle used the gas compositions shown in Table 6.

The cost of fuel produced is not sensitive to plant size. The cost analysis for both SRI and Battelle were for dual trains fed 500 to 750 tons/day of solid feed and there is little cost differential in sizes above 500 tons/day.

The cost estimates given above provide evidence that biomass can be converted to a medium Btu gas at a cost of \$2.00 to \$2.50 per 10^6 Btu while municipal waste will produce essentially the same quality gas for about \$1.75 to \$2.25 per 10^6 Btu. These estimates assume that the feed was delivered to the plant site. The cost at the site was zero for municipal waste and \$1.00/10^6 Btu for biomass. For municipal waste there would be a drop charge. This would decrease the cost of the gas produced. For biomass, there is a cost associated with transporting the biomass from the location it was harvested to the place where it is to be converted to a gas. The cost associated with the transportation of the biomass and fuel gas are not considered in these estimates.

System Applications

In this section several applications using intermediate energy gas made from biomass are described and some economics developed.

Application 1

The Department of Energy is supporting a study on the fueling of a 50 Mw electric plant with wood. The most straight forward approach is to burn the fuel in a power plant. An alternative to this approach is to produce a gas (to be called wood-gas) and fire this gas in a gas fired power plant. There are several advantages to this approach. These include:

a) Lower power plant costs - A gas fired power plant costs about one half as much as a solid fueled plant.

b) Higher combustion efficiency - A gas fired system uses less excess air and combustion is more complete than in a wood fired system.

c) Enviromentally attractive - A gas fired plant needs no stack clean-up to meet environmental standards.

Table VI. Estimated Composition of Output Gas -- Bailie Process from Pyrolysis of Bagasse (5)

	Mole Percent Dry Basis Bagasse
CO	23.29
CO_2	23.04
H_2	35.38
CH_4	8.53
C_2 hydrocarbons	8.66
C_3 hydrocarbons	1.10
H_2O (if wet)	(25.56)
Net Heating Value	14.96 MJ per scm (401 Btu per scf) (dry)

d) Lower maintenance - A gas fired plant has much less maintenance than a solid fueled plant.

The disadvantages include the cost of the plant needed to generate the fuel gas and the loss of heating value resulting from the additional processing.

In comparing the direct wood fired system to the wood-fuel gas-electric plant, the following assumptions were made:

1. Capital cost and operating cost of a wood fired electric generating plant is twice that of a natural gas fired system.
2. To produce 1 Kw-hr of electricity from wood requires 12,750 Btu and from natural gas is 11,000 Btu. This is for a 55 Mw system.
3. Fuel gas can substitute directly for natural gas.

A study by Mitre (9) estimated the capital cost of a 55 Mw boiler firing wood is \$55 x 10^6 and the operating cost is \$3.6 x 10^6/year. The previous section provided estimates of fuel costs from wood to be 2.1 x 10^6 Btu and 2.4 x x 10^6 Btu for low debt and high debt economics. Using these values and the assumptions given above, it is possible to compare direct wood fired systems to wood-gas fired systems. The results are summarized in Table 7.

The results shown in Table 7 indicate:

a) The cost of electricity from direct wood firing are equal to or more expensive than wood gas firing systems.

b) The cost is strongly affected by the method of financing.

c) For low debt financing electricity from wood gas is about 20% less than direct wood firing. For high debt financing the cost is comparable.

Figure 7 shows the effect of the raw material costs (basic wood cost) on the cost of electricity. The differential between the two alternatives remains constant while the percentage difference decreases with fuel cost.

Application 2

It was suggested earlier that the conversion of wood to a fuel gas would be particularly important if there was an existing facility that would be shut down if natural gas was curtailed. The industry would not only pay the cost of a new wood burning

Table VII. Comparison of Electric Costs for Wood-Electric and Wood-Gas-Electric Systems

	Wood-Electric		Wood-Gas-Electric	
	High Debt	Low Debt	High Debt	Low Debt
Fuel Costs (Wood @ $\$1/10^6$)	4.1	4.1	8.4	8.4
Operating Costs ($\$10^6$/yr)	3.6	3.6	1.8	1.8
Annualized Capital Costs	4.8	8.5	2.4	4.2
	12.5	18.2	12.6	14.4
Mill/Kw-hr	(40)	(58)	(40)	(46)

facility, but would be forced to pay the annualized capital cost for the idle natural gas plant. Figure 8 shows the electric costs for a new wood fired plant and for wood-gas used in an existing natural gas plant as a function of wood fuel cost. The only difference in this figure and that shown in Figure 7 is the annualized capital cost for the idle gas plant is added to the cost of a new wood burning plant. For either financing scheme, the wood gas plant provides electricity at a significantly lower cost than the direct fired system. For the case where the basic wood fuel costs are $\$1.00/10^6$ Btu, the differentials are 15.6 and 35.7%, respectively.

Application 3

The economies of scale are significant in utility power stations. The Mitre report (10) provides cost estimates for plants using 850, 1700 and 3400 tons of wood per day. These values were converted to annualized capital costs in the same manner as previously described. In comparing costs it was assumed that the cost of the wood to gas conversion facility did not benefit from economies of scale. This is not true for other types of conversion facilities. For example, the Union Carbide Process benefits greatly from an increase in scale. This is because the economies resulting from the larger oxygen plant needed in this process. There are some small economies of scale that would lower the curves for the wood gas systems at larger capacity.

The curves in Figure 9 reflect the increase in trasnportation costs for larger plants. In evaluating the transportation costs the following assumptions were made:

1. Cost of transportation is $0.10 per ton mile.

2. The average distance traveled for various size plants were:

850 tons/day	35 miles
1700 tons/day	46.7 miles
3400 tons/day	66 miles

In Figure 9 the cost of electricity in direct fired plants are seen to cross the wood-gas fired curves for high capacity. This is a direct result of not benefiting from economies of scale for the wood-gas generators. Since there is no cost advantage in larger gas generating facilities, it would be more attractive to locate the plants closer to the wood source and pipeline the gas to a large central gas fired utility. This would reduce the distance the wood is hauled and the cost associated with moving the wood to a central plant. This would reduce the cost below those shown in Figure 9 for the gas fired systems. In Figure 9 the wood-gas curves do not reflect either the small but real economies

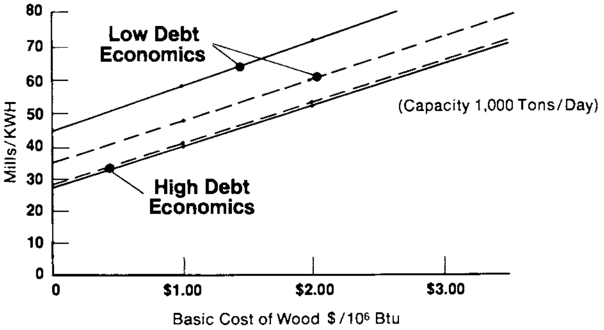

Figure 7. Electric cost comparison between wood fired and wood gas fired: (——) wood fired; (– – –) wood/gas fired

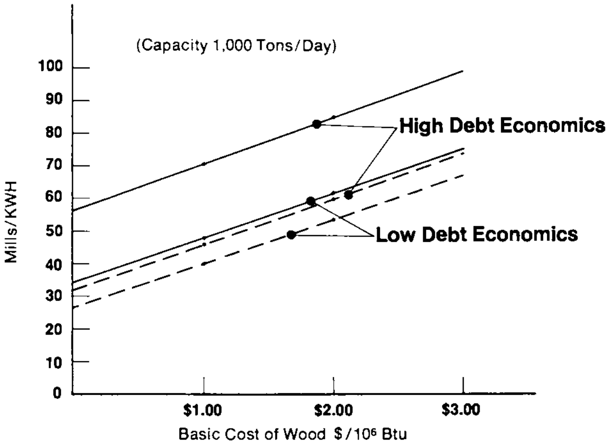

Figure 8. Cost comparison of electricity between firing wood gas in existing plants and building new direct fired plants: (——) new wood fired plants; (– – –) existing plant fired with wood gas

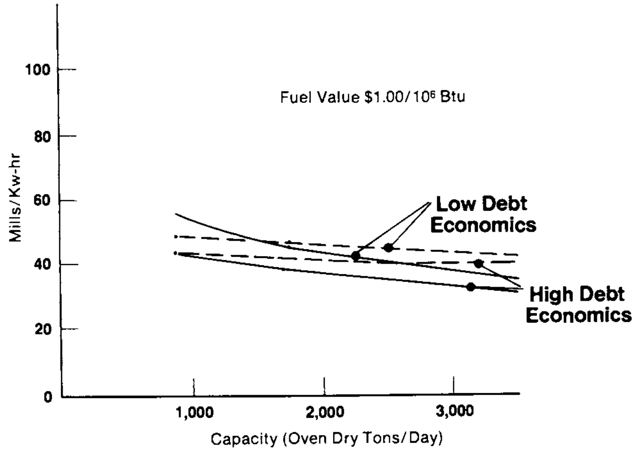

Figure 9. Economies of scale

of scale or the lower transportation costs. The situation for wood gas would be better than that shown in Figure 9.

Application 4

The final hypothetical situation considers the use of municipal waste to provide a fuel gas to a large mall, group of commercial establishments or industrial park. Table 2 provided estimates for the cost of fuel gas from municipal waste. In this table no charge was made for the fuel. For the case of municipal waste, a drop charge is imposed in order to leave the waste at the facility. Eight dollars per ton is considered to be a reasonable charge. If four dollars of this could be assigned to subsidize the waste-to-gas process, a credit of $0.79 per 10^6 Btu is realized. This reduces the gas cost to $1.06 to $1.53 per million Btu. At this price it becomes a highly competitive fuel. This fuel gas can be transported by pipeline to the user to burn in commercial gas-fired equipment. This becomes important because the comparative capital cost between gas fired equipment to solid fired equipment is often as low as 1/6 for small commercial units.

Conclusions

1. Biomass and municipal waste can be converted to an intermediate energy gas.

2. The gas produced in the twin fluidized bed can substitute for natural gas in existing equipment.

3. The energy lost in the biomass to gas conversion system is largely made up in higher combustion efficiency in the power plant.

4. The cost of the wood-gas is in the price range of $2.00 to $3.00 per 10^6 Btu.

5. The economic choice between direct fired or wood gas fired is highly dependent on the method of financing. If there is an advantage, it would appear to be with wood-gas.

6. Wood gas systems producing intermediate gas have not seen commercial operation and has no track record. Direct fired systems are operating commercially but have a spotty record.

7. When operating natural gas fired unit must be replaced by a wood fired unit, it becomes more attractive to convert wood to gas and use the existing facility.

8. For power generating systems, economies of scale are significant for direct fired systems. The wood gas to electric system is less sensitive to scale because the wood gas generating system is insensitive to scale. Multiple units are built for large size applications. For the Union Carbide Purox system economies of scale are achieved and this conclusion would not apply if this unit were selected to produce wood-gas.

9. Municipal waste can produce a fuel gas at a lower cost because there is no charge for the feed and the system will receive a subsidy through a drop charge.

The paper has emphasized wood which was selected to represent biomass. Other biomass sources may also be used. The fluidized bed system will more readily accept other biomass materials with differing physical characteristics than the Purox system or direct fired boilers. If it is not possible to provide assurances of the economic advantages that may be obtained from wood gas there is little doubt that the economics of today show this option in a more favorable light than the economics of a few years ago.

The longest coal strike in history is underway. It shows how dependent the nation is on coal. The cost of coal will without question rise significantly under any new contract. This will see a further shift toward favorable economics for biomass utilization.

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The Andco-Torrax High-Temperature Slagging Pyrolysis System

PAUL E. DAVIDSON and THEODORE W. LUCAS, JR.
Andco-Torrax Division, Andco Incorporated, 25 Anderson Road, Buffalo, NY 14225

The Andco-Torrax system is a new high temperature refuse conversion process known technically as slagging pyrolysis. Although the pyrolysis of solid waste is a relatively new innovation, pyrolysis processes have been used for years by industry. Typical examples are found in the production of char and methanol from wood and coal gasification. Pyrolysis, commonly referred to as destructive distillation, is defined as an irreversible chemical change brought about by the action of heat in an oxygen deficient atmosphere. The pyrolysis of organic material causes the volatile fraction to distill, forming combustible liquids and vapors. The vapors are composed primarily of methane, hydrogen, carbon monoxide, carbon dioxide, water and the more complex hydrocarbons such as ethane, propane, oils and tars.

The exact components in percent composition of these gases formed by pyrolysis of municipal waste cannot accurately be predicted in that, in a real system, the complex multi-component fractions would be converted to more stable gases such as ethylene through a continuing pyrolysis action and are a result of complex time/temperature kinetic reactions. The material remaining after pyrolysis is char, a charcoal-like substance consisting primarily of fixed carbon residue. In the Andco-Torrax system, the char remaining from the pyrolyzed material is burned to carbon monoxide and carbon dioxide using high temperature air, thus releasing sufficient heat energy to convert all non-combustibles contained in the solid waste to molten slag and to further pyrolyze incoming waste material.

PROCESS DESCRIPTION

The principal components of the Andco-Torrax system are the gasifier, secondary combustion chamber, primary air preheating equipment, waste heat boiler, and gas cleaning system. These components are shown in Figures 1 and 2. These figures illustrate the primary air preheating system as a regenerative tower system, although other alternatives may be used which could include a heat recuperator or a fossil fuel fired air preheater. The gasifier and secondary combustion chamber comprise the main process

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In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

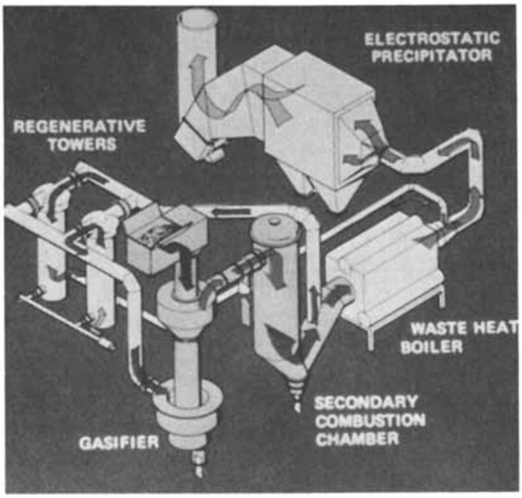

Figure 1. Andco–Torrax system

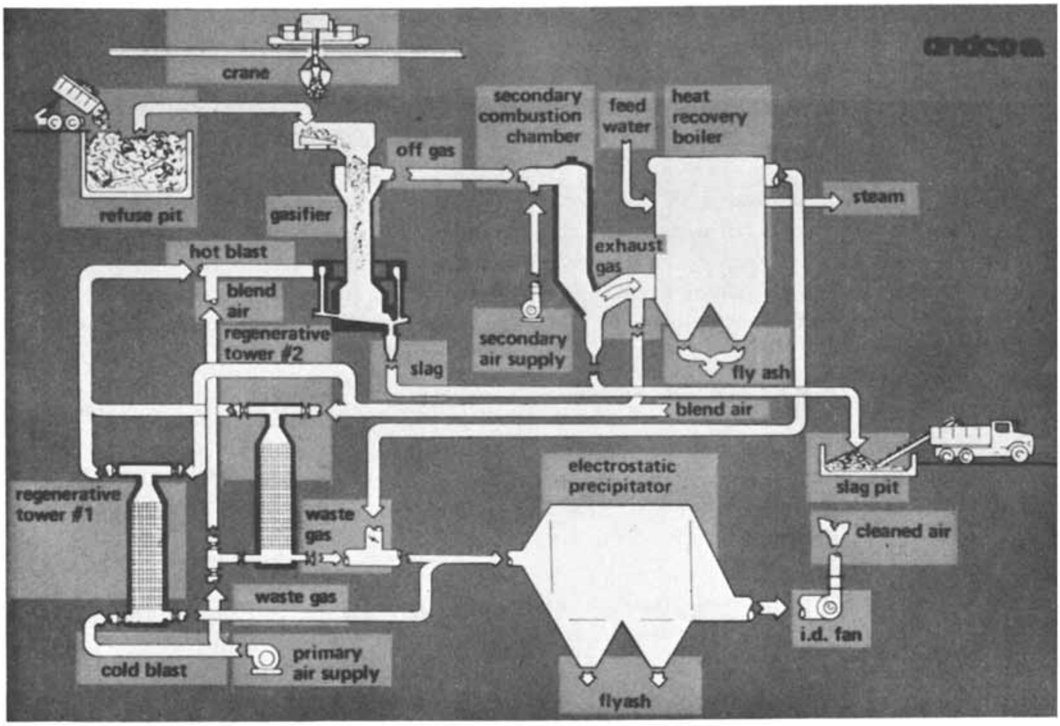

Figure 2. Andco–Torrax system schematic

components and are shown in Figure 3.

Gasifier

A crane with grapple bucket removes refuse from the refuse pit without pretreatment other than the shearing of oversize items to approximately one meter dimensions. The refuse is directed into a hopper from which it is fed into the top of the gasifier by either a reciprocating ram or a vibrating feeder. The gasifier is a vertical shaft furnace so designed that the descending refuse burden and the ascending high temperature gases become an effective countercurrent heat exchanger. The gasifier shaft is 12–15 meters (40'–50') in overall height, 1.8–2.7 meters (6'–9') in diameter, and is fabricated from mild steel with refractory lining in the lantern and hearth areas only. The main stack between the lantern and hearth is externally cooled over its length by a jacketed water cascade. The hopper has a loose sealing mechanism to decrease the infiltration of air into the top of the gasifier. As the refuse descends in the gasifier, three distinct zones are encountered - drying, pyrolysis, and primary combustion. The function of the drying zone is to evaporate the moisture in the refuse and to act as a plug to further restrict the in-flow of air during charging. Refuse entering at the top of the gasifier moves downward past the gas offtake plenum (lantern). From this point down, the refuse dries and then, in the pyrolysis zone, is heated from 260–1093°C (500–2000°F) in a reducing atmosphere where the rate of decomposition to pyrolysis products increases with temperature. Various oils formed in the low temperature region of pyrolysis continue to pass down and are cracked into gases and char at the higher temperatures. These particles of char and oil that are entrained in the hot gases are to a large extent scrubbed out by the descending refuse and are recycled down into the higher temperature zone.

The heat for drying and pyrolyzing the refuse is supplied by the combustion of the carbon char with preheated air in the primary combustion zone. The preheated air at 1037°C (1900°F) is directed from the regenerative towers through a hot blast main into a circular bustle pipe. The air then passes through several downcomer–tuyere assemblies radially into the gasifier hearth where it is used to combust the char. The heat generated by this combustion process (up to 1650°C or 3000°F) also transforms the non–combustible materials to a molten slag. The temperature profile through the refractories results in the formation of a coating of solidified slag over the refractories in the hearth area. This coating assists in protecting the refractories from the high temperatures and corrosive action of the molten slag. The molten slag is drained continuously through a sealed slag tap into a water quench tank to produce a black, glassy, sterile aggregate. The sealed slag tap arrangement allows removal of the frit while maintaining pressures in the hearth slightly above atmospheric. The slag formed in the hearth of the gasifier and in the secondary combustion chamber accumulate in a common slag agitation/holding tank, and several methods are available for transporting it from the tank. At the demonstration plant in Orchard Park, the material flowed into a slag pit and a front-end loader was used to manually carry it to a stockpile for later removal by truck. One automatic method uses drag and belt conveying equipment, with drag conveyors removing slag from the main slag pit and depositing it at a continuous rate on horizontal or inclined belt conveyors feeding a vertical bucket elevator. The bucket elevator directs the material to external storage bins or stockpiles for removal by transportation vehicles. Another method used is slurry pumping of slag from the main slag pit to dewatering tanks. The suspended solids quickly settle out in the dewatering tank for removal by grapple bucket or drag conveyor system. The conveying liquid may be pumped back through the system from the dewatering tanks for reuse in the quench tanks.

Volume reduction from the raw refuse to the slag residue is approximately 95-97 percent and the weight reduction is 80-85 percent, depending on the inert fraction in the refuse. Table I presents the physical and chemical properties of the slag residue produced at the Orchard Park demonstration plant.

At least 90% of the energy content of municipal refuse is contained in the gas stream which leaves the gasifier. This energy is in the form of combustible gases, vapors, and entrained particles and as sensible and latent heat. The temperature of this gas is approximately 400°C-500°C (752°F-932°F). The complete combustion of this gas stream produces about the same volume of products of combustion per unit of heat released as would be the case with other gaseous fuels. The composition and properties of the combustible gas stream are dependent, of course, on the refuse mix. The heating value of the gas will normally be in the range of 937-1593 kcal/nm^3 (100-170 Btu/scf).

Although this combustible gas stream, with or without cleaning to remove entrained material, has potential application as an energy source where close coupling of the gasifier with the recipient combustion device (such as kilns, boilers, etc.) is possible, the present commercial Andco-Torrax plants employ a secondary combustion chamber to burn this gas to completion.

Secondary Combustion Chamber

The offgas exits from the gasifier via the lantern, and is drawn into the secondary combustion chamber through a refractory lined cross-over duct by the system's induced draft fan. At the inlet to the secondary combustion chamber, the offgas is mixed with air in a high energy burner and is admitted to the chamber in a tangential fashion for a high turbulence, spiral flame. A small quantity of excess air is normally used to maintain the exit temperature from the secondary combustion chamber between 1150°C and 1250°C (2100-2280°F). The secondary combustion chamber is a vertical, refractory lined vessel in which temperatures up to 1400°C

TABLE I - SLAG RESIDUE CHARACTERISTICS

Constituent	Average % (by weight)	Range %
SiO_2	45	32.00 - 58.00
Al_2O_3	10	5.50 - 11.00
TiO_2	0.8	0.48 - 1.30
Fe_2O_3	10	0.50 - 22.00
FeO	15	11.00 - 21.00
MgO	2	1.80 - 3.30
CaO	8	4.80 - 12.10
MnO	0.6	0.20 - 1.00
Na_2O	6	4.00 - 8.60
K_2O	0.7	0.36 - 1.10
Cr_2O_3	0.5	0.11 - 1.70
CuO	0.2	0.11 - 0.28
ZnO	0.1	0.02 - 0.26
Trace Oxides	1.1	--
	100.0	

Dry Bulk Density..........1.40 gm/cc
True Residue Density...... 2.80 gm/cc

(2550°F) are realized, and where sufficient residence time, up to 1.7 seconds, is maintained to assure complete combustion. The particulate matter entrained in the offgas from the gasifier is burned and the inert portion is fused, melted and slagged out of the stream. This slag is also water quenched and is approximately 10% of the total residue produced in the process.

Air Preheating System

As discussed previously, there are a number of possible methods for the preheating of primary combustion air by utilizing the energy from the hot combustion products which exit from the secondary combustion chamber or by burning fossil fuel. One such equipment system which has been built at both the SIDOR (Luxembourg) plant and the Frankfurt plant is shown in Figures 1 and 2. This system employs two regenerative towers, a system successfully used for many years in the steel and glass industries for preheating air. Regenerative towers are refractory filled steel vessels containing a high heat capacity refractory with a multiplicity of vertically aligned flues which readily absorbs the heat from the hot products of combustion passing through them. During the heating cycle, approximately 10% of the total products of combustion from the secondary combustion chamber are introduced into the top of the regenerative towers. The heat from these gases is transferred to the refractories, bringing them up to temperature levels of approximately 1150°C (2100°F) at the top and 260°C (500°F) at the base. The waste gas exiting the regenerative tower subsystem is returned to a duct at the inlet of the gas cleaning system. A modulating damper valve in the boiler exit duct controls the amount of gas used in heating the regenerative tower refractory.

During the "blast" cycle, the combustion products from the secondary combustion chamber are diverted to the second regenerative tower to heat its refractory. Ambient process air is introduced at the base of the fully heated regenerative tower and passes up through the refractory absorbing the stored heat. The exit temperature of the air from the tower ranges from 1037°C (1900°F) to 1110°C (2030°F). A constant primary air, or blast, temperature is maintained by blending the heated air with ambient air before introduction into the gasifier.

A second method for preheating the primary combustion air employs a heat recuperator as used in the Grasse and Creteil plants. The metallic recuperator recovers heat from a portion of the products of combustion from the SCC exit and produces preheated air at a temperature of 600°C. The additional 400°C temperature differential is achieved through the use of a supplementary oil or gas fired burner.

A third method for preheating the primary combustion air employs a fossil fuel fired silicon carbide cross flow shell-and-tube heat exchanger similar to that used in the demonstration plant in Orchard Park, New York. While this unit proved reliable during operation, the cost and availability

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of liquid or gaseous fossil fuels makes such a unit unattractive for commercial plants. Using a regenerative or recuperative type of heat exchanger results in both energy and cost savings.

Waste Heat Boiler

With regenerative towers as the method of air preheating, approximately 90% of the gas flow exiting the secondary combustion chamber is directed to the energy recovery system, a waste heat boiler. Typically, the waste heat boiler is of the combination radiation-convection type and is designed with certain Andco-Torrax process characteristics in mind. These characteristics are high temperature inlet gases (from 1150°C to 1250°C), constant temperature input, homogeneity of inlet gases and absence of unburned material, low volumetric gas flow, and low particulate loading. Gases leave the boiler at approximately 290°C (554°F).

Gas Cleaning Equipment

At the exit of the waste heat boiler, the cooled waste gases from the regenerative towers (or alternative preheating systems) are combined with the exiting flow from the waste heat boiler and are ducted to the gas cleaning system. The gas cleaning system is typically a hot gas electrostatic precipitator of conventional design which can effect an outlet particulate loading of less than 100 gm/nm^3 (.04 grains/SCF).

DEMONSTRATION PLANT

The development of the Andco-Torrax process began in early 1969 with the formation of a company called Torrax Systems, Inc. which was jointly owned by Andco Incorporated and the Carborundum Company. The purpose of Torrax Systems, Inc. was to acquire and develop all necessary technology for a new type of high temperature slagging pyrolysis process for disposal of municipal solid waste and then to make the necessary arrangements to build a prototype plant and to operate it successfully to prove out the technology. On July 1, 1969, Torrax Systems, Inc. entered into a contract with the County of Erie, a municipal corporation of the State of New York, to build and operate a 75-ton-per-day demonstration plant in Orchard Park, a suburb of Buffalo. This new plant program was made possible largely through the support of the U.S. Environmental Protection Agency acting under the Solid Waste Disposal Act of 1965 which provided a program of demonstration grants for the development and evaluation of new solid waste disposal technology. Other funds for the program were provided by New York State, Erie County, the American Gas Association, the Carborundum Company, and Andco Incorporated. Ground breaking ceremonies for the demonstration plant were held in July of 1970 and operation commenced in the second quarter of 1971. Between 1971 and 1973, the plant operated as an engineering development facility to evaluate and prove system design features. During that period of time, the plant operated for 2316 hours and processed 7664 tons of municipal solid waste. The longest period of continuous operations was 120 hours. Although the plant design capacity was 3.1 tons per hour, the equipment operated at throughput rates as low as 1.3 tons per hour and as high as 4.7 tons per hour.

In addition to normal municipal refuse, the Andco-Torrax demonstration plant handled other wastes, particularly those from industry. These wastes were mixed with municipal refuse and minor changes were made to the equipment and the operating procedures.

Some of the tests at the demonstration facility included the following:

Sewage Sludge. Undigested sewage sludge with 78 percent water content was charged with the municipal refuse in quantities averaging 28.5 percent of the total 3.8 TPH charge.

Waste Oil. Waste automotive lubricating oil was charged with municipal refuse in average quantities of 6.1 percent of the 4.4 TPH charge.

Combined Sludge and Oil. A test combining sewage sludge with waste oil and municipal refuse was accomplished. The total refuse rate was 3.4 TPH of which 30.1 percent was sludge and 3.0 percent was oil.

Tires. Unshredded automotive tires were charged by bucket with the normal refuse. The average addition was 30 tires per hour or about 10 percent of the total 3.3 TPH consumption.

Polyvinylchloride (PVC). Bags filled with PVC plastic waste were charged with municipal refuse in quantities averaging 7 percent of the 3.2 TPH charge. In all cases no significant changes to the process operation were encountered.

In these tests, changes in the process parameters (flows, temperatures, gas composition) were evident and were related to the changes in the input refuse heating value and composition.

The Orchard Park plant employed equipment which for the main part was similar to that used in commercial Andco-Torrax plants. However, as discussed, the demonstration plant used a natural gas fired heat exchanger, rather than regenerative towers, to supply the high temperature primary combustion air. Other differences between the demonstration plant and commercial plants include the following:

- Refuse was charged into the top of the gasifier through an open cone without any type of hopper feed.
- The residue was removed from the slag quench area with a front end loader rather than automatically as is the case in commercial plants.
- A wet scrubber was used rather than an electrostatic precipitator.

MARKET DEVELOPMENT

It was decided by Carborundum and Andco in early 1973 that the original objectives of Torrax Systems, Inc. had been successfully concluded. Since then the demonstration plant has only operated for sales demonstration purposes. In 1973, agreements were executed between Carborundum and Andco giving Andco commercial rights to the process in Canada, Australia and Europe. A further agreement was executed in 1976 which expanded Andco's commercial rights to the process to include the United States, the USSR, Japan, New Zealand, Indonesia, and countries in Polynesia, Latin America, Africa, and the Middle East.

Commercial Plants

Orders for four commercial Andco-Torrax systems have been received to January 1978. One order was placed in 1975 with Antox Wurth, Andco's West German licensee, by the City of Frankfurt. The unit capacity is 200 metric tons per day (220 TPD) and the system is located within an existing conventional incineration facility. Another order was placed in 1975 with Caliqua S.A. by the municipality of Grasse, France. The construction of the new 168 metric ton-per-day (185 TPD) plant is under the authority of Caliqua for use by the "Syndicat Intercommunal pour le Traitement et l'enlevement des Ordures Menageres et des Dechets Urbains de la Region de Grasse." The Grasse plant is currently undergoing its initial start-up phase and the Frankfurt plant construction will be complete in the first quarter of 1978. A third plant, with a capacity of 400 metric tons per day (440 TPD), is being built in Creteil, France, and will be completed in 1979.

The first order for a commercial Andco-Torrax System was awarded to S. A. Paul Wurth, Andco's Benelux licensee, in January 1974. This system has a 192 metric-ton-per-day (211 TPD) capacity, and is owned and operated by the Syndicat Intercommunal pour la Destruction des Ordures des Cantons de Luxembourg, Esch, et Cappellen, (SIDOR), a group of 35 municipalities in Luxembourg. Process design and sizing of the hardware components for this system were based largely on prior demonstration plant experience gathered at the Orchard Park facility, correlation with specific refuse analysis data for Luxembourg, and utilization of a process simulation computer program. The design point refuse composition, expressed in weight percent, for the SIDOR Andco-Torrax System is:

Refuse Composition		Proximate Analysis		Ultimate Analysis	
Combustibles	51.4	Fixed Carbon	9.0	Carbon	29.3
Water	23.7	Volatiles	42.4	Hydrogen	3.6
Ash-Inerts	24.9			Oxygen	18.5

This refuse analysis corresponds to a lower heating value of 2500 kcal/kg (4500 Btu/lb). Although the plant was designed for 2500 kcal/kg refuse, actual refuse quality as received is closer to 1600 kcal/kg (2880 Btu/lb). Mass and heat balances for these two operating points are presented in Table II. The differences in system efficiencies and steam production per unit of refuse for these two points are noteworthy.

The over-all SIDOR plant consists of two conventional grate-type incinerators built by CNIM (Constructions Navales et Industrielles de la Mediteranee) and one 192 ton-per-day Andco-Torrax unit built by S. A. Paul Wurth. The Andco-Torrax equipment includes the refuse feeder,

TABLE II - HEAT AND MASS BALANCES
ANDCO-TORRAX (S.I.D.O.R.) SYSTEM

Refuse Quality	1600 kcal/kg Refuse (2880 Btu/lb.)		2500 kcal/kg Refuse (4500 Btu/lb.)	
	Heat (kcal/kg)	Mass (kg/kg)	Heat (kcal/kg)	Mass (kg/kg)
Input				
Refuse	1600	1.00	2500	1.00
Auxiliary Fuel	69	0.01	69	0.01
Combustion Air	3	3.49	4	6.26
Feedwater	244	1.74	408	2.90
Electrical Power	69	--	69	--
Total	1985	6.24	3050	10.17
Output				
Steam	1323	1.74	2206	2.90
Slag & Flyash	92	0.29	82	0.27
Exhaust Gases	296	4.21	464	7.00
Losses	369	--	393	--
Total	2080	6.24	3145	10.17
Steam Production	1.7 kg steam/kg refuse		2.9 kg steam/kg refuse	
System Efficiency*	62%		68%	

$$\text{*System Efficiency} = \left(\frac{\text{Steam-Feedwater}}{(\text{Refuse + Fuel + Electrical Power})} \right) \times 100\%$$

gasifier, secondary combustion chamber, regenerative towers, waste heat boiler, steam condensation unit, electrostatic precipitator, induced draft fan, chimney, water cooling plant, controls system, and all secondary equipment for compressed air, auxiliary fuel and waste oil subsystems. Figure 4 shows the SIDOR Andco-Torrax gasifier, Figure 5, the system.

The greater portion of the hot gases exiting from the SIDOR Andco-Torrax secondary combustion chamber is fed to a waste heat boiler, where the sensible heat content is used to generate steam. The boiler has been designed for a maximum gas throughput of 36,800 nm^3/h (21,660 scf/min.), a maximum gas inlet temperature of 1370°C (2500°F) and a nominal gas inlet temperature of 1250°C (2280°F). The boiler is a 3-pass, vertical, combination radiation/convection boiler, with the first of the three passes comprising the radiation section. The boiler has a super heater at the end of the radiation section and an economizer in the third pass. With a feedwater temperature of 140°C (284°F), the boiler is designed to produce a maximum of 60,000 lb/hr of steam at a pressure of 35 kg/cm^2 (498 psi) and a temperature of 385°C (725°F).

The steam produced by all three refuse conversion units is directed to a 7 MW turbine generator producing electricity at 10 kV. The turbine generator is designed for a throughput lower than the plant's corresponding refuse capacity of 600 metric tons per day, in that the amount of waste currently generated and available to the plant is lower than the total design capacity. A second turbine generator will be installed as refuse production increases.

Theoretically, 1 kg (2.2 lb) of 2000 kcal/kg (3600 Btu/lb) refuse can be converted to 1.22 Kwh of electricity. However, due to practical turbine efficiencies, line losses, and flue gas losses, the final conversion factor of refuse to electrical power is approximately 0.48 Kwh per kg of refuse. Assuming the availability of the system as 85%, the total amount of electricity produced per year from the Andco-Torrax system will be approximately 28.6×10^6 Kwh. Since the Andco-Torrax system electrical requirements are about 80 Kwh/ton of refuse, the total amount of electricity available to the local power distribution company is about 23.8×10^6 Kwh per year.

The waste gases from the boiler at 290°C (554°F) are combined with the exhaust gases from the regenerative towers and enter the electrostatic precipitator. The two field precipitator is designed to produce particulate emissions no greater than 100 mg/nm^3, corrected to 7% CO_2. The cleaned gases then pass through the induced draft fan to a triple flue stack, 80 meters in overall height.

ECONOMICS

Since 1973, Andco Incorporated and its licensees have been involved in several studies and have made qualitative and firm bids for refuse conversion plants. Four such bids have resulted in the sale of Andco-Torrax

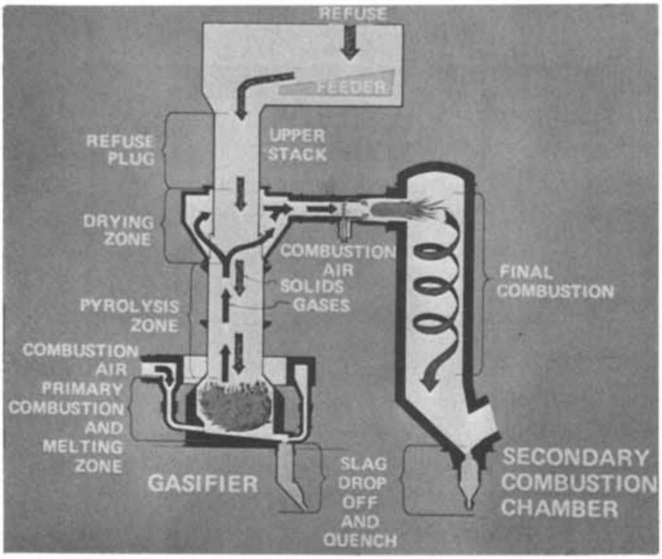


Figure 3. Andco–Torrax unit

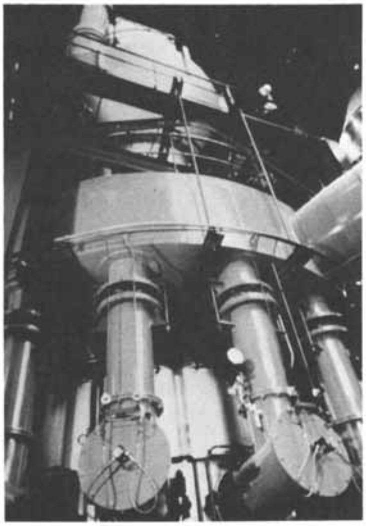


Figure 4. A 192-metric TPD SIDOR gasifier

systems to municipalities in Europe. An estimate of 1977 capital and operating costs are presented in Table III. It may be noted that construction costs for Andco-Torrax equipment alone range from a low of $22,000 per daily ton of plant capacity for a 1500 ton-per-day plant to a high of $30,000 per daily ton of capacity for a plant with a capacity of 250 tons per day. Net disposal costs per ton for plants of 250, 500, 1000 and 1500 tons respectively are $17.51, $11.90, $3.30 and $2.20. These net disposal costs after energy credits are currently competitive with other commercially available waste-to-energy technologies. The graph in Figure 6 utilizes the cost information presented in Table III with variable tipping (disposal) fees, and variable product credits. The family of curves thus generated depict Andco-Torrax plant economics for tipping fees ranging from zero dollars per ton of refuse to $20.00 per ton of refuse. Thus for any plant capacity and for a tipping fee between zero and $20.00 per ton, breakeven estimates for plant costs or energy credits may be obtained from the vertical axes. For example, a 750 ton-per-day plant having a $5.00 per ton tipping fee has breakeven plant costs of $20.00 per ton without energy credits. If plant costs are to be covered by product sales, then steam must be sold at $4.00 per thousand pounds or electricity must be sold at 48 mills/Kwh.

CONCLUSION

The Andco-Torrax System has been proven through successful development and operation of a 75-ton-per-day demonstration plant in the United States over a four-year period. The first two commercial Andco-Torrax Systems have recently commenced operation in Luxembourg and Grasse, France. The third unit, located in Frankfurt, is scheduled for start-up in the second quarter of 1978. The fourth plant, in Creteil, France, will be completed in early 1979.

The Andco-Torrax high-temperature slagging pyrolysis system accomplishes the three objectives required of modern refuse conversion units:

1. Maximum Volume Reduction - Volume reduction in the Andco-Torrax System is 95-97%. Future tests at the Frankfurt plant will be made to determine what further volume reduction can be achieved with conventional incinerator residue by adding varying amounts of incinerator residue to the Andco-Torrax System.

2. Inert Residue - The fritted, glassy aggregate residue is completely sterile and inert, consisting entirely of oxidized material as a result of the high-temperature process conditions.

3. Resource Recovery - The production rate of energy in the form of steam and/or electrical power is high as a result of all of the combustible material in the refuse being totally consumed. The economics of an Andco-Torrax energy recovery plant are similar to those of other commercially available, competitive waste-to-energy technologies.

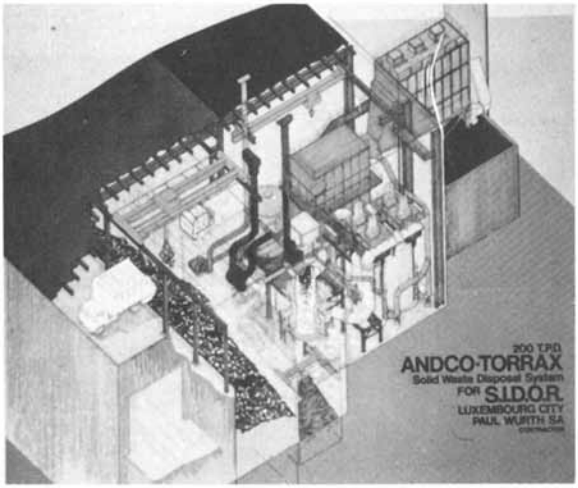
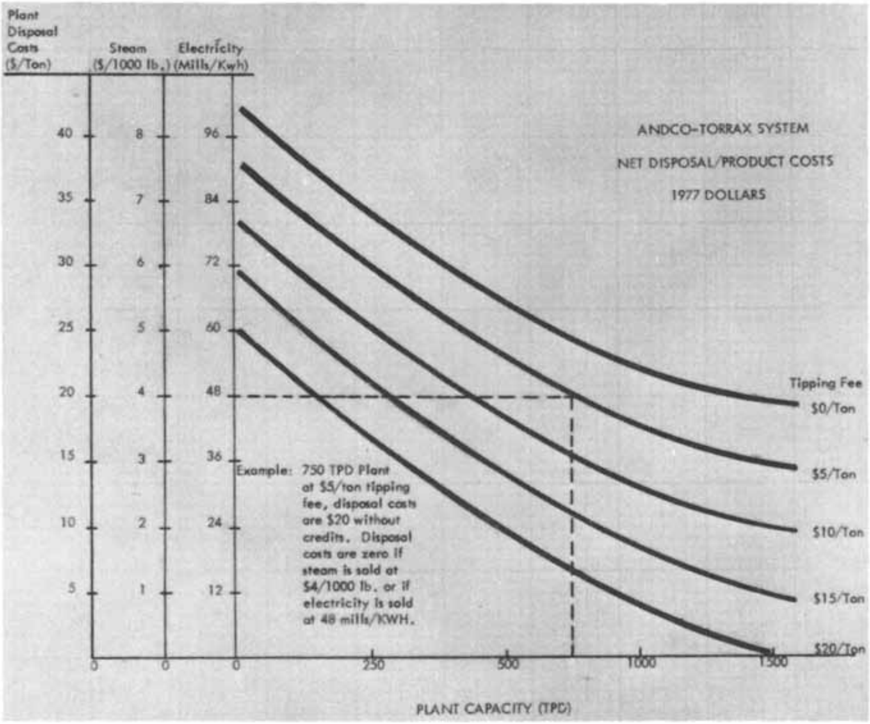


Figure 5. SIDOR plant—Luxembourg: Andco Torrax equipment train



Example: 750 TPD Plant at $5/ton tipping fee, disposal costs are $20 without credits. Disposal costs are zero if steam is sold at $4/1000 lb. or if electricity is sold at 48 mills/KWH.

Figure 6. Andco–Torrax system net disposal/product costs

TABLE III - ANDCO-TORRAX SYSTEM ECONOMICS (1977)

	250 TPD	500 TPD	1000 TPD	1500 TPD
No. of Andco-Torrax Units	1	2	3	5
Annual Refuse Throughput (TPY)	82125	164000	329000	492750
CAPITAL COSTS ($000 omitted)				
Andco-Torrax Equipment	7500	14900	20000	33000
Buildings & Utilities	5800	8080	12860	14420
Interest During Construction	1130	1953	2793	4030
Start-up Expense	740	850	1080	1730
Working Capital	466	672	980	1375
Total Capital Costs	15636	26455	37713	54555
Amortization Cost/Ton (8-1/2% interest, 20 year plant life)	19.83	16.80	11.94	11.53
OPERATING COSTS ($/Ton)				
Labor & Administration (No. of Personnel)	6.94(19)	5.48(30)	3.56(39)	2.88(47)
Maintenance, Power & Utilities	9.64	8.52	6.70	6.49
Total Operating Costs/Ton	16.58	14.00	10.26	9.37
Total Plant Costs/Ton	36.41	30.80	22.20	20.90
CREDITS				
Steam	18.90	18.90	18.90	18.90
NET DISPOSAL Cost/Ton	17.51	11.90	3.30	2.20

ASSUMPTIONS:

1. A 1000 TPD plant will have 3-330 TPD units. A 1500 TPD plant will have 5-300 TPD units.
2. Plant availability is 90%.
3. No escalation of capital cost to completion of project.
4. Interest during construction is estimated at 8$\frac{1}{2}$% of construction costs.
5. Start-up expense is estimated at 3%-6% of construction costs.
6. Working capital is estimated at 3% of construction costs.
7. Average total wages and benefits per employee per annum is $28,000.
8. Annual maintenace costs are 4% of capital (equipment and buildings).
9. Plant electrical consumption is 80 KWH/ton of refuse. Power costs 20 Mills/KWH.
10. Auxiliary fuel usage is 2.5 nm^3 (88.3 ft^3) natural gas/ton. Natural gas costs $3.00/1000 ft^3.
11. Refuse is assumed to have an LHV of 2275 kcal/kg (4100 Btu/lb.). Steam produced will be 2.7 kg/kg refuse. Steam credits will be $3.50/1000 lb.

ABSTRACT

In our society today, the disposal of municipal solid waste poses ever increasing social and technical problems. During recent years, widespread attention has been focused on both real and potential scarcities of conventional fossil fuels. The search for economically attractive solutions to these two problems has given rise to the development of different systems to process municipal solid waste by thermal and chemical techniques to provide alternate energy sources. This paper describes one such system called the Andco-Torrax System which is used to convert municipal solid waste into useful energy and a granulated slag byproduct by the pyrolysis and primary combustion of organic materials and by the melting of non-combustible materials at temperatures up to 1650°C (3000°F).

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Co-Disposal of Sludge and Refuse in a Purox Converter

C. T. MOSES and K. W. YOUNG

Union Carbide Corporation, Linde Division, Tonawanda, NY 14150

G. STERN and J. B. FARRELL

U.S. E.P.A., MERL, Cincinnati, OH 45268

The development of the PUROX System for the conversion of municipal refuse into a useful fuel gas, a treatable wastewater, and an inert slag stream has been described in several publications (1,2,3). The key element in this system is the converter which is basically an oxygen-fed shaft furnace in which solid waste is dried, pyrolyzed, combusted, and the inorganic residues slagged. The operation of the converter is supported by several subsystems: a feed material storage and preparation system, a gas cleaning system, a wastewater treatment system, and an oxygen generating system. Figure 1 presents a mass balance indicating the general process flow and typical material rates for operation with municipal refuse. One mass unit of as-received refuse undergoes shredding and magnetic separation of ferrous material. The magnetically separated, shredded material is combined with 0.22 mass units of oxygen and is converted into 0.7 mass units of product gas saturated with water at ambient condition, 0.28 mass units of wastewater, and 0.22 mass units of slag. In the S. Charleston, WV demonstration plant a small amount of natural gas is injected with the oxygen into the slagging zone to provide preheating of the entering oxygen prior to contacting the molten slag pool. In a commercial installation this heat input would be provided by recycle of a portion of the product gas stream.

Figure 2 presents an energy balance for the generalized process described above. Over 70% of the entering energy is converted into product gas. The product gas is a medium-Btu fuel with its major constituents being CO, H_2, CO_2 and CH_4 having a heating value of 350 Btu/S ft^3.

A natural extension of the solid waste processing capability of the PUROX System was the use of dewatered sewage sludge cake as a feed material in combination with municipal refuse. This combined processing scheme has the potential advantage of

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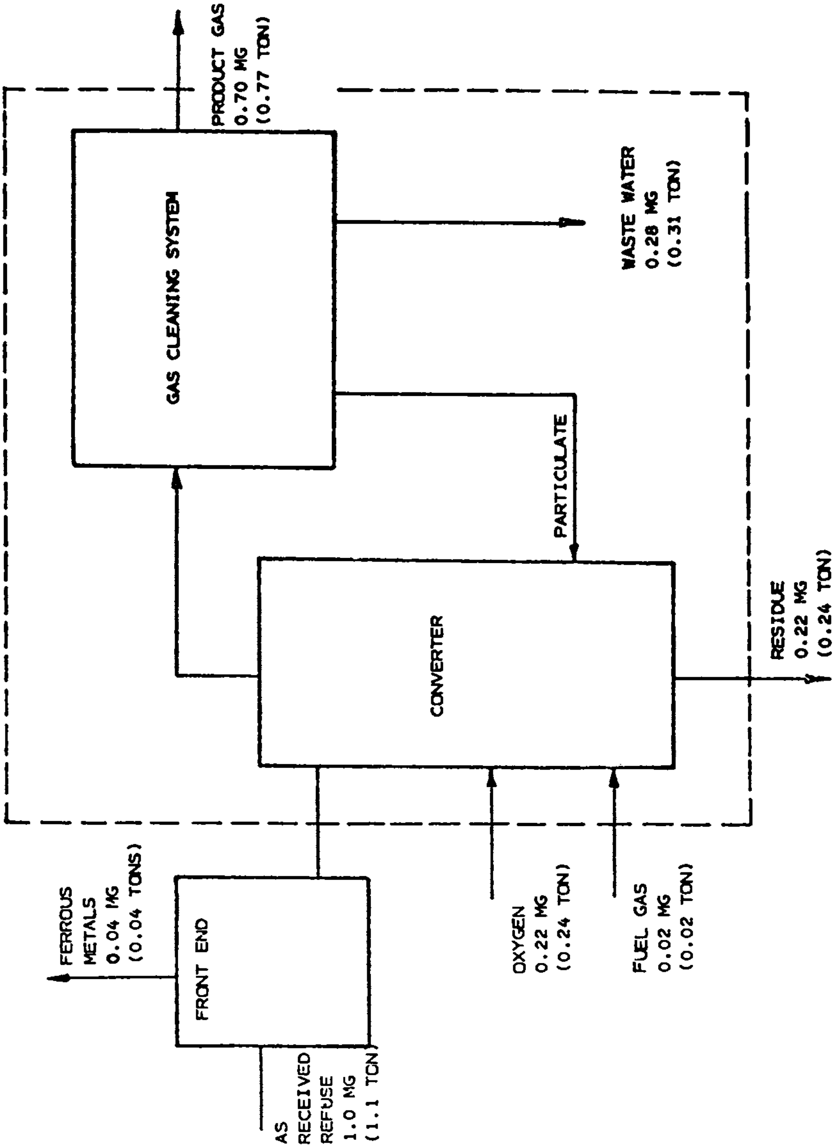
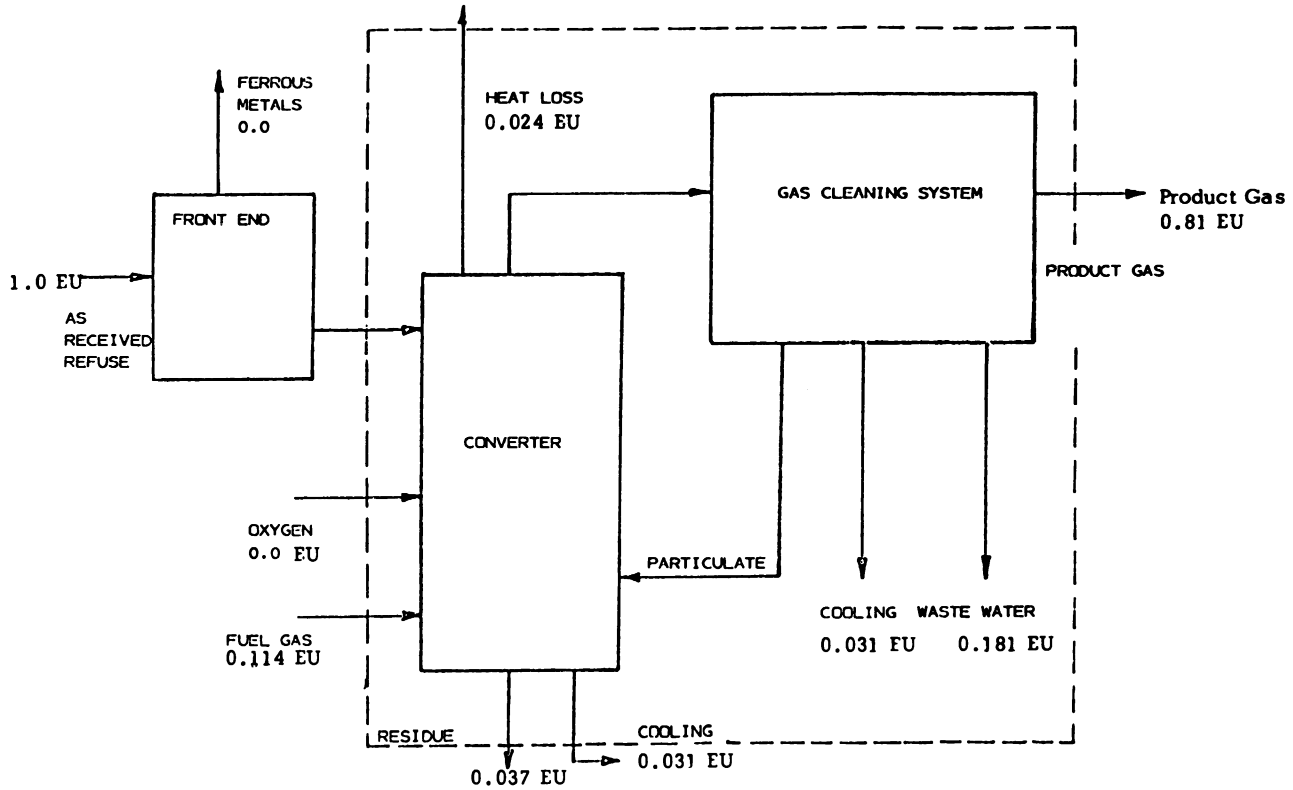


Figure 1. Typical PUROX System mass balance



Basis: 1 EU of Feed (EU = energy unit)

Figure 2. Typical PUROX System energy balance

disposing of sludge and refuse with a net production rather than the net consumption of energy in conventional incineration techniques. Also, the positive environmental features of slagging the inorganic residues in a non-leaching, glassy matrix suggests a major environmental improvement in thermal processing techniques for sludge disposal.

Following some preliminary analysis and pilot scale testing of the concept of sludge/refuse codisposal in the PUROX System, Union Carbide Corporation proposed through the Sanitary Board of the City of South Charleston, WV, to the U.S. Environmental Protection Agency that a large scale testing program be conducted utilizing the PUROX System demonstration plant in South Charleston. The EPA approved the proposed program and construction work began in the fall of 1976. Construction was completed in February of 1977 with the test program starting in mid April. The testing was conducted for a two month period. A discussion of the test program and its results is presented below.

Figure 3 presents an isometric view of the South Charleston PUROX System facility arranged for codisposal of sludge and refuse. Major pieces of processing equipment are described in the legend on the figure. The front end system consisting of the refuse storage building, the vertical-shaft, hammermill-shredder, and the drum magnetic separator was expanded to provide a dewatered-sludge-cake processing capability by the following additions. Dewatered sludge cake from the wastewater treatment plant was delivered in dump trucks and stored in a sludge inventory area for subsequent use. Sludge cake was periodically loaded into a live-bottom metering hopper from which it was steadily metered into the shredded refuse stream going to the converter.

The mixture of sludge cake and refuse was fed into the converter where it underwent the drying, pyrolysis, combustion, and slagging steps to convert it into product gas, slag, and wastewater. The inorganic residues (slag) left the converter as a continuous molten stream which was quenched in a water tank and discharged into a storage dumpster. The products of the drying, pyrolysis and combustion reactions passed overhead through the gas cleaning system which consisted of a scrubber tower, electrostatic precipitator, and condenser. Wastewater was collected from the scrubbing and condensing operations. The product gas was continuously flared in a combustor. Figures 4 and 5 provide additional views of the actual sludge handling operation showing the sludge filter cake in the storage area being placed in the metering hopper for discharge onto the belt conveyor.

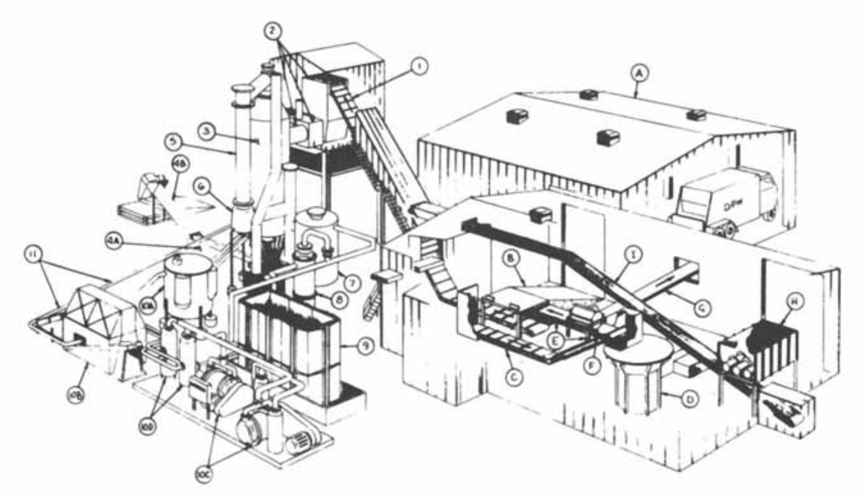


Figure 3. PUROX System facility for sludge/refuse codisposal in South Charleston, West Virginia.

PUROX System: (1) PUROX System leveling conveyor; (2) PUROX System pelletizing feeder; (3) PUROX System converter; (4) PUROX System slag quench system: (a) PUROX System slag quench tank, (b) PUROX System slag conveyor; (5) PUROX System scrubber; (6) PUROX System gas liquid separator; (7) PUROX System electrostatic precipitator; (8) PUROX System condenser; (9) PUROX System combustor; (10) PUROX System solid liquid separation system: (a) PUROX System filtrate tank, (b) PUROX System vacuum filter, (c) PUROX System vacuum pumps, (d) PUROX System vacuum receivers; (11) PUROX System char conveyors. Front End (not PUROX System): (A) refuse storage building; (B) refuse weigh bridge and load ramp; (C) refuse feeder pit and shredder feed conveyor; (D) shredder; (E) shredder discharge conveyor; (F) magnetic separator; (G) magnetic material conveyor; (H) sludge metering hopper with augers; (I) sludge feed conveyors.

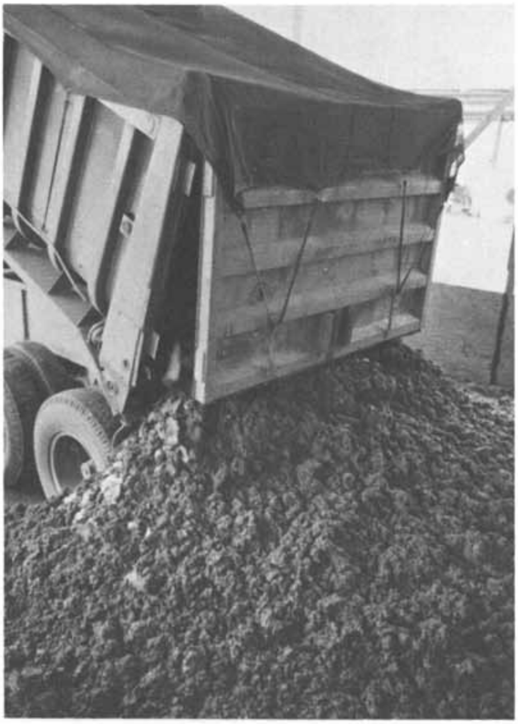


Figure 4. Sludge cake at dumping station

Figure 5. Front loader scooping sludge cake

Codisposal Test Program

Two types of dewatered sludge cake were selected for testing in this program. Raw primary sludge cake with a nominal moisture content of 25% solids, a volatiles content of ~ 45% and a heating value of 5000 Btu per lb of dry solids, was available in sufficient quantity from the wastewater treatment plant at the City of Huntington, WV. Raw mixed sludge cake was not readily available in the South Charleston environs. It was necessary to go to the Alleghany County Sanitary Authority Waste Treatment Plant in Pittsburgh, PA to obtain a sufficient supply of dewatered, raw mixed sludge. This sludge was a mixture of 60% raw primary and 40% secondary which had been dewatered to ~20% solids. The mixed sludge had a volatiles content of ~60%, and a heating value of 6500 Btu per lb of dry solids.

These two types of sludge were processed with municipal refuse from the S. Charleston area. The refuse was typical mixed municipal refuse with a moisture content of 25-35%, a dry inert content of ~25%, and a heating value of 5000 Btu/lb. As noted above, the refuse was shredded and magnetic separation of the ferrous components was carried out prior to combination with the sludge.

The test program consisted of a series of seven tests during which comprehensive sampling and analysis of the process streams entering and exiting the system were conducted. Test A was a baseline operating condition using refuse-only which was conducted to provide a reference point for comparison of results from the codisposal tests. The converter operating rate averaged 81 tons per day for this test.

Test B was a combination of primary sludge and refuse in a sludge-dry-solids (SDS) to as-received refuse (ARR) ratio of 0.016. The plant was operated at an average rate of 61 tons per day (TPD) of refuse and 3.7 TPD of 26% solids primary sludge for a total operating rate of 64.7 TPD.

Test C was conducted using primary sludge at an SDS/ARR ratio of 0.021. The plant operated at an average rate of 85 TPD of as-received refuse and 5.6 TPD of 32.4% solids primary sludge for a total operating rate of 90.6 TPD.

Test D was conducted using primary sludge at an SDS/ARR ratio of 0.027. The plant operated at an average rate of 83 TPD of refuse and 8.1 TPD of 27.5% solids sludge for a total operating rate of 91.1 TPD.

Test E was conducted using primary sludge at an SDS/ARR ratio of 0.052. The plant operated at an average rate of 95 TPD of refuse and 22.2 TPD of 22.3% solids sludge for a total operating rate of 117.2 TPD.

Test F was a combination of mixed secondary sludge and refuse with an SDS/ARR ratio of 0.027. The plant operated at an average rate of 86 TPD of refuse and 11.6 TPD of 20.3% solids mixed sludge for a total operating rate of 97.6 TPD.

Test G was conducted using raw mixed sludge at an SDS/ARR ratio of 0.031. The plant operated at an average rate of 67.3 TPD of refuse and 10 TPD of 21% solids mixed sludge for a total operating rate of 77.3 TPD.

Table I presents a summary of the test conditions described above. Summary tables of the mass flows, energy flows, product gas and slag analyses for each of these tests are presented in Tables II - V. As can be noted from the tables, there was no major departure from the baseline refuse-only results in the quantities reported. The slight variations that do appear are in part the result of changes in refuse composition that occur in a ramdomly varying fashion. Mechanical problems prevented operation of the particulate recycle system throughout the test program. This resulted in an inability to recycle particulate from the electrostatic precipitator and scrubber to the converter and the loss of representative wastewater samples due to the use of substantial amounts of makeup water in the scrubbing system. During subsequent plant operation with refuse-only, stable operation of the particulate recycle system was achieved. A model was developed for projecting the codisposal data to the performance obtained during particulate recycle. The projected results for refuse-only (Test A) were in good agreement with the experimental data from the particulate recycle period which gave the projection model results for the codisposal period a measure of credibility.

The mass flows in Table II are presented on a per-unit-of-feed basis to aid in illustrating the general similarity of the codisposal results with those obtained for refuse only. The energy flows in Table III are based on the total quantity of energy leaving the system since it was not possible to obtain reliable measurements of the energy in the input refuse stream. The mass and energy flows were also adjusted to the recycle case where particulate is returned to the converter. In examining these results, test G appears to be the only period that differs significantly from the refuse-only and other codisposal results for key parameters such as oxygen usage and conversion to product gas. These differences are the result of mechanical problems encountered using the existing feeding equipment to blend the sludge and refuse. Improper blending produced a variable porosity condition in the converter bed which permitted a high heat loss condition to develop resulting in elevated oxygen consumption and heat loss factor while

TABLE I. SLUDGE CODISPOSAL TEST SERIES

Test #	Source/Type	Sludge % Solids	Feed Rate Mg/d (ton/d) Refuse	Sludge (Wet)	Total	Sludge/Refuse Wet	SDS/ARR
A	-	-	73.5 (81.0)	-	73.5 (81.)	-	-
B	H/RP	26.23	55.3 (61.0)	3.4 (3.7)	58.7 (64.7)	0.061	0.016
C	H/RP	32.38	77.1 (85.0)	5.1 (5.6)	82.2 (90.6)	0.066	0.021
D	H/RP	27.47	75.3 (83.0)	7.4 (8.1)	82.7 (91.1)	0.098	0.027
E	H/RP	22.32	86.2 (95.0)	20.1 (22.2)	106.3 (117.2)	0.234	0.052
F	P/MS	20.25	78.0 (86.0)	10.5 (11.6)	88.5 (97.6)	0.135	0.027
G	P/MS	21.01	61.0 (67.3)	9.1 (10.0)	70.1 (77.3)	0.149	0.031

H/RP - Raw primary sludge from Huntington, West Virginia

P/MS - Secondary sludge from Pittsburgh, Pennsylvania mixed with primary sludge in approximate ratio of 60% raw primary, 40% secondary

TABLE II. PROJECTED MASS FLOW (PER UNIT OF TOTAL FEED)

	A	B	C	D	E	F	G
Refuse-Sludge Feed	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Fuel Gas	0.023	0.029	0.016	0.014	0.016	0.020	0.018
Oxygen	0.236	0.241	0.205	0.223	0.223	0.217	0.277
Total In	1.259	1.270	1.221	1.237	1.239	1.237	1.295
Metal	0.084	0.043	0.051	0.050	0.039	0.043	0.036
Product Gas	0.605	0.645	0.571	0.614	0.586	0.576	0.616
Slag	0.242	0.176	0.258	0.248	0.191	0.213	0.193
Wastewater	0.328	0.406	0.341	0.325	0.423	0.405	0.450
Total Out	1.259	1.270	1.221	1.237	1.237	1.237	1.295

TABLE III. PROJECTED ENERGY FLOW* BASED ON TOTAL ENERGY EXITING THE PROCESS

(Referenced to 60 deg. F)

Exiting Streams**	A	B	C	D	E	F	G
Product Gas	77.0%	74.8	75.1	76.8	74.8	74.3	71.2
Slag	4.6%	3.2	5.2	4.8	3.7	4.2	3.9
Wastewater	4.7%	5.9	5.2	5.0	6.5	6.0	6.7
Heat Losses	13.7%	16.1	14.5	13.4	15.0	15.4	18.2
	100%	100%	100%	100%	100%	100%	100%

* Since it was not possible to obtain representative measures of the input energy streams, the output energy flows were measured.

** Exiting streams are given as a % of total energy leaving.

TABLE IV. SUMMARY OF PRODUCT GAS ANALYSES

	Mole Percents						
Component	Test A Vol. %	Test B Vol. %	Test C Vol. %	Test D Vol. %	Test E Vol. %	Test F Vol. %	Test G Vol. %
H_2	28.14	29.64	31.25	29.62	26.81	25.30	30.32
CO	37.28	33.74	36.24	36.80	34.19	36.72	31.56
CO_2	23.07	25.58	23.52	23.85	26.85	25.01	28.86
CH_4	6.21	5.89	4.46	5.02	6.12	6.48	4.82
C_2H_2	0.46	0.52	0.25	0.41	0.72	0.60	0.48
C_2H_4	1.90	1.91	1.32	1.68	2.19	2.26	0.68
C_2H_6	0.44	0.34	0.33	0.32	0.40	0.46	0.23
C_3H_6	0.32	0.27	0.27	0.23	0.28	0.39	0.20
C_3H_8	0.23	0.12	0.25	0.18	0.19	0.21	0.02
C_4*	0.40	0.28	0.47	0.34	0.40	0.44	0.23
C_5+	0.29	0.22	0.35	0.27	0.36	0.31	0.17
C_6H_6	0.24	0.26	0.17	0.25	0.34	0.26	0.22
HHC‡	0.26	0.17	0.32	0.29	0.37	0.23	0.12
H_2S	0.02	0.02	0.00	0.04	0.03	0.02	0.01
CH_3OH	0.11	0.04	0.14	0.08	0.09	0.11	0.12
O_2	0.03	0.04	0.05	0.03	0.04	0.26	0.24
Ar	0.59	0.61	0.62	0.59	0.63	0.95	0.74

* C_4 is a composite of butanes, butenes, butadiene, vinyl acetylene and diacetylene

+ C_5 is a composite of pentanes, pentenes, isoprene, and cyclopenetadiene

‡ HHC are higher hydrocarbons C6 and above excluding benzene

TABLE V. SUMMARY OF SLAG ANALYSES

Major Components	Test A	Test B	Test C	Test D	Test E	Test F	Test G
C (wt. %)	1.8	0.8	1.0	0.8	0.4	0.9	0.2
SiO_2	61.5	56.1	60.0	59.8	56.8	66.5	64.0
Al_2O_3	11.6	9.7	10.0	11.9	9.3	10.5	11.0
C_aO	9.8	8.5	12.0	10.5	11.3	9.9	12.1
FeO	8.0	14.6	8.1	8.1	8.3	6.3	6.7
MgO	0.9	1.9	1.9	0.9	1.0	1.6	1.2
P_2O_5	0.6	0.4	0.9	0.6	0.8	0.8	0.6
BaO	0.1	0.1	0.2	0.2	0.2	0.1	0.1
MnO	0.2	0.3	0.2	0.1	0.2	0.2	0.3
Cr_2O_3	0.5	0.5	0.05	0.1	0.5	0.5	0.05
TiO_2	0.7	0.6	0.51	0.52	0.6	0.6	0.56
Ni_O	0.05	0.05	0.3	0.2	0.1	0.05	0.05
	95.75	93.55	95.16	93.72	90.4	97.95	96.86

Trace Components

	Test A	Test B	Test C	Test D	Test E	Test F	Test G
Cd (ppm)	5	5	10	10	5	5	6
Cr	3420	3420	891	645	3420	3420	363
Cu	4000	3400	4300	4000	3300	2860	3800
Fe	61970	113100	55000	53000	64300	48800	53000
Hg	0.03	0.01	0.01	0.02	0.15	0.05	0.01
Mn	1600	1300	1200	1700	1400	1140	1400
Ni	305	959	2700	1300	1500	150	145
Pb	94	75	119	119	243	120	128
Zn	309	404	455	482	1100	350	456

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

depressing overall conversion to useful fuel gas. This condition was corrected as noted in the results for Test F in which similar operating rates with similar sludge were used. Careful control of the sludge-refuse blending conditions in Test F resulted in normal performance.

The environmental effects of the codisposal process were the key measurements to be made during these tests. The process product streams - product gas, wastewater, and slag were carefully monitored. In addition to the overall composition of the product gas reported earlier, sampling and analysis of the particulate remaining in the product gas after cleaning were conducted. Isokinetic samples of the product gas were collected, and the particulate obtained was thoroughly analyzed for key trace metal contaminants. Since the exiting gas leaves the system at $\sim$100 deg. F, it seemed likely that any significant amounts of trace metal vapor would be present in a condensed state on the residual particulate. Table VI presents a summary of the product gas particulate levels It should be noted that these levels are for gas prior to combustion. Since the particulate regulation is based on combustion products at standard conditions with 12% CO_2, these numbers must be converted. Table VII presents the calculated particulate levels in product gas combustion products converted to 12% CO_2. All the levels reported are well below the regulated value of 0.08 grains per standard cubic foot at 12% CO_2. Table VIII reports the average trace metal levels found in the particulate material during the codisposal test period. Levels are reported on a combustion product basis for both the refuse-only and codisposal operation. The apparent improvement in trace metal levels of the codisposal period relative to the refuse-only period is misleading. At the time the refuse-only data were collected, the electrostatic precipitator in the gas cleaning system was operating unreliably which resulted in non-optimum gas cleaning performance. In general the trace metal levels during refuse-only operation should be very comparable to those observed during codisposal tests. A comparison of these combustion product levels is also made with time-weighted-average (TWA) threshold-level-values (TLV) proposed by the American Conference of Industrial Governmental Hygienists. In the absence of more complete emission regulations, these work-place regulations, although not specifically applicable, give an indication of regulated levels for these materials.

TABLE VI. PRODUCT GAS PARTICULATE LEVELS

Test Period	Type Feed	Particulate Loading	
		g/Nm^3	gr/Sft^3
A	Refuse	0.0176	0.007
B	Refuse + Primary Sludge	0.0223	0.009
C	Refuse + Primary Sludge	No samples taken	
D	Refuse + Primary Sludge	0.372	0.015
E	Refuse + Primary Sludge	0.0449	0.018
F	Refuse + Mixed Sludge	0.0201	0.008
G	Refuse + Mixed Sludge	0.0188	0.008

TABLE VII. CALCULATED PARTICULATE LEVELS IN PUROX SYSTEM GAS COMBUSTION PRODUCTS AT 12 PERCENT CO_2*

Test Period	Particulate Level	
	g/Nm^3	gr/Sft^3
A (Refuse)	0.0031	0.0013
B (Primary)	0.0034	0.0014
C (Primary)	-	-
D (Primary)	0.0056	0.0023
E (Primary)	0.0063	0.0026
F (Mixed)	0.0029	0.0012
G (Mixed)	0.0031	0.0013

TABLE VIII. TRACE METAL CONCENTRATIONS IN PRODUCT GAS COMBUSTION PRODUCTS

Combustion Product Concentration (mg/m^3, dry @ 12% CO_2)*

Metal	Refuse +	Refuse + Sludge (∿5% sludge (ds)/refuse)‡	TLV-TWA # (mg/m^3)
Cd	0.0022 mg/m^3	0.0001 mg/m^3	0.05 mg/m^3
Cr	0.0090	0.0007	0.5
Cu	0.079	0.0097	0.2
Fe	1.62	0.03	5
Hg	0.0084	below detection limit	0.05
Mn	0.012	0.0004	5
Ni	0.058	0.002	1
Pb	0.038	0.002	0.015
Zn	0.047	0.009	5

*Computed from metal content in isokinetically collected particulate in the product gas.

+ Refuse-only with particulate recycle and non-optimum gas cleaning.

‡ Based on average of measurements using primary and mixed sludges with more optimum gas cleaning performance.

#(4) ACGIH, National Safety News, pp. 83-93, September 1977.

TABLE IX. PUROX SYSTEM WASTEWATER TREATABILITY DATA

Wastewater Analysis		Reactor Operating Conditions	
$TBOD_5$	1350 ppm	$F/M \frac{BOD_5/day}{MLVSS}$	0.68
$SBOD_5$	1321 ppm	t_d, days	1.18
TCOD	3078 ppm	MLVSS, ppm	1684
SCOD	3012 ppm	So ($TBOD_5$), ppm	1350
COD/BOD_5	2.28	Se ($SBOD_5$), ppm	59

The wastewater production is obviously strongly affected by the moisture content of the feed material. The strength of the wastewater is related to the quantity produced. In general it has been found that the wastewater can be readily treated to a level sufficient for discharge into a municipal sewer system. A 95% reduction in biological oxygen demand can be achieved by diluting at a ratio of 20 to 1 with municipal sewage to provide nutrients and control for the biological process. Table IX summarizes the wastewater treatability data. Table X presents a summary of trace metal concentrations measured in the wastewater during operation of the particulate recycle system with refuse-only. Pretreatment requirements that are also given in the table are an estimate of the current status of regulations that are now being formulated. As can be seen from the table, the raw wastewater exceeds some of the projected standards. However, a relatively easy solution to this problem can be obtained through the use of lime (CaO) precipitation. The small amount of precipitate could then be disposed of in an environmentally acceptable fashion.

TABLE X. WASTEWATER METAL CONCENTRATION FROM REFUSE ONLY OPERATION WITH PARTICULATE RECYCLE

Metal	Raw Wastewater	Wastewater and Precipitation*	Pretreatment Requirement
Cd	0.22 ppm	0.07 ppm	0.1 ppm
Cr	<0.07	< 0.04	1.0
Cu	0.12	<0.04	0.5
Fe	39.9	0.04	20.0
Pb	43.0	<0.05	0.5
Mn	19.9	1.0	1.0
Ni	0.11	< 0.04	5.0
Zn	83.7	<0.03	5.0
Hg	< 0.02	<0.003	0.02

*These values are essentially the detection limits for these metals where a < sign is used. Based on solubility calculation for hydroxides at pH 10.

The trace metal content of the slag was reported in the earlier summary tables. The slag analysis do not reflect significant differences for the major constituents. The trace components, particularly zinc and lead, show an increase with increasing proportion of sludge. To evaluate the leaching potential of the

TABLE XI. SLAG LEACHATE ANALYSIS

Analysis	Test A	Test B	Test D	Test E	Test F	Test G
pH	7.7	7.8	-	7.9	7.7	7.9
F	0.07 ppm	0.06 ppm	- ppm	0.08 ppm	- ppm	- ppm
Cl	18.0	4.0	4.0	4.0	3.0	4.0
P	0.2	0.7	2.0	0.6	1.0	1.0
As	< 0.1	< 0.1	< 0.1	< 0.1	< 0.1	< 0.1
Ba	< 0.03	0.09	< 1.0	0.10	< 1.0	< 1.0
Be	< 0.03	< 0.03	< 1.0	< 0.03	< 1.0	< 1.0
Cd	0.02	0.02	< 1.0	0.02	< 1.0	< 1.0
Cr	< 0.04	< 0.04	< 0.1	< 0.04	< 0.1	< 0.1
Cu	< 0.04	< 0.04	< 0.1	< 0.04	< 0.1	< 0.1
Hg	< 0.003	< 0.003	< 0.05	< 0.003	< 0.05	< 0.05
Fe	0.18	0.04	0.1	0.05	0.4	0.5
Pb	< 0.05	< 0.05	< 0.1	< 0.05	< 0.1	< 0.1
Mn	< 0.02	0.05	< 0.1	0.04	< 0.1	< 0.1
Ni	< 0.04	0.05	< 0.1	< 0.04	< 0.1	< 0.1
Zn	< 0.03	< 0.03	< 1.0	< 0.03	< 1.0	< 1.0
Ag	0.1	0.1	< 0.1	0.1	< 0.1	< 0.1

slag, a simple test procedure was devised in the absence of a standard method for such a test. A 375 gram representative slag sample was placed in a beaker with 1500 ml of distilled water. The beaker was gently agitated for at least 48 hours. The water was then analyzed for trace metals. Table XI summarizes the results of these analyses. An essentially negligible amount of material has leached from the slag.

Table XII gives an overall summary of the disposition of trace metals in the three process effluent streams during refuse-only operation with particulate recycle. Hg, Cd, Zn and Pb show more volatility than the bulk of the metals; however, over 75% of these four end up in the slag. Except for mercury, over 99% of these metals are captured in the slag or wastewater. As the particulate loadings presented earlier indicate, the amounts that escape in the product gas are exceptionally low. Even Hg, which has the highest percentage loss, is in absolute quantity, an order of magnitude below the regulated level cited in Reference (4).

TABLE XII. TRACE METAL DISPOSITION IN PROCESS EFFLUENT STREAMS

Metal	Percent of Total Metal Emitted in Product Gas Before Combustion	Percent of Total Metal Emitted in Slag	Percent of Total Metal Emitted In Wastewater*
Cd	0.50 percent	96.4 percent	3.10 percent
Cr	0.05	99.92	0.03
Cu	0.04	99.95	0.01
Fe	0.05	99.85	0.10
Pb	0.30	77.60	22.10
Mn	-	99.20	0.80
Ni	0.60	99.30	0.10
Zn	0.80	88.70	10.50
Hg	15.0	82.50	2.50

*Trace metals would be precipitated prior to wastewater treatment.

The mechanism of capture of volatile metals in the slag as indicated in the data here is believed to be related to the formation of relatively stable inorganic salts which become phsico-chemically bonded within the silicate matrix of the slag. The high removal of mercury in the slag is especially impressive because this represents permanent removal from the environment.

Projected Economics of the Codisposal Process

The economics of codisposal of sewage sludge and refuse in the PUROX System are strongly dependent upon the allowable cost of refuse disposal. If a relatively high tipping fee ($15/ton) can be charged for refuse disposal, then an attractive net sludge disposal cost can be developed. However, if low cost ($5/ton) refuse disposal alternatives are available, then the net sludge disposal cost will appear less attractive since a relatively small quantity of sludge must carry a larger portion of the total system cost.

Processing alternatives can also have a substantial effect upon sludge disposal costs. The degree of dewatering strongly affects the size and cost of equipment required to handle a given amount of sludge dry solids. Since the process is a capital-intensive operation, improved dewatering including drying prior to mixing the sludge with refuse can generally result in reduced total capital cost. Four cases have been developed to examine these effects. Case A is based on thickened sludge at 3% solids dewatered to 20 percent solids and delivery of the 20 percent sludge cake to the PUROX facility. Case B is based on 3% solids sludge dewatered to 25 percent solids and delivery of the sludge cake to the PUROX facility. For these cases, the PUROX facility is operated in a similar manner to the test program in which separately dewatered sludge is fed to the converter along with municipal refuse.

Economics and practical considerations favor removing as much moisture as is reasonable in a combination of dewatering and drying steps before introduction of the sludge with refuse into the PUROX converter. This minimizes use of the PUROX converter and gas cleaning capacity for processing moisture. It maximizes the use of the PUROX System capacity for the critical task of producing energy and disposing of the sludge solids. Cases C and D are included in the projections of sludge disposal costs to reflect improvements in the codisposal process compared to that which was practiced at South Charleston. Case C is based on delivery of 3 percent sludge solids to the PUROX facility. At the facility, the 3 percent sludge solids would be combined with the PUROX System scrubber slurry and jointly dewatered to 35 percent solids with a belt filter press. The scrubber slurry contains particulate, produced from the pyrolysis of refuse and sludge, which enhances the dewatering characteristics of sludge. Bench scale evaluations of the dewaterability of the particulate and sludge mixture have shown significantly reduced polymer requirements from that of separate sludge dewatering and substantially higher cake solids following the filtering process.

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

Case D is based on codewatering of 3% solids sludge and particulate to 35 percent combined solids as in Case C and drying the 35 percent cake solids in a separate dryer to approximately 70 percent solids. In Cases C and D the particulate/sludge mixture, after dewatering (and drying in Case D), is fed into the converter together with municipal refuse. To properly compare these cases, all costs associated with dewatering from 3 percent solids and ultimately disposing of the sludge have been included. Dewatering costs associated with separate dewatering were taken from published literature as $55 to $66 per dry Mg ($50 to $60 per dry ton) for dewatering from 3 percent to 25 percent solids (5).

Figures 6 and 7 illustrate the effect of refuse tipping fee and extent of dewatering upon sludge disposal costs at the national per capita sludge-to-refuse ratio of 0.05. Sludge disposal costs range from several hundred dollars per ton for small volume plants to a potential net credit for very large plants with drying and a $15/ton refuse tip fee. The situation specific nature of the system economics is readily apparent from these curves.

Many municipalities may require higher sludge-to-refuse ratios than the national per capita due to the diversified nature of the refuse supply and the centralized wastewater treatment plant. To examine this case, economics for a Case D type system with codewatering and drying of sludge at a sludge to refuse ratio of 0.3 have been prepared in Table XIII. The plant would have a nameplate capacity of ~2000 tons per day of sludge/refuse mixture at the specified ratio. Capital equipment costs, amortization of capital, annual operating costs and revenues are presented to develop a net sludge disposal cost for processing 400 tons per day of sludge dry solids.

The resultant sludge disposal cost of $40/ton of dry sludge is only meant to be illustrative. It is based on a budgetary economic projection. Site and project-specific costs such as: financing costs, capital escalation, interest during construction, land cost, insurance cost, consulting fees and miscellaneous legal fees are not included.

In conclusion, individual site requirements will strongly affect project economics. At very low dry sludge-to-refuse ratios, refuse disposal economics (the available refuse tipping fee) will control rather than sludge related costs. If a refuse-only facility can be economically supported then, sludge at the national per capita ratio of sludge-to-refuse can be processed in the facility with very attractive disposal costs. If refuse only operation does not appear viable, then the relatively small amount of sludge processed

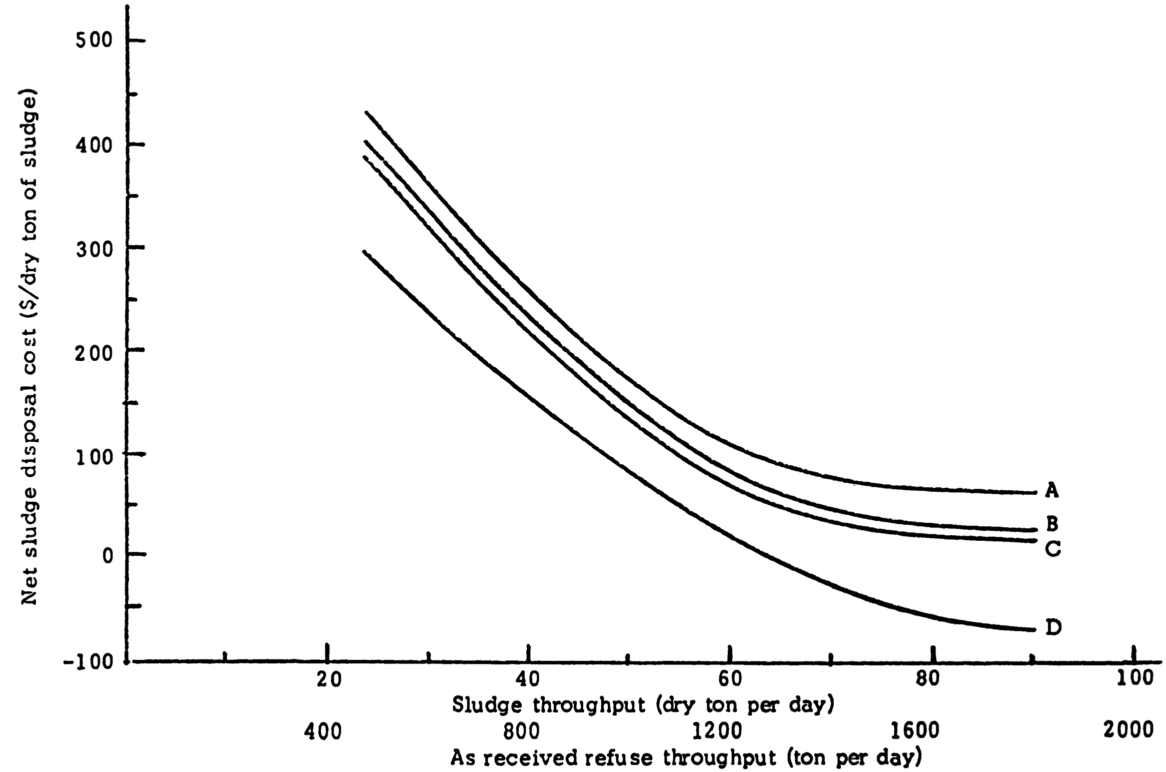


Figure 6. Estimated net sludge disposal cost with tip fee of $15/ton of refuse: 0.05 dry sludge to refuse ratio; $15.00/ton of refuse

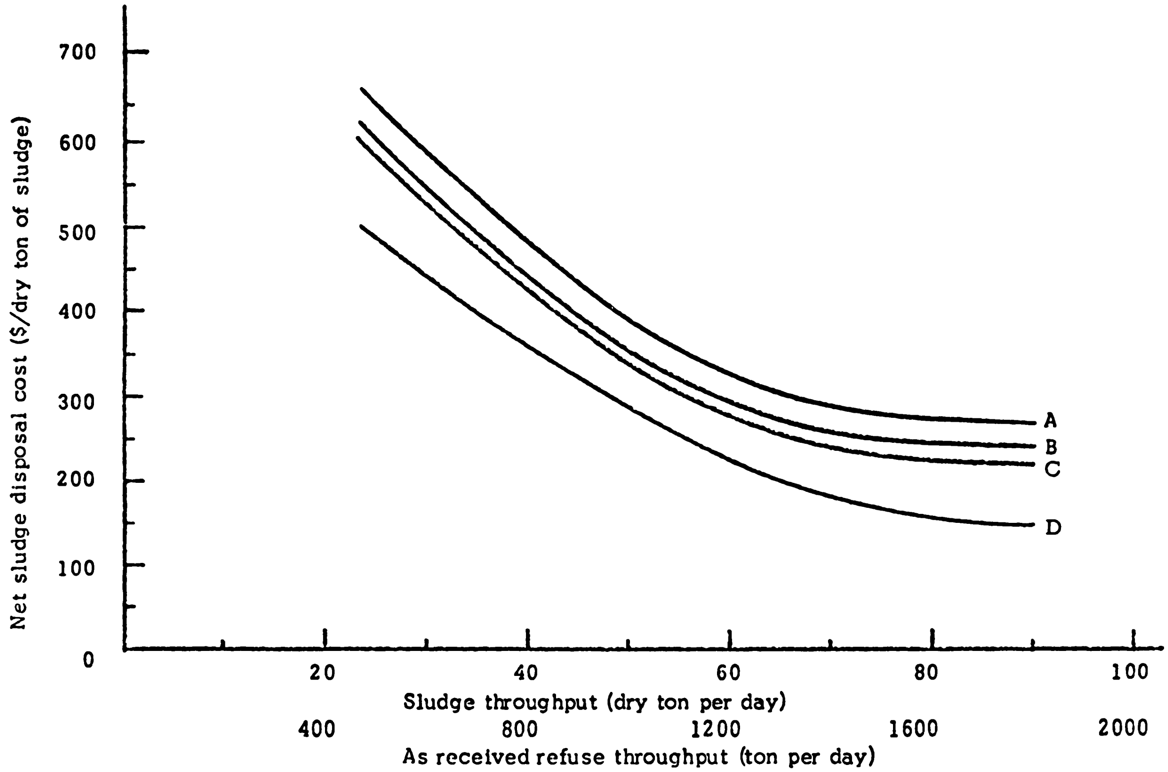


Figure 7. Estimated net sludge disposal cost with tip fee of $5/ton of refuse: 0.05 dry sludge to refuse ratio; $5.00/ton of refuse

TABLE XIII. ESTIMATED ECONOMICS - PUROX SYSTEM CODISPOSAL FACILITY

Design Capacity = 400 tons per day of sludge dry solids
0.3 sludge dry solids/refuse

Capital Costs (in thousands of 1977 dollars)	
PUROX System, Oxygen Plant, Wastewater Pretreatment System	$ 57,000
Front End System	15,000
Site Facilities	13,000
Buildings and Site Improvements	4,000
Codewatering and Drying Equipment	22,000
Total	$111,000
Annual Operating Costs (in thousands of 1977 dollars)	
Facility Operation and Maintenance	$ 6,370
Facility Power @ $0.03/KWH and Production Materials	7,330
Levelized Amortization of Capital @ 7% for 20 years	10,500
Total	$ 24,200
Annual Revenues (in thousands of 1977 dollars)	
Refuse Disposal @ $15/ton of as-Received	$ 7,870
Fuel Gas @ $2.35/million Btu	9,000
Ferrous Metal @ $40/ton	1,470
Total	$ 18,340
Net Sludge Disposal Cost (in 1977 dollars)	$40/ton of dry sludge

cannot appreciably alter the economics without developing unreasonable sludge disposal fees. At higher sludge to refuse ratios the sludge disposal fee carries a larger portion of the total project cost thus presenting potentially more favorable economics.

ABSTRACT

CODISPOSAL OF SLUDGE AND REFUSE IN A PUROX CONVERTER

Combinations of dewatered sewage sludge and shredded, magnetically separated municipal refuse were processed in the slagging, oxidative pyrolysis process incorporated in Union Carbide's PUROX System. Both raw primary sludge at a solids content of $\sim$25% and mixtures of primary and secondary sludge at a solids content of $\sim$20% were tested. Sludge dry solids-to-as received refuse ratios varied from 0.016 to 0.052 in the program. The general conclusion of the testing was that codisposal in this fashion does not produce any significant changes from operation experienced with refuse only. The environmental effects of this process are of particular interest since they represent a significant improvement in the handling of heavy metals from sludge disposal. Economics of this process are very situation specific and are strongly tied to refuse disposal costs.

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MARCH 31, 1978.

5

The Caloricon Process for Gasification of Solid Wastes

HARALD F. FUNK[1]

68 Elm Street, Murray Hill, NJ 07974

Solid waste management practices such as ocean dumping, open burning and uncontrolled landfills are now being banned in most areas. Even incinerators which are not equipped with adequate pollution control devices have been or are going to be shut down.

The new approach is to recover usable portions of solid waste such as metals, glass and even paper while the remainder goes to landfills or is burned in modern incinerators which may be furnished with moving grates, water walls for steam generation, electric precipitators, secondary combustion chambers, scrubbers and high stacks. Unfortunately most of the incinerator models are not geared to process industrial waste because of the higher heating value of such materials or the pollution problems resulting from the combustion.

Saarberg Fernwaerme, a division of Saarbergwerke A.G. in Saarbruecken, Germany owns and operates several incinerators, which yield steam for either district heating or power generation. Since they are aware of the shortcomings of incinerators they wanted to build a pilot plant for waste gasification as an improvement.

The West-German Federal Ministry of Research and Technology decided to carry 60% of this 5 million dollar project for reasons of energy conservation and pollution control.

A pilot plant was built for an intake capacity of one ton per hour of solid waste, which by now has been in operation for more than 2 months. The technical feasibility has been established and now the design criteria have to be studied.

This system consists of the following stages of operation as can be seen in Fig. 1.

<u>Sorting</u> of metals and glass,
<u>Size reduction</u> to maximum lump size of 2",
<u>Gasification</u> in a shaft furnace applying an autothermic process combined with a carburetting effect by recycling

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[1] Current Address: A-5084 Grossgmain, 44, Austria

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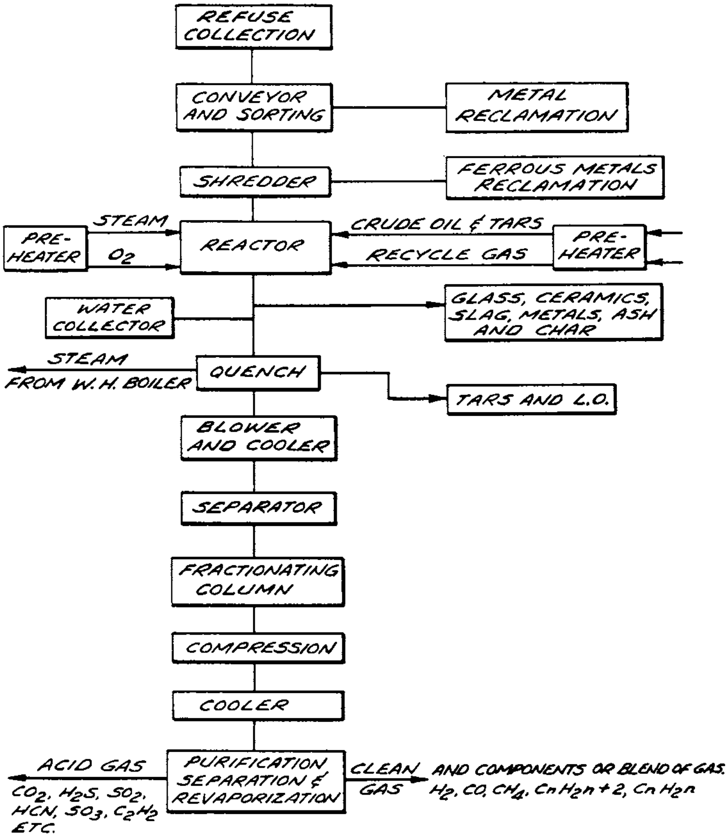

Figure 1. The Caloricon Process (U.S. Patent No. 3,970,524)

tars and oils,

Condensation section to separate tars, oils and condensate and compress the cooled gas,

Purification section applying a cryogenic technique to physically retain impurities,

Neutralization to annihilate undesirable components by chemical treatment. A U.S. Patent No. 3,970,524 has been granted for this type of process.

Utilization of the clean gas.

This closed system is designed to yield a gas of relatively high heating value. In comparison with an incinerator which sends 200,000 SCF of stack gas per ton of solid waste to a scrubber using a minimum of 120 Gal. of water for the purification, the Caloricon Process counts on an output of 20,000 SCF of gas at a lower heating value of 475 BTU/SCF, while only 4,000 to 6,000 SCF of impurities are discharged to a neutralization unit. Thus, the chemical treatment of the gasification plant processes only between 2% and 3% of the stack gas discharged from an incinerator.

This is one reason why a cryogenic method was adopted for gas purification.

The other reason was to allow for a synthesis gas preparation combined with this purification system. Then carbon monoxide and hydrogen can be fed to a methanol synthesis plant which provides an ideal use of the gas.

A shaft furnace has been chosen for this autothermic process initiated by a partial oxidation of the solid waste. The oxygen or oxygen enriched air is preheated as well as the steam which has to be charged if the material is too dry or the reaction becomes too hot. The solid waste is charged to the top of the furnace; it slides down through locks which prevent the escape of gas and passes through the various zones of heating, drying, carbonizing, reducing and partial combustion.

Enough oxygen is injected to maintain a temperature of at least 800°C but not more than 1000°C to prevent slagging which would not only affect the grating but consumes too much of the carbon and lowers the efficiency of the furnace. There is a certain flexibility in this operation since the gas carries more methane and C_2 fractions if the temperature is maintained at a lower level, while a water gas reaction is favored in the higher temperature range which might be preferable if the gas is used for synthesis gas makeup.

The oxygen consumption ranges from 10% to 17% per ton of solid waste as charged to the reactor.

Based on the following assumption of maintaining an exit temperature of gas and solids at a maximum of 300°C and a feed of 80% municipal and 20% industrial waste (in this case rubber tires) is charged to the reactor, the thermal efficiency was calculated to be 81%. If the light oils, however, are added to recoverable heating value the efficiency increases to 87% of heating value charged within solid waste. Deducting the energy consumption of the process it will result in an overall heat recovery of better than 70%. The average elemental analysis shows a composition of carbon at 56.3%, hydrogen 6.7%, oxygen 33.9%, nitrogen 1.4% and sulfur with other impurities 1.7%.

This material in comparison with coal which contains only 4% to 5% hydrogen is more suitable for gasification.

The gas emerging from the reactor is cooled to ambient temperature and separated from tar, oil and condensate. The latent heat of the vapors and steam can be utilized for district heating if facilities are available. In this phase of the operation, all the water is recovered which is charged to the reactor in form of moisture or steam.

Then the gas is compressed to about 30 PSIG and passed on to the purification section. The gas is processed here in 3 phases:

1. The loading phase
2. The subgas recovery phase
3. The cooling down phase

In the first phase, the gas is cooled down to at least minus 120°C in one run. The heavier components, mainly consisting of impurities are retained in the vessel while the pure gas passes on to an expansion turbine where it is cooled further by means of adiabatic expansion. This colder gas is returned through the last regenerator in a return flow thus preparing it for the next phase, the loading period.

The heavier components which were retained during the loading phase are recovered by means of a vacuum at an absolute pressure of 1.5 PSIA and this stream is diverted to the neutralization section for chemical treatment. The cycles are switching in periods of 6 to 8 minutes.

The regenerators are very efficient heat exchangers and the temperature differences between in and outgoing streams is kept between 2° and 3°C. This explains the low power consumption to compensate for the loss of refrigeration.

The clean gas can be used for various purposes:

A. Heating boilers,
B. Driving a gas turbine for power generation,
C. As a reducing gas in metallurgical processes,
D. For synthesis gas preparation to produce methanol,
E. For peak shaving to convert methanol to pure methane.

Methanol as such can be converted to a high octane gasoline as suggested by Mobil Oil Corporation which has developed such a process.

Methanol can be used as fuel for vehicles, gas turbines or boilers.

Methanol can be the basic raw material for a chemical process industry if acetic acid or formaldehyde has to be

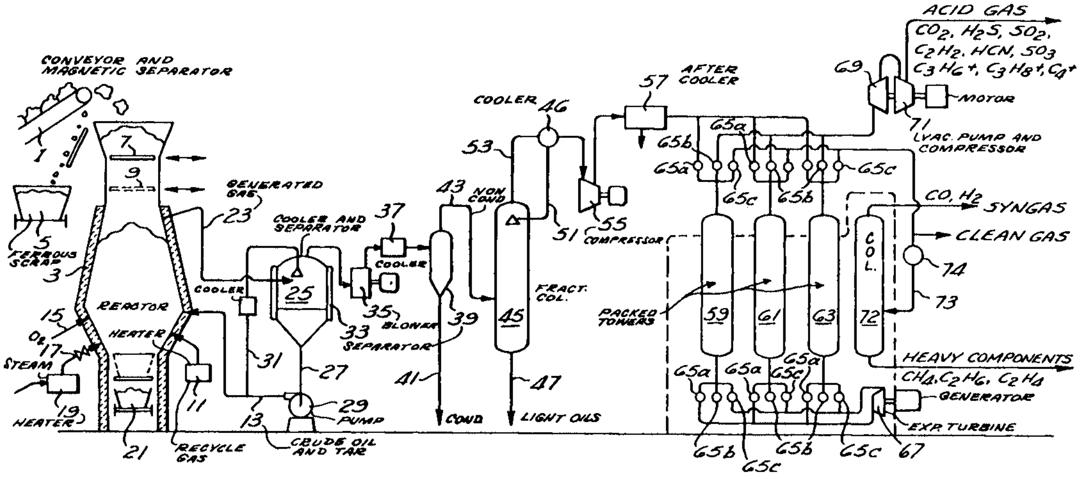

Figure 2. The Caloricon Process

produced.

Cities with a daily collection rate of about 5000 tons, for instance Pittsburgh, Cleveland or Toronto could produce between 600 and 1000 tons per day of methanol if all solid waste was to be gasified. Such a quantity of methyl alcohol would enable a city to fuel all its own vehicles and it could support a chemical complex.

An economic evaluation could be made only after operating conditions have been defined from the pilot plant, but it seems to be justified to assert that such a gasification plant which does not require moving grates, electric precipitators, secondary combustion chambers, scrubbers, and a stack should not cost more than incinerators but can offer a better utilization of the gas either for heating or for use as base product for chemicals, plastics or pharmaceuticals.

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6

A Vertical-Bed Pyrolysis System

M. D. BOWEN and E. D. SMYLY
Tech–Air Corporation, Atlanta, GA 30341

J. A. KNIGHT and K. R. PURDY
Engineering Experiment Station, Georgia Institute of Technology, Atlanta, GA 30332

The initial work on the Tech-Air pyrolysis system was started at the Georgia Institute of Technology in Atlanta, Georgia, approximately ten years ago. At that time the disposal of agricultural wastes due to new and impending air pollution regulations was a serious problem and a study was initiated to design a small, cheap incinerator which would meet emissions regulations and be suitable for use by small and intermediate sized, agricultural businesses. An evaluation of the inherent combustion characteristics of these ligno-cellulosic wastes indicated that such an incinerator would be large and very expensive. For example, an agricultural waste like peanut hulls, when heated in an air stream will first decompose emitting gases from the solid particles. These gases will burn in the gas phase in an envelope around each particle preventing oxygen from reaching the solid. This process continues until the peanut hulls have decomposed to pieces of char containing practically all of the sand imbedded in the original hulls and the other ash. The combustion of a piece of char is limited by the diffusion of oxygen to the solid surface. This results in a significant residence time and large excess air requirements for complete burning. Further, although the gases produced during decomposition of the hulls are gases with relatively low heating value, they have a low air to fuel ratio at stoichiometric conditions and, consequently, the flame temperatures observed are surprisingly high. It was clear from this and other early problems that a different approach to the disposal problem should be considered.

Historically, ligno-cellulosic materials have been used to produce chars for many different uses by means of thermal decomposition, or pyrolysis. This approach appeared to offer a number of advantages over incineration and resulted in thinking about the waste materials more in terms of utilization rather than disposal. This can significantly affect the economics of a particular case when the cost of pollution controls and fuel values are considered. Generally, pyrolysis can be considered as one very efficient means of increasing the options available for utilizing

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waste materials. In this case the wastes are converted into more readily useable, and environmentally more acceptable, fuels — char, gas, and oil.

The first pyrolysis unit built in this program was a retort approximately five feet high with a single air tube, an electric starter, and a movable grate for periodic char removal. The retort was built and operated with dry agricultural wastes in the late 1960's. Information on the process and potential products obtained from this work was used as the basis for designing and building at the Georgia Institute of Technology the first continuous pilot plant in 1970. This unit was designated Blue I and is shown in Figure 1. Some information on the pilot plant work has been reported earlier by Knight and Bowen (1, 2). The pilot plant was designed for a short life and minimum cost. The system incorporated a vertical bed, gravity fed, counter flow pyrolysis chamber with a continuously operating char output system. The input feed system consisted of the small storage bin shown in the front of the picture, an on-off screw at the base of the bin, a roto -lift and an input chute feeding the top of the rectangular shaped pyrolysis chamber. Two leveling screws running the width of the chamber were used to distribute the material fed in at the chute. In this system the process was initiated by means of a natural gas burner and sustained by bleeding a small amount of air low in the bed. The system was operated slightly above atmospheric pressure with the off-gases being taken off at the top of the chamber and burned in four flare stacks shown at the rear of the picture.

Blue I was operated for approximately one year at temperatures below 1600°F, on dry agricultural wastes. This plant demonstrated that the process could be operated continuously with simple means of control and, most important, that the mechanical char output system was reliable. Information obtained on the process showed the importance of distributing the process air within the bed and the wide variations in char yields which could be obtained without producing slag. The interior surfaces of the pyrolysis chamber were eventually damaged by overheating while operating at high temperatures. This was not surprising in view of the initial design criteria and the materials of construction used in Blue I.

The results of this first pilot plant program were very encouraging and led to the construction in 1971 of the second pilot plant, Blue II, shown in Figure 2. This system used some of the components from the earlier unit, but was designed for a longer life with refractory walls and more instrumentation. The input system consisted of a bin, a covered belt conveyor, and a rotary air lock. The off-gas system was changed significantly by installing an air cooled condenser, an off-gas control fan, and a refractory lined, swirl chamber for combustion of the non-condensed gases. The off-gas fan permitted the pyrolysis chamber to be operated at sub-atmospheric pressure and allowed the installation of simple, weighted doors for pressure relief on each component in the system.

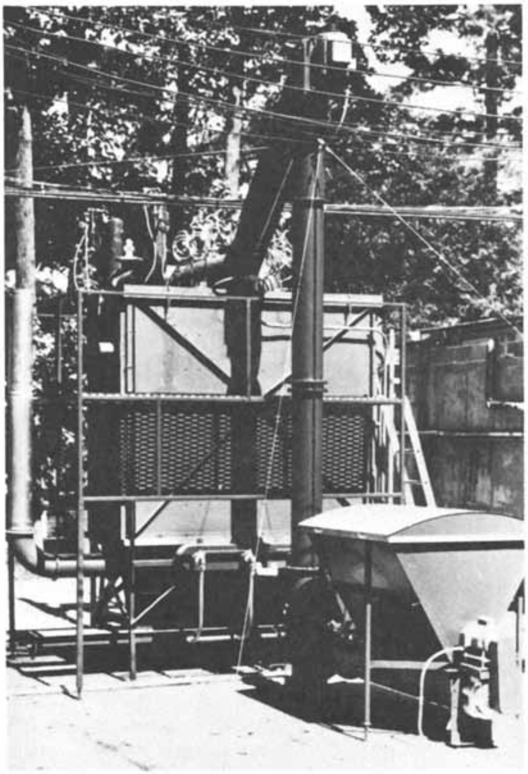


Figure 1. Pilot plant—Blue I, 1970–1971

Blue II was operated for approximately four years on a wide variety of feedstocks. These included bark, sawdust, wood chips, cotton gin trash, various nutshells, automobile shredder wastes and municipal wastes. In each case, significant variations in processing characteristics and in the quality of the different products, char, oil and gas were observed.

In 1971, based on the technology represented by Blue II, Tech-Air Corporation designed two field test units, each with a nominal capacity of two tons per hour. These systems were built and installed at a peanut shelling plant in Georgia. These two units are shown in Figure 3 while in operation. No provision was made for char storage and handling since these were test systems. For the most part the products were either landfilled or burned. However, one truckload of char was sold to the briquette industry for evaluation purposes. The two pyrolysis units were operated for approximately one year and resulted in a greatly improved air injection system, a simplified electric start system, better safety interlock devices and based on experience an insight into the scaling and material handling problems associated with this type of equipment. The successful operation of these units, although on a test and development basis, and using only dry agricultural wastes, resulted in a decision by Tech-Air Corporation to design and build a commercial prototype plant to process wood wastes. This decision was not surprising in view of the fact that 160,000 tons of peanut shells are produced each year in Georgia compared with several million tons of wood wastes. This design work was initiated in 1972.

Early in this work it became apparent that in the past attention had been focused primarily on the pyrolysis unit and the material handling, the metering and the drying problems required considerable work, particularly the drying of wood wastes. Tech-Air Corporation undertook the development of a compartmented, conveyor type dryer during this same period which could use waste heat or low temperature gases for drying. Consequently, the commercial prototype plant included the development of a completely new piece of major equipment in the system. This development work significantly extended the total time which otherwise would have been required for the first wood waste system.

The commercial prototype plant was installed in a small lumber mill in Cordele, Georgia in 1973 and was operated intermittently for approximately two years. During this period the dryer was developed and successfully demonstrated, the char sold to the briquette industry and work performed on burning the oil on a demonstration basis in several commercial applications. It should be noted that the servicing requirements of the off-gas system were high, but acceptable, for this plant operating intermittently on a one shift basis. Each morning prior to start-up the off-gas system which was cool after standing idle for sixteen hours could be opened and the accumulated solids and tars removed. Later, after placing the plant on a twenty-four hours per day,

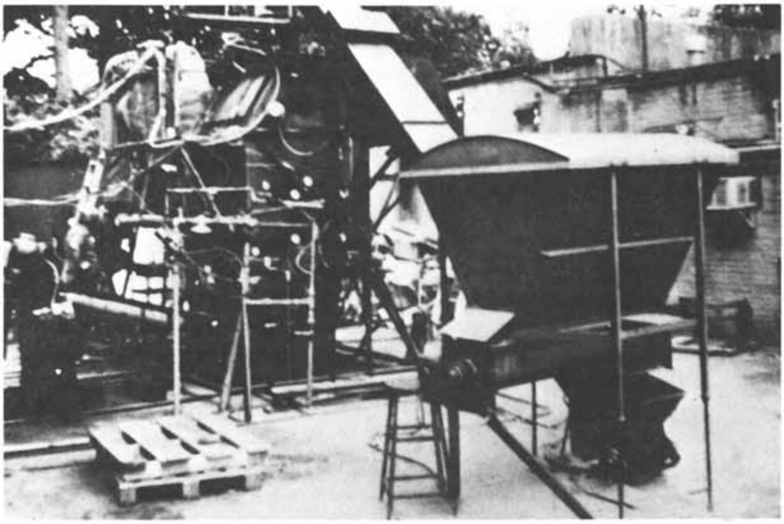


Figure 2. Pilot plant—Blue II, 1971–1974

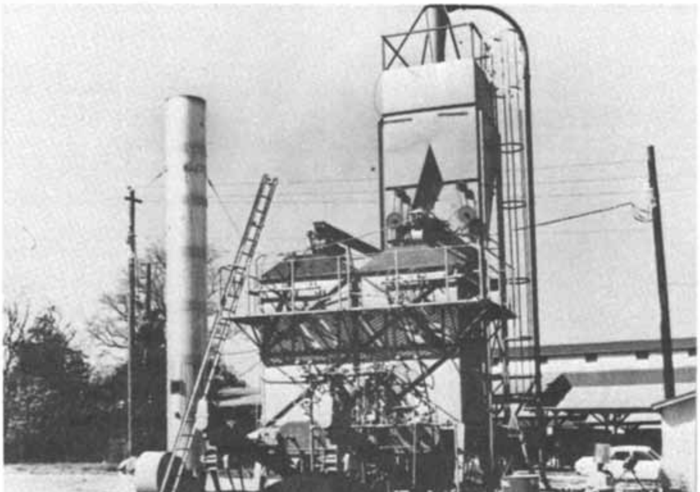


Figure 3. Field test units, 1971–1972

seven days per week basis, this servicing problem became very difficult and resulted in considerable down time for the system.

By this time pyrolysis was being considered by several companies for a variety of different materials or applications. The American Can Company through their resource recovery business unit, Americology, was involved in processing municipal refuse through a resource recovery system. American Can Company recognized the potential value of a pyrolysis system in this area to produce more readily acceptable fuel fractions, to reduce the volume of material, and to allow development of new char-based products. This interest resulted in the design and construction of a third pilot plant, nominally rated at one ton per hour, for processing municipal refuse. This pilot plant was designated Blue III and is shown in Figure 4.

This system was significantly different from the earlier ones due to the differences in feedstocks. A double gate, input air lock was used at the top, acid resistance refractory interior walls were installed in the pyrolysis chamber, a stirrer specifically designed to level and slightly compact the refuse was installed in the pyrolysis chamber and the output feed system was modified to pass large chunks, or pieces, of charred material.

Blue III has been in operation since 1974 processing a wide variety of materials. Approximately fifty-nine runs have been made on municipal refuse alone. These runs included light fraction, heavy fraction(with and without metals), whole garbage, sewage sludge blended with light fraction and light fraction blended with shredded tires. These runs demonstrated that the technology developed for agricultural and forestry wastes could be applied successfully to processing municipal refuse and sewage sludge. Work in this area of application is currently continuing, particularly in new product development and the next step for the municipal refuse application is a field test program for at least one year duration.

In 1975 Tech-Air Corporation became a wholly owned subsidiary of The American Can Company. At that time two different efforts were initiated to carry forward the work of commercializing the wood waste system and to establish a continuing research and development program at the Georgia Institute of Technology to support the area of waste utilization. First, a six month program was carried out to upgrade and extend the capacity of the commercial prototype plant and to permit a long term, around-the-clock operation. Second, a fourth, smaller pilot plant was constructed for further study and development of the process. This fourth pilot plant was designated Blue IV and is shown in Figure 5. The commercial prototype plant, known as the Cordele plant, is shown in Figure 6.

The weak links in the material handling system were identified early at the Cordele plant with around-the-clock operation. These were not major problems but they did require some beefing-up of certain conveyors and the establishment of a routine

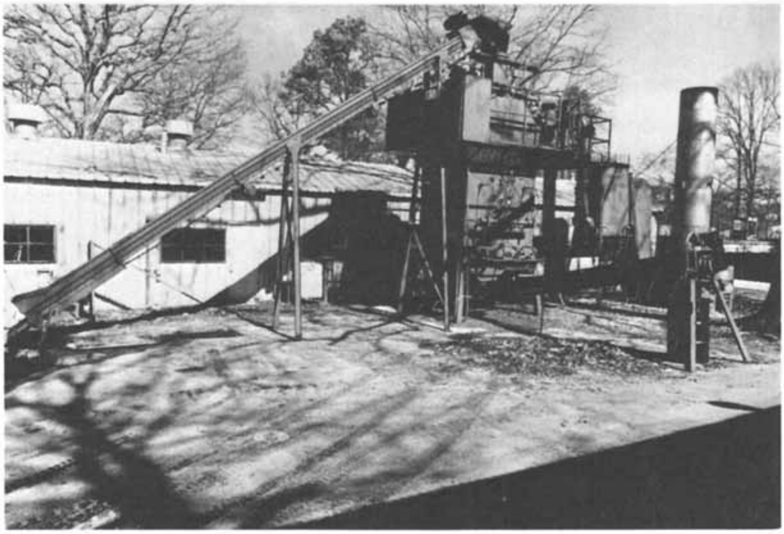


Figure 4. Pilot plant—Blue III, 1974–present

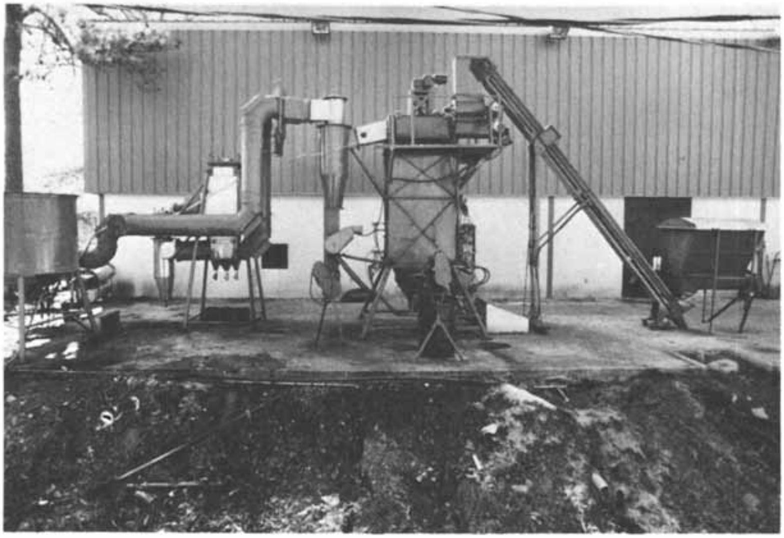


Figure 5. Pilot plant—Blue IV, 1975–present

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

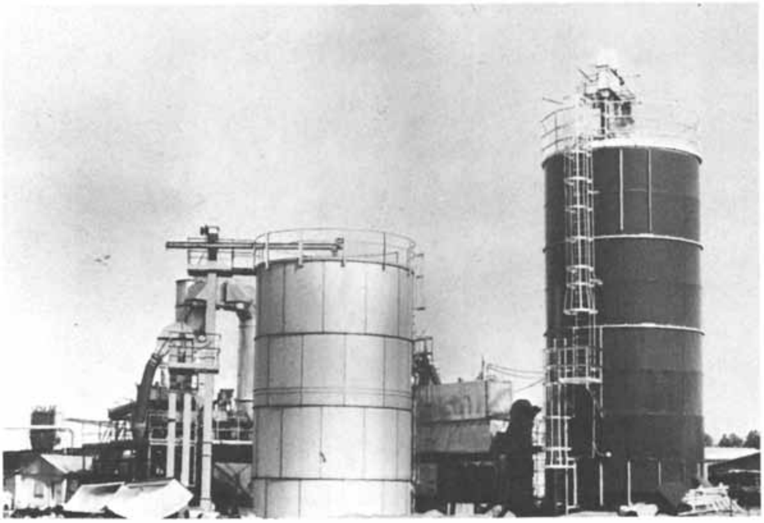


Figure 6. Cordele plant, 1973–present

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

servicing program. As mentioned earlier the largest problem encountered involved servicing or cleaning the off-gas system which resulted in considerable down time for the plant. This down time was drastically changed by the installation of a scrubbing system using a portion of the pyrolysis oil condensed from the off-gases. This technique was first considered in 1972 in the prototype design but the unquestionable requirement for its use accompanied the around-the-clock operation of the Cordele plant.

The Cordele plant was operated on a twenty-four hours per day basis for eighteen months until June of 1977. All the products, char and oil, produced at Cordele were sold during this period in the bulk char and fuel oil markets. With this successful demonstration the value of pyrolysis in waste utilization and resource recovery has been established. In retrospect it was indeed fortuitous that the direction of this effort started from a disposal standpoint which required the development of less complex and less expensive equipment to accomplish the process.

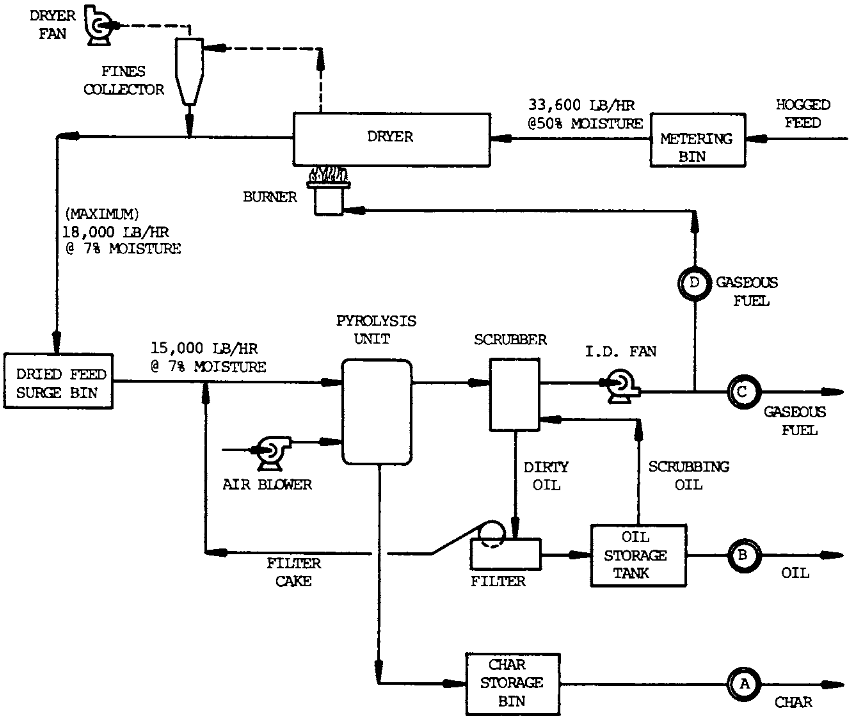
Wood Waste System

Plant Description. The process flow diagram for a 7 dry tons per hour wood waste system is shown in Figure 7. The system receives wet waste which has been hogged. The size of the hogged feed is not critical but a maximum particle size on one inch is desirable. The hogged wood waste is received in a metering bin which supplies a metered feed rate to the dryer. The dryer operates with some of the gaseous fuel generated in the process and pyrolysis oil can be used as a back-up. The dryer was designed by Tech-Air Corporation and is a compartmented, screw conveyor dryer. The inlet temperature to the dryer operates within the range of 400 to 600°F with a bulk exhaust temperature of 130 to 140°F. The dryer reduces the moisture content from a nominal value of 50 percent to a final value below 7 percent.

The dried feed is conveyed to the storage bin which supplies feed to the pyrolysis unit on demand and provides surge capacity. The dried feed from the storage bin is fed through a rotary airlock into the pyrolysis chamber where it is thermally decomposed into a char and a hot gas. The char is discharged at the bottom of the unit and the pyrolysis gases flow upward through the vertical bed and exit at the top of the unit. The rate of char discharge controls the throughput rate and a bed height sensing device is used to control the input to the chamber.

The char is discharged into a sealed conveyor, cooled with a water spray and fed through a rotary airlock into a char conveyor. The char is conveyed to a storage bin from which it is retrieved by gravity flow for subsequent shipment. The char bin incorporates a pressure-relief deck and low pressure, self-closing, relief doors.

The pyrolysis gases leave the top of the converter at a temperature ranging from 350 to 500°F and at a pressure near at-



PRODUCT MASS AND ENERGY YIELDS FOR A MIXTURE OF BARK AND SAWDUST

CASE	CHAR YIELD (LBS/ LB DRY FEED)	Ⓐ CHAR		Ⓑ OIL		Ⓒ GASEOUS FUEL TO PROCESS		Ⓓ GASEOUS FUEL TO DRYER	
		MASS (LB/HR)	ENERGY (MMBTUH)	MASS (LB/HR)	ENERGY (MMBTUH)	MASS (LB/HR)	ENERGY (MMBTUH)	MASS (LB/HR)	ENERGY (MMBTUH)
MAXIMUM CHAR	.35	4900	60.2	2520	22.8	2738	8.3	8116	24.6
MAXIMUM OIL	.25	3500	46.2	3150	28.7	4842	16.7	7132	24.6
INCREASED GAS	.20	2800	37.8	2660	24.4	7118	29.3	5976	24.6

NOTE - USED AVERAGE PROCESSING RATE THROUGH DRYER OF 15,000 LB/HR

Figure 7. Wood waste pyrolysis system—7 dry tons/hr

mospheric. The gas stream contains non-condensible gases, oil vapors, water vapor and entrained particulates. These gases are immediately sprayed with cooled pyrolysis oil in a scrubber-condenser which serves to remove the particulates and cool the gas stream to a temperature between 180 and 200°F. The cooling process is controlled to condense pyrolysis oil and limit the amount of water vapor condensed. The cooled gases flow through a rotary demister which removes the small liquid droplets from the gas stream. The unfiltered pyrolysis oil from the scrubber and demister is discharged through a rotary valve directly into a continuous filter. The filtered oil is pumped to a small holding tank from which the oil is recirculated through a cooler back to the scrubber. As the oil level in the holding tank increases due to condensation, oil is pumped to bulk storage. The filter cake, which has a dry appearance even though it contains about 30 percent solids, is conveyed back to the input feed system and reinjected into the pyrolysis chamber. Upon reheating of the filter cake, most of the oil is revaporized, but some is decomposed to a light gas, water vapor, and char.

The pyrolysis gases leaving the demister contain water vapor and some low boiling point fractions in addition to the non-condensible gases and this mixture is nearly saturated. Hence, it is desirable to utilize these gases close to the pyrolysis plant.

An induced draft fan controls the pressure in the pyrolysis chamber and directs the flow of gases through the off-gas system from the converter. A portion of the gases leaving the fan are piped to a burner supplying heat for drying the feed. The remainder of the gases are supplied to a burner which provides heat for a boiler or, other heat device. A flare stack is used to burn the gases during start-up or in case of a rapid change in gas demand.

Mass and Energy. The Tech-Air pyrolysis unit can be operated to provide a variable yield of char from the feed material with a corresponding variation in the gas and oil yields. For a given feedstock, the char yield is affected primarily by the maximum bed temperature which is controlled by the air to feed ratio. For purposes of presentation, the distribution of energy among the three products has been correlated versus char yield and is shown in Figure 8 for a mixture of pine bark and sawdust. These curves are based on data obtained at the Cordele plant. The Cordele plant has been operated at char yields ranging from 19.5 to 38.6 percent and the pilot plants have been operated at char yields from 8 to 45 percent.

The heating values of the three products vary with the char yield as shown in Figure 9. The energy recovered in the form of char continuously increases as the char yield increases with a corresponding decrease in the energy in the gaseous fuel. Note that a constant amount of energy is shown to provide the

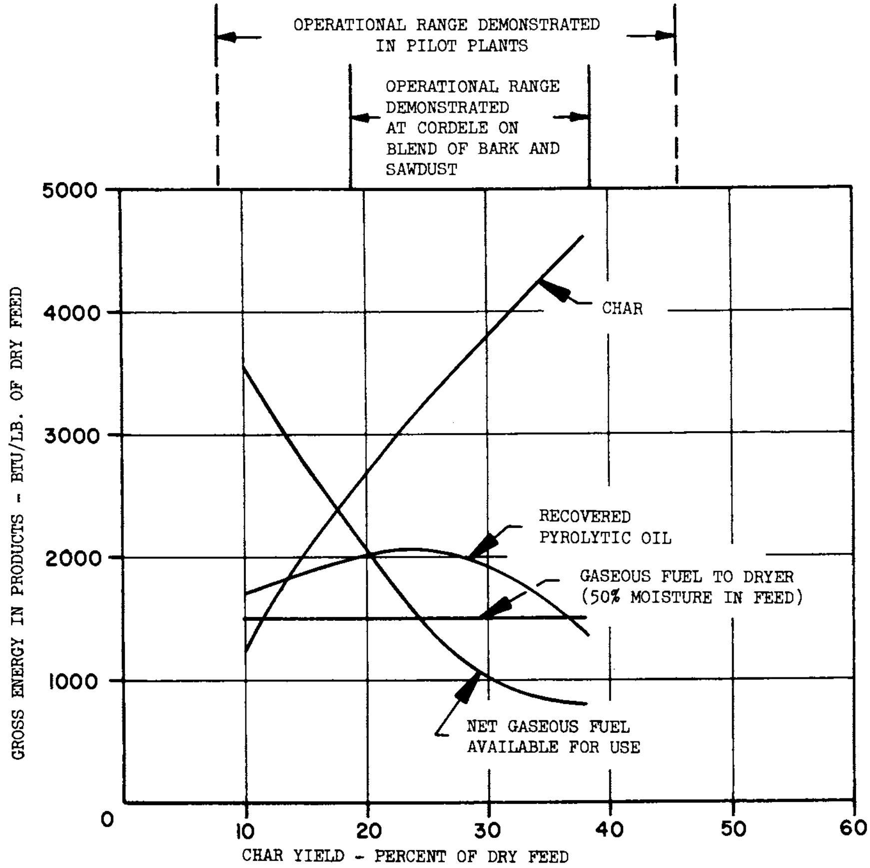


Figure 8. Energy distribution vs. char yield for product of pyrolysis from blend of pine sawdust and bark

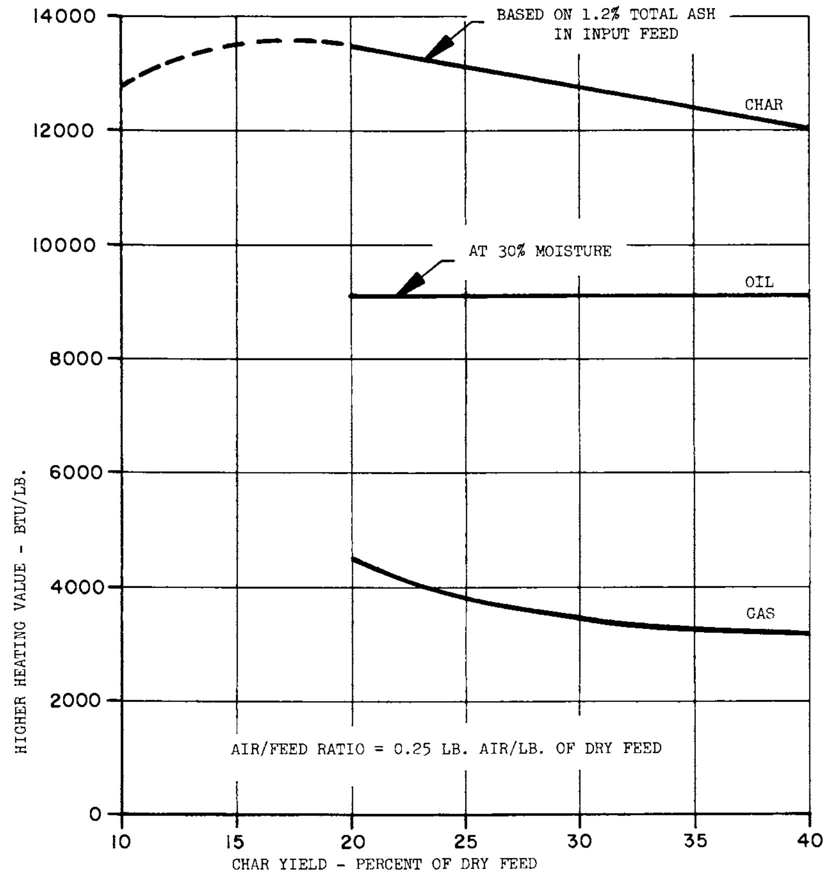


Figure 9. Higher heating value of pyrolysis products from pine bark-sawdust mixture

energy for drying the wood. The drying energy was approximately 1600 BTU per pound of water removed. The energy in the net gaseous fuel is that available after the drying requirement has been satisfied. The energy recovered as pyrolysis oil increases slightly as the char yield increases from 10 to 25 percent and then decreases.

The gross energy recovery in the form of products is approximately 95 percent of the energy in the dry feed. The higher heating value normally associated with a mixture of pine bark and sawdust is 8700 BTU per pound. Note in Figure 8 that the sum of the energies of the products at 10 percent char yield is 8000 BTU per pound of dry feed and at 35 percent char yield is 8260 BTU per pound of dry feed.

The energy distribution curves shown in Figure 8 point out the flexibility of the system. The process can be controlled to maximize those products of greatest value, and to match the system to on-site energy requirements.

The distribution of energy among the products, char, oil, and gas also varies with the type of feed material. Data are presented in Table 1 for pine tree top chips, pine bark, and a mixture of pine bark and sawdust. These data are from measurements at the Cordele plant and the Blue III and Blue IV pilot plants. Shown are the heating values of the input feeds, mass yields and energy yields. The portion of the energy contained in the char is approximately the same for all of the feeds at the same char yield. However, at a char yield of approximately 20 percent, the energy recovered as pyrolysis oil was highest for the chips and lowest for the bark and, as expected, the value for the mixture of sawdust and bark lies between these two. Also, the highest char yield obtained to date of 45.7 percent was from bark feed.

An initial study has been performed on the minimum char yield obtainable while operating the unit as a non-slagging, pyrolysis system. The data from this study, obtained from a 72 hour Blue IV pilot plant run, are presented in Table 2. The feed material used was pine chips. Two modes of operation were used. In the once-through mode the material made a single pass through the unit as is normal. In the recirculation mode, the char output was increased to an equivalent 20 percent yield and screened. The coarse particles were reinjected into the unit with the feed material and the fine char and ash passing the screen were removed. A char yield of 8.2 percent was obtained in the once-through mode. With recirculation of char, the net yield (fines passing the screen) was lowered to 3.8 percent. Even at the low char yields the unit was operated as a pyrolysis system with an oil yield only slightly below that obtained on pine tree top chips (see Table 1) at a 20 percent char yield. Char recirculation simply offers another option with pyrolysis, the efficient conversion of the wastes to gas and oil only. This techniques can also be used to avoid slagging conditions in the bed and obtain low char yields when the ash is high.

TABLE 1

SUMMARY OF MASS AND ENERGY DATA FOR PYROLYSIS OF SEVERAL WOOD FEEDSTOCKS

MATERIAL	HIGHER HEATING VALUE (DRY) (BTU/LB.)	PERCENT MOISTURE	MASS YIELD (PERCENT OF DRY FEED)			ENERGY YIELD (BTU PER LB. OF DRY FEED)		
			Char	Oil	Fuel Gas	Char	Oil	Fuel Gas
Pine Tree Top Chips	8935	5.7	20.5	34.6	76	2631	3498	2470
Pine Tree Top Chips	8935	5.8	25.6	31.9	71	3412	3174	1987
Pine Bark-Sawdust	8837	1.4	20.5	19.2	103	2768	2179	3067
Pine Bark-Sawdust	8609	7.5	24.4	23.7	103	3311	2249	2953
Pine Bark-Sawdust	8663	13.5	38.2	17.5	111	4664	1076	2222
Pine Bark	8912	12.2	18.9	7.8	116	2422	834	5250
Pine Bark	8801	7.4	45.7	12.8	88	5738	1235	1316

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

TABLE 2

SUMMARY OF DATA FOR MINIMUM CHAR YIELD TESTS ON PINE CHIPS

OPERATIONAL MODE	AIR INPUT (Lbs. per Lb. of Dry Feed)	MAXIMUM BED TEMPERATURE (°F)	MASS YIELD (Percent of Dry Feed)		ENERGY DISTRIBUTION (Percent of Energy in Dry Feed)			
			Char	Oil	Gas	Oil	Char	Heat Losses
Once Through	.56	1865	10.9	25.3	50.7	28.0	16.7	4.6
Once Through	.69	1885	8.2	32.3	47.1	36.1	12.2	4.6
Recirculation of Screened Char Into Feed	.92	2065	3.8	30.7	56.8	33.2	5.4	4.6

Municipal Refuse System

Plant Description. A system to process municipal refuse is somewhat more complex than a system for wood waste. The additional complexity results from the added difficulty in the handling and preparation of municipal refuse for pyrolysis feed. In a full-scale resource recovery facility the refuse is shredded and air classified into heavy and light fractions. The heavy fraction is processed further for the removal of ferrous metals, aluminum, and glass. Whether or not all of these operations are performed depends upon the quantity of refuse to be processed and its composition. Pilot plant runs have been made on light fraction, heavy fraction, single shredded whole refuse with the metals removed and on the co-pyrolysis of light fraction with both sewage sludge and shredded tires.

Although not limited to this, the municipal refuse system described herein is one applicable to small communities which have insufficient waste quantities to justify a complete resource recovery facility. Toward this end the system is designed to require minimum processing of the refuse as preparation for pyrolysis feed. Of course, the pyrolysis system can be efficiently used for the light and heavy fractions from a complete facility.

The process flow diagram envisioned for a system to process limited quantities of refuse is shown in Figure 10. The refuse is delivered to a receiving floor, placed in a conveyor and processed through a shredder. After the shredding operation the ferrous metals are magnetically separated. The shredded refuse with ferrous removed is then placed in storage.

From storage the feed material is fed at a metered rate to a dryer in which the moisture is reduced to approximately ten percent. The dried material is screened to remove glass and debris.

The process flow diagram given in Figure 10 includes the input of sewage sludge for co-pyrolysis with the dried refuse. This sludge addition is an optional feature which appears feasible if the quantity of sludge is small relative to the quantity of refuse. Data will be presented which were obtained on the Blue III Pilot plant for the co-pyrolysis of light fraction and a larger percentage of sludge than that shown in the figure. The mass and energy values given in Figure 10 are based on the co-pyrolysis of dried refuse and sludge. If the system is operated as shown without the addition of the sludge the average steam output will increase from 47,000 to 48,000 pounds per hour.

The pyrolysis portion of the system is very similar to the system for wood waste. There are differences in equipment design details, materials of construction, and the conveying and handling of the refuse, but the process functions are the same.

Mass and Energy. Mass and energy data for the pyrolysis of dried municipal refuse are presented in Figures 11 and 12, respectively. The data in these figures were obtained on the

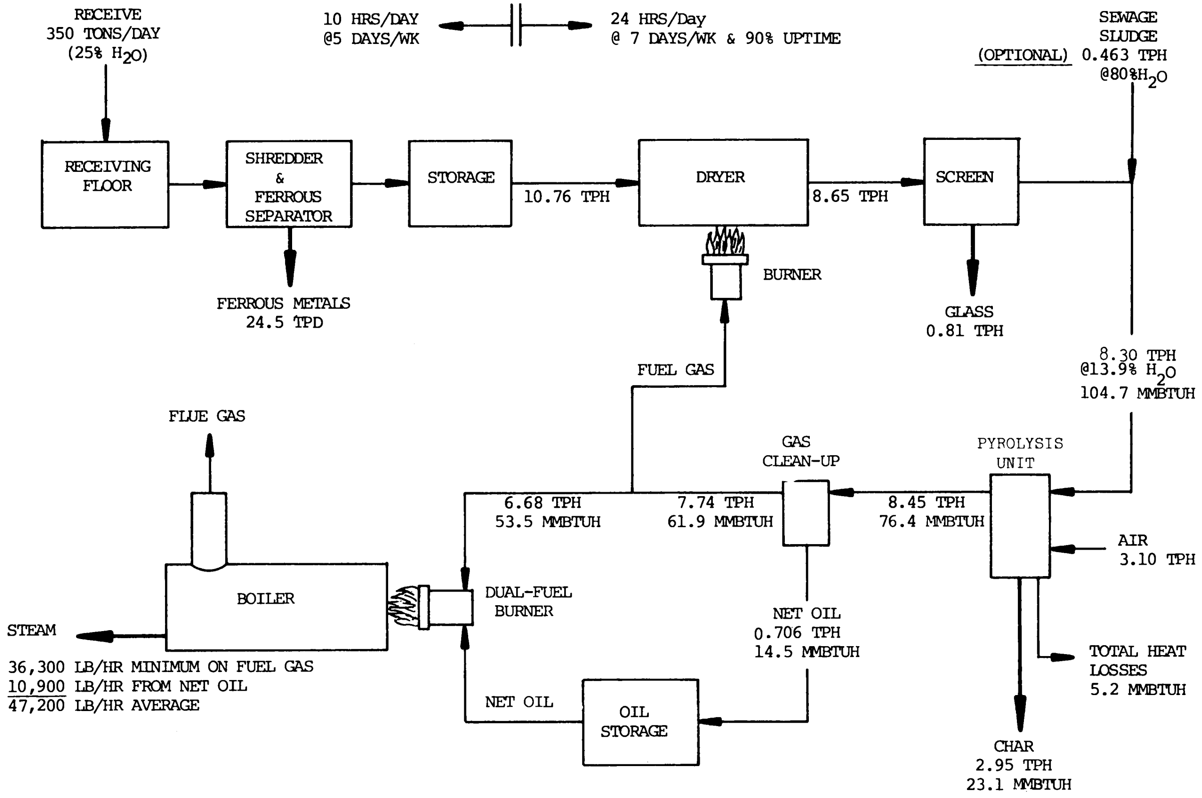


Figure 10. Small community municipal refuse pyrolysis system—350 tons/day

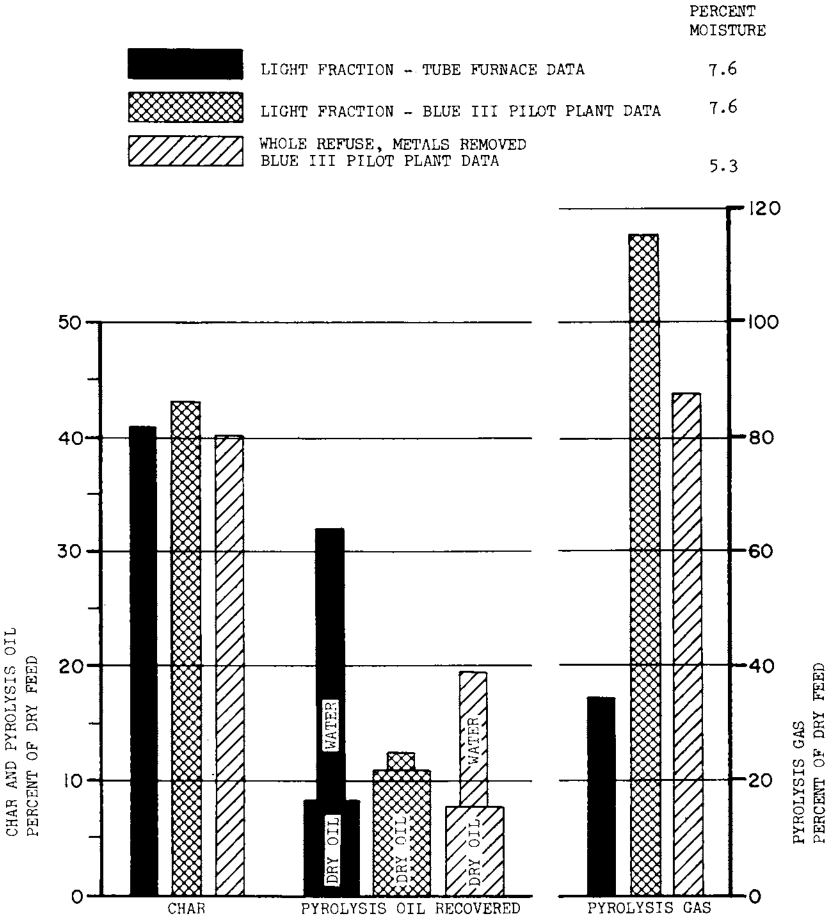


Figure 11. Mass balance for dried municipal refuse

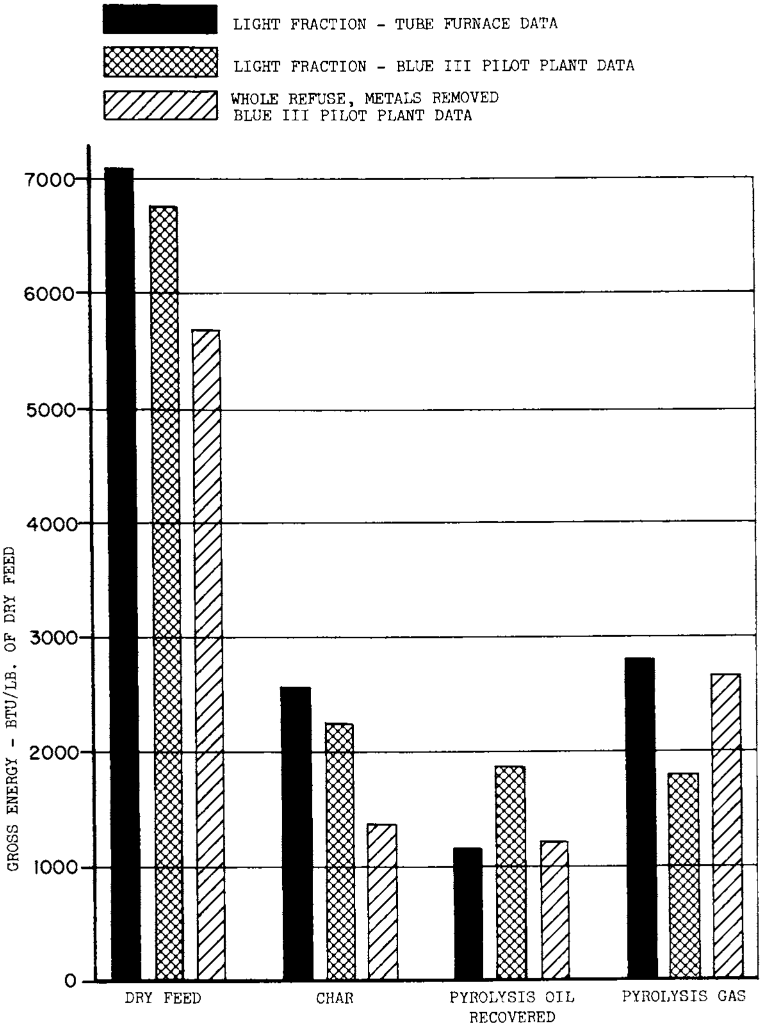


Figure 12. Energy balance for dried municipal refuse

light fraction of municipal refuse and on single-shredded, dried, whole refuse with the metals removed. For the light fraction, both tube furnace data and data obtained in the Blue III pilot plant are shown. For the single shredded, whole refuse with metals removed, data obtained in Blue III are presented. In the tube furnace, electrical heating is used and the material is pyrolyzed with no air input. The Blue III pilot plant operates on the same partial oxidation principal as the Cordele System.

The mass balance data in Figure 11 give the yield of char, the yield of oil actually condensed from the gases, and the yield of gas including the non-condensed oil. Approximately the same yield of char was obtained for the light fraction in the tube furnace and in Blue III. Further, the yield of char from the single-shredded, whole refuse was approximately the same as the yield from the light fraction. The values shown for the recovered oil show the total oil yield broken down into water and dry oil, which is the moisture-free fraction. The tube furnace data show a very high moisture content since all liquid was recovered with a cold-trap as opposed to the pilot plant operation which sought to minimize water condensation with the oil. The oil from the single shredded, whole refuse contained more water than the oil from the pilot plant run on light fraction because of poor control. The oil yield on a later run in the pilot plant with single-shredded, whole refuse was 19.5 percent with only 11.3 percent water. The gas yields shown in Figure 11 represent the total vapor stream after the condensation of the recovered oil. The muffle furnace gas yield was the lowest since there was no air input. The higher yields for the pilot plant runs reflect the air input.

The energy yields are more difficult to define than the mass yields. For the mass balances the quantities of feed material, char, condensed pyrolysis oil and air input were measured directly. Thus, the quantity of gaseous fuel could be directly determined by difference. From an energy standpoint the energy associated with the char and oil could be reasonably well characterized since representative samples could be obtained. For the gaseous vapors, the non-condensible portion could be accurately determined from gas chromatograph analysis of representative samples; however, the quantity of aerosols or low-boiling point fraction were extremely difficult to measure accurately. Again, the energy in this fraction can be calculated by difference if the heating value of the input feed is accurately known. Experience has shown that it is very difficult to accurately sample the input feed for heating value measurements.

Experimental data for the higher heating value of dry municipal refuse is plotted in Figure 13 as a function of ash content. Note that there was a wide variation in the ash content of the samples taken. In treating the experimental data from the pilot plant, the average heating value measured on samples of the input feed was adjusted to the ash content calculated from the char

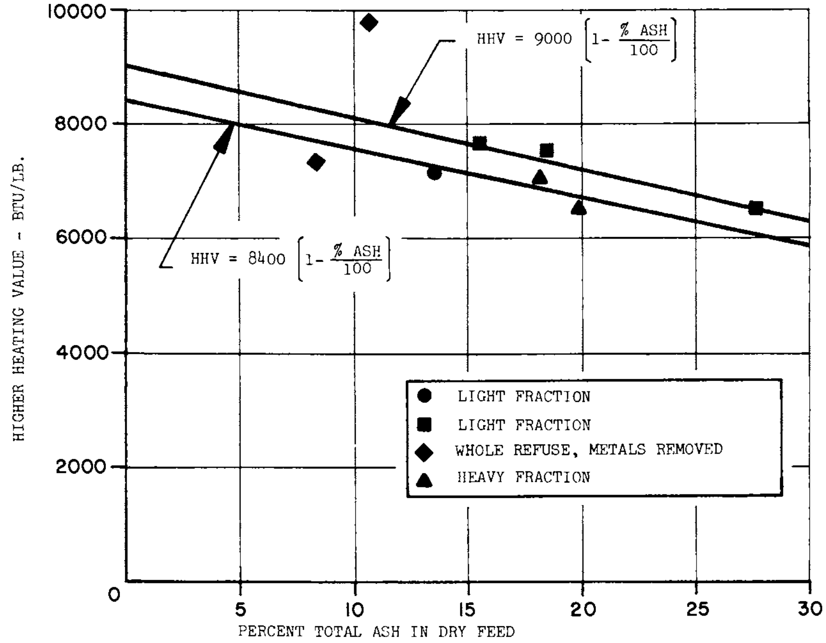


Figure 13. Higher heating value of dry municipal refuse correlated vs. total ash content

yield and ash content of the char. This value was then used to calculate the energy in the condensible fraction which remained in the gaseous vapors. The heat losses were calculated by assuming that each pound of input air released 1333 BTU which was lost as sensible or latent heat.

The energy data for light fraction and single-shredded, whole refuse with metals removed are presented in Figure 12. The heating value for the dry feed was 5700 BTU per pound for the single-shredded, whole (metals removed) refuse and approximately 6900 BTU per pound for the light fraction. For the pilot plant run the energy was reasonably well distributed among the products for the light fraction, with values of 1750 BTU per pound of dry input for the gas, 1850 BTU per pound of dry feed for the oil, and 2200 BTU per pound of dry input for the char. For the single-shredded, whole (metals removed) refuse the gas contained the largest fraction of the energy, 2650 BTU per pound of dry input, and the oil contained the least, 1200 BTU per pound of dry input. For the subsequent run made on single-shredded, whole refuse (metals removed) refuse, which was mentioned earlier, the oil contained 11.3 percent moisture and approximately 2300 BTU per pound of dry input.

Data from pilot plant runs in Blue III on the co-pyrolysis of light fraction with sewage sludge and light fraction with shredded rubber tires are presented in Table 3. The average moisture content of the feed material for the run with light fraction and sewage sludge was 16.2 percent. This high moisture content required a significant increase in process air which increased the mass yield of gaseous fuel. The energy yield of the char was 2400 BTU per pound of dry input for the light fraction-sludge blend and 2808 BTU per pound of dry input for the light fraction-rubber blend. The energy recovered as oil was 674 BTU per pound of dry input for the light fraction-sludge blend and 1002 BTU per pound of dry input for the light fraction-rubber blend. The energy yield of gaseous fuel was 2516 BTU per pound of dry feed for the light fraction-sludge blend and 3048 BTU per pound of dry feed for the light fraction-rubber blend.

Pyrolysis Products from Wood Waste

Characterization. Typical average heating values of the char, oil, and gaseous fuel from the pyrolysis of a mixture of pine bark and sawdust are given in Figure 9 at various char yields. Generally, measured heating values have ranged from 12,300 to 13,500 BTU per pound. The elemental carbon content of the char can be as high as 90 percent. The bulk density of char from hogged wood waste is normally within the range of 10 to 13 pounds per cubic foot. The volatile content has been measured within the range of 3 to 26 percent, depending upon operating conditions and feed material.

Typical heating values for the gaseous fuel are shown in

TABLE 3

SUMMARY OF MASS AND ENERGY DATA FOR CO-PYROLYSIS OF SEWAGE SLUDGE WITH LIGHT FRACTION MUNICIPAL REFUSE AND SHREDDED RUBBER TIRES WITH LIGHT FRACTION OF MUNICIPAL REFUSE

MATERIAL	HIGHER HEATING VALUE (BTU/Lb)	PERCENT MOISTURE	MASS YIELD (Percent of Dry Feed)			ENERGY YIELD (BTU per lb. of Dry Feed)		
			Char	Oil	Fuel Gas	Char	Oil	Fuel Gas
Dried Light Fraction with 13 Percent Wet Sewage Sludge (80% Water) 5/21/75	6590 (1)	16.2	47.5	10.9	136	2400	674	2516
Dried Light Fraction with 20 Percent Shredded Rubber Tires 1/15/75	7391(2)	6.0	45.7	17.2	83	2808	1002	3048

Notes:

1. Measured heating value was 7631 BTU/Lb. at average ash content of 20.1%. Ash content of feed calculated to be 31 percent based on char yield and ash content in char. Heating value corrected to 31 percent ash.

2. Based on ash in char, heating value of lights was estimated to be 6000 BTU/Lb. Heating value of shredded rubber was measured to be 12,600 BTU/Lb. These values were used to calculate the average heating value of the input.

Figure 9 at various char yields. The heating value decreases as the char yield increases. Normally, the gas is at a temperature of approximately 200° F and the density is approximately 0.05 pounds per cubic foot. At low char yields, the volumetric higher heating value will generally be on the order of 225 BTU per actual cubic foot at 200° F and at a moisture content of approximately 30 percent by weight.

Typical data for the pyrolysis oil are presented in Table 4. Note that the moisture content is 26 percent. Normally, the condensation process is operated to provide an oil with a moisture content of about 20 percent. This quantity of moisture serves to lower the viscosity for improved handling and atomization and appears to enhance combustion. At 26 percent moisture as shown in the table, the heating value is 9081 BTU per pound or approximately 90,000 BTU per gallon which is 60 percent of the heating value of No. 6 Fuel Oil. The viscosity of the oil can be reduced below 100 SSU by heating to 150°F. The flash point of the oil places it in the Code 1 category for explosivity. The oil is highly corrosive to mild steel, but the corrosive rate is a fraction of a mil per year for 304 stainless steel and copper.

Uses. The char from the pyrolysis of wood waste is widely used in the manufacture of charcoal briquettes. In addition, it has fuel value as a low sulfur coal extender or substitute and can be used to produce a water grade activated carbon. The combustion of pulverized char from wood waste produced by the Tech-Air system was studied in a joint program by EPA and ERDA (3). Carbon combustion efficiencies were 97.3 to 98.6 percent in a pulverized coal fired test facility and NO_x emissions were much lower than those obtained with coal. Also, blended with coal, the SO_2 emissions were much lower than the coal alone. A blend of pulverized char and pyrolysis oil combined with No. 6 fuel oil performed well.

Char produced from wood waste at the Cordele plant has been activated in the laboratory to produce a water grade activated carbon. In addition, laboratory studies have demonstrated the effectiveness of the activated char in the removal of color from kraft mill effluents.

The pyrolysis oil has been demonstrated as a fuel. It also has a potential as a chemical raw material. The oil has been sold commercially for use as a fuel in a cement kiln, a power boiler and a lime kiln. In the cement kiln, several months production from Cordele was fired as a 20 percent blend with No. 6 fuel oil. The remainder of the oil produced has been sold as a fuel which was fired in parallel with several No. 6 fuel oil guns in a power boiler and also directly fired in a lime kiln. Prior to these applications the oil was test fired in a Trane Thermal Vortex burner (4) and at KVB (5) in a test boiler system. The burner used at Trane Thermal performed equally well with air or steam as the atomizing media. The maximum oil pressure required was 3 psig.

TABLE 4

TYPICAL DATA FOR PYROLYSIS OIL FROM PINE BARK-SAWDUST

ITEM	UNITS	VALUE
CHEMICAL COMPOSITION		
CARBON	% by weight	49.4
HYDROGEN	"	4.7
OXYGEN	"	19.7
NITROGEN	"	0.16
WATER	"	26.0
ASH	"	0.04
VISCOSITY		
@ 68°F	SSU	276
@ 100°F	"	114
@ 150°F	"	90
HIGHER HEATING VALUE	BTU/LB	9081
DENSITY	LB/GAL	9.88 to 10.27
FLASH POINT (Open Cup)	°F	262 to 305
POUR POINT	°F	+2

TABLE 5

TYPICAL DATA FOR PRODUCTS FROM PYROLYSIS OF SINGLE SHREDDED, DRIED MUNICIPAL REFUSE WITH METALS REMOVED

OIL

HIGHER HEATING VALUE	(BTU/Lb.)	10,400
WATER CONTENT	(Wt. Percent)	30
DENSITY	(Lb./Gal.)	9.5

FUEL GAS

COMPOSITION (Wt. Percent)	
Water Vapor	26
Pyrolysis Oil	13
"Dry" Gas	61
HIGHER HEATING VALUE (BTU/Lb.)	3081
DENSITY @ 200°F (Lb./Ft.^{3})	0.05
HIGHER VOLUMETRIC HEATING VALUE @ 200°F (BTU/ACF)	154

CHAR

HIGHER HEATING VALUE	(BTU/Lb.)	3972
TOTAL ASH	(Wt. Percent)	69
ACID INSOLUBLE ASH	(Wt. Percent)	60
DENSITY	(Lb./Ft.^{3})	20

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

The unburned hydrocarbons were measured to be 11 ppm at 6 percent excess air and the combustion chamber temperature was measured at 2520°F. In the tests at KVB the carbon monoxide level was about 15 ppm, and the NO_x level was about 110 ppm at 4.5 percent excess air. The oil, also, has a potential use as a fuel for a hot gas turbine. Research work along these paths is expected to be performed in the near future.

Work done in the laboratory has indicated the potential of the pyrolysis oil as a raw material for the production of phenolic resins and rubber tackifiers. Other work has shown that it may be used as a substitute for wood tars produced by other methods.

The gas from the pyrolysis system is useful as a fuel. At the Cordele plant, the gas is burned as a fuel to supply heat to the dryer. Preliminary studies have been made of the use of the gas as a fuel for an internal combustion engine. A dry gas mixture which simulated the composition of the non-condensible pyrolysis gas fraction was run in an internal combustion engine study (6). The higher heating value of the mixture was 180 BTU per cubic foot. The power output of the engine was slightly over 60 percent of the output with gasoline.

Calculations have shown that it is reasonable to expect that the heating value of the gas can be increased to around 400 BTU per cubic foot so that engines already developed for sewage gas can be used. These modifications would involve removing all of the condensible fractions from the gas and reducing or eliminating the process air. Preliminary estimates have indicated only a small penalty in energy with incineration of the low-energy content liquid effluent produced in this case.

Pyrolysis Products from Municipal Refuse

Characterization. The products of pyrolysis from municipal refuse are char, oil, and gas. The yields and the properties of the products are related to the make-up of the input feed, the degree of shredding, the moisture content of the feed, and the pyrolysis conditions. Classification of the refuse prior to pyrolysis is one direct means of influencing the quality of the products. For example, a char which can be readily upgraded and improved can be produced from light fraction. With whole garbage it is more difficult to utilize char except as a filler or coloring material for brick and similar products.

The oil obtained from single-shredded garbage is usually more viscous than that obtained from wood. Its heating value ranges between 10,000 and 11,000 BTU/pound (at 30 percent moisture content), which is higher than the heating value of oil from wood. The density of both oils is very similar, approximately 9.5 pounds per gallon.

The fuel gas produced from garbage will vary due to the physical characteristics of the refuse as well as the composition of the refuse. For example, the permeability of a bed of single-shredded refuse is very low compared to wood wastes and usually results in operating at higher air-to-feed ratios, which directly affects the heating value of the off-gas. Both agitation of the bed and improvement in the distribution of the injected air can be used to improve this situation. Generally, the gas will contain some non-condensed oil and water vapor with the bulk being a non-condensible fraction. Its heating value usually lies in the 3000 to 4000 BTU/pound range resulting in volumetric values from 150 to 200 BTU per actual cubic foot.

The char produced from refuse is usually very high in ash content, particularly acid insoluble ash. Its heating value usually is below 4000 BTU/pound with a total ash of 65 percent or higher. This char will contain little or no slag, will be relatively free flowing and dusty, and have a bulk density of 20 pounds per cubic foot or higher.

Uses. Oil production results from deep bed pyrolysis with low air-to-feed ratios. In the Tech-Air system this is required in order to have a suitable scrubbing medium, and this is the initial use of the oil. In addition, the oil from garbage is a storable, transportable fuel oil and has a potential use in the production of carbon black, but little has been done in this area to date. Classification of the refuse into specific fractions is probably the more realistic approach to producing an oil from garbage that would have value outside the fuel area. This is not meant to suggest that its fuel oil value is low. To the contrary, energy production from garbage at the present time is the largest, single, economic factor involved.

The fuel gas from garbage, although its heating value is low, is an acceptable fuel suitable for use in the vicinity of the plant. The gas has a very low air-to-fuel requirement for combustion which results in an attractive ratio of total pounds of combustion products per unit of heat released which is not too different from the more common fuels. Also, since it is a scrubbed fuel gas it can be readily adapted for use in many different heat devices.

Further, it is well known that low BTU gas can be used in an internal combustion engine. However, because of the derating involved, and the possibility of variable loading, its use directly for this purpose without drying, or upgrading, is questionable. At the present time reduction of the air requirements and moisture removal from the gases appear to offer the best approach to producing a gas suitable for an internal combustion engine. Research work in this area is being conducted by Tech-Air Corporation and the Georgia Institute of Technology at the present time.

An additional potential use of the gas can be suggested in view of the low air-to-fuel ratio characteristic. A clean, dry fuel gas can be compressed for efficient use in a hot gas turbine provided the compressor work on the fuel can be offset with a similar, or larger, reduction in air compression work. This appears to be possible with gases produced in the pyrolysis process.

Char produced from municipal refuse, generally, is not considered of significant value. For this reason a pyrolysis system operating on refuse is primarily aimed at gas and oil production, minimizing the energy loss in the char. Research work has been done on upgrading char from light fraction refuse, and it holds significant promise in the area of sewage treatment. Generally a carbon with an acceptable adsorptive capacity can be produced from the light fraction. Research work in this area has been underway by Tech-Air for some time. This effort is expected to intensify and increase in the future.

Discussion

The development of the Tech-Air pyrolysis system has been briefly traced from the initial laboratory work in the 1960's through the construction and operation of four pilot plants, two large scale field test systems and one large, commercial prototype plant, the Cordele plant. This work was accomplished over a ten year period and culminated in 1977 with the successful demonstration for eighteen continuous months of a commercial prototype plant operating on wood waste.

The Tech-Air process is based on a vertical bed, gravity fed, counterflow pyrolysis chamber which probably gives the best heat balance attainable. To accomplish this process the distribution of the air and the permeability of the bed are very important. The air is injected in a manner so that partial oxidation of the gases occurs rather than complete burning in a small region. The result is the ability to produce char yields equivalent to that obtained in the complete absence of air.

Another important feature of this process is the mechanical feedout mechanism for the char. The device, originally designed for Blue I, has proven very reliable and adaptable to municipal refuse systems which occasionally require the passage of large chunks. This one device permits considerable automation to be introduced into the process. An output rate can be fixed by selecting the rotation rate of the output mechanism and with a bed level sensor controlling the input rate directly, the throughput rate is effectively controlled by the one device. To complete the process control the air input can be controlled within predetermined limits by temperature sensing at points within the bed. An alternative to this arrangement is to select an output rate and the desired off-gas temperature and vary the bed level continuously to maintain this temperature. This technique has been used successfully in both the pilot plant operations and at the Cordele plant. This technique is easily accomplished since the off-gases can be rapidly changed by raising or lowering the bed level. Approximately 40°F change in off-gas temperature occurs with a change of only one inch in bed level. This procedure, however, results in a greater departure from what could be called equilibrium bed conditions.

On first inspection, drying the wastes in a separate step may appear to be undesirable, but experience with this system has shown that the added complexity is more than off-set by the gains. The latent energy associated with the moisture will be extracted with any process or device. If this moisture is evolved within a device utilizing the energy in the solid there are additional penalties involved. If it is being burned, as in a bark boiler, additional air must be supplied to overcome the decrease in oxygen concentration due to the presence of the moisture. The increase in the mass of the flue gas due to the moisture and the additional air results in an increase in the sensible heat losses in the flue gases. Further, the increased mass results in higher velocities through the boiler which, in practice, results in operation well below rated capacity. In effect, if pre-drying can be accomplished at approximately 1500 BTU per pound of moisture removed, there will be little or no net energy penalty involved with a separate drying step.

On the other hand, the gains to be had with dry feed material are significant in pyrolysis, as well as other processes. The throughput is greatly increased, the temperature control of the process is easier and the products produced are of higher heating value.

The Tech-Air pyrolysis process is a low temperature, non-slagging process which produces char, oil and gas. In cases of high ash feedstocks, or if it is desirable to produce oil and gas only, it has been demonstrated that the process can handle the feed and gasify the char by the simple expedient of raising the air to feed ratio and recirculating char, but maintaining a significant char output from the bed. This technique has demonstrated that 90 percent of the energy in the dry feed can be recovered in the form of gas and oil. In view of these data, considering the much higher net thermal efficiencies achievable with oil and gas compared to burning a wet solid, the value of pyrolysis in waste utilization becomes clearer.

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7

Gasification of Solid Waste Fuels in a Fixed-Bed Gasifier

L. K. MUDGE and C. A. ROHRMANN

Battelle–Northwest, P.O. Box 999, Richland, WA 99352

Gasification of solid fuels is a technology that has been around for many years. The process of solids gasification with air is as old as steel. Industrial use of combustible gas produced from organic solids started with blast furnaces during the industrial revolution. Probably the original gas generator was the blast furnace, in which iron was smelted in the presence of coke. The combustible gases produced from the coke were used for various heating purposes by the year 1800.

By the year 1840 gases generated by the partial oxidation of coal, charcoal and peat were used for heat generation. Gas generators were in operation in Europe for the specific purpose of producing combustible gases.

Research and development studies on solids gasification at Battelle Northwest were initiated under private sponsorship in 1969 for the development of a process to recover silver from photographic film. In a subsequent one-year study sponsored by EPA and the City of Kennewick, Washington, the process was applied to the conversion of municipal refuse to a low Btu gas.

Recent studies have been sponsored by DOE and a number of private concerns. In these studies, a variety of materials, including wood, agricultural residues, shredded rubber, and mixtures of plastics with cellulosic wastes, have been processed in a small-scale fixed bed gasifier.

This paper describes the gasification process and equipment used for conversion of a variety of solid materials to a low Btu gas. Operating results obtained in the studies are presented. Preliminary economics for processing of municipal refuse are estimated.

Process Description

Gasification of solid waste fuels as practiced in the Battelle Northwest studies occurs in a cylindrical, vertical

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In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

retort. Processes that occur in the retort are illustrated in Figure 1.

Waste feed material enters through a lock hopper atop the retort and gravitates through zones of increasing temperature. As it passes through the retort, the waste is first dried and pyrolyzed to yield a char. The char is then gasified and combusted to leave an ash residue. Gases pass countercurrent to the flow of solids. An air-steam blast preheated to about 150°C enters the retort through openings in the grate and passes through an ash zone. The ash zone serves the dual purpose of protecting the grate from the high temperatures of the combustion zone and of distributing the gases over the retort's cross section. Oxygen in the air reacts with carbonaceous residue in the combustion zone generating heat required for the process. The sensible heat in the oxygen-depleted gases from the combustion zone sustains the endothermic gasification reactions of steam and CO_2 with carbonaceous residue from the pyrolysis zone. Gasification reaction rates become negligibly small at temperatures below 700°C. Sensible heat in gases from the gasification zone is used to pyrolyze and dry the incoming waste feed material.

Pyrolysis of the solid waste feed material produces CO, H_2, CO_2, hydrocarbons, and condensibles that are carried in the product gas and carbonaceous residue. The carbonaceous residue reacts with gas phase constituents according to the following reactions:

$$C + O_2 \rightarrow CO_2 \quad (1)$$

$$C + CO_2 \rightarrow 2CO \quad (2)$$

$$C + H_2O \rightarrow CO + H_2 \quad (3)$$

$$C + 2H_2 \rightarrow CH_4 \quad (4)$$

The oxidation of char by reaction (1) is highly exothermic, extremely rapid and proceeds to completion with respect to oxygen disappearance. The endothermic water gas reaction and Boudouard reaction, reactions (2) and (3) respectively, are thermodynamically favored at temperatures over 700°C. These endothermic reactions are slow and rarely at equilibrium in coal char systems at temperatures below 1100°C. Reaction (4) is highly exothermic and thermodynamically favored at temperatures below 600°C. This reaction is of negligible importance in the system discussed in this report.

The following reaction, the water gas shift reaction, is also important in this system:

$$CO + H_2O \rightarrow CO_2 + H_2 \quad (5)$$

This reaction appears to involve a heterogeneous phenomemon on the surface of the fuel.(1) Equilibrium is approached as a function of steam decomposition, although other factors such as fuel reactivity and catalytic activity of the ash constituents should be considered. Detailed discussions of gasification reaction thermodynamics and kinetics are contained in several sources.(1,2,3,4,5)

A hypothetical scheme showing mass and energy flows in the reactor is illustrated in Figure 2. Equilibrium concentration based on reactions (1) through (4) at some temperature may occur in the combustion gasification zone. The compositions and flows shown in Figure 2 are based primarily on yields actually obtained but are not experimental data. The actual product includes a condensible fraction similar to pyroligneous acids. Data are presented in a later section for processing different feedstocks.

The experimental equipment used in the majority of the studies is illustrated in Figure 3. Tars and condensates were usually flared in a close-coupled secondary burner. Some recovery of tars and oils was effected in a water cooled condenser. Product gases were sampled and analyzed to determine compositions.

Description of Testing Program

Studies were conducted in pilot plant facilities that include a three-foot diameter reactor (6) and a one-foot diameter reactor (7). The three-foot diameter unit was a simple, verticle shaft with an operating depth of up to 10 ft. The small scale unit with an operating depth of up to 5 ft was equipped with a mechanical agitator for distribution of the charged solid waste fuels. A rotary grate for ash discharge was also provided. A schematic of the one-foot diameter reactor is shown in Figure 4. Photographs of the one-foot diameter reactor and three-foot diameter reactor are shown in Figures 5 and 6 respectively as they exist today.

Feedstocks that have been processed in the BNW facilities include municipal refuse, wood wastes, agricultural residues, shredded tires, and simulated wastes from nuclear processing facilities. Properties of these materials and products obtained from their gasification are presented below.

Shredded municipal refuse obtained from the City of Vancouver, Washington, and wood chips from the Boise-Cascade plant at Wallula, Washington, were gasified in the three-foot diameter reactor. Composition of the municipal refuse is shown in Table I. Ultimate and proximate analyses of this refuse are given in Table II. Density of the shredded refuse ranged from 38 to 50 lb/ft^3 (600 to 800 kg/m^3).

The wood chips (Western soft wood) were mostly 5/16 to 1 inch in size and had a dry calorific value of 8500 Btu/lb. Moisture content of wood chips processed in the three-foot unit ranged from 20 to 45%.

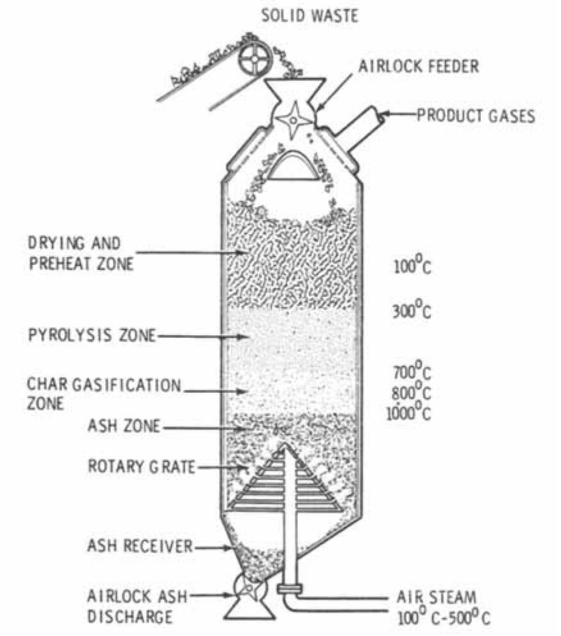

Figure 1. Schematic of Battelle gasification process

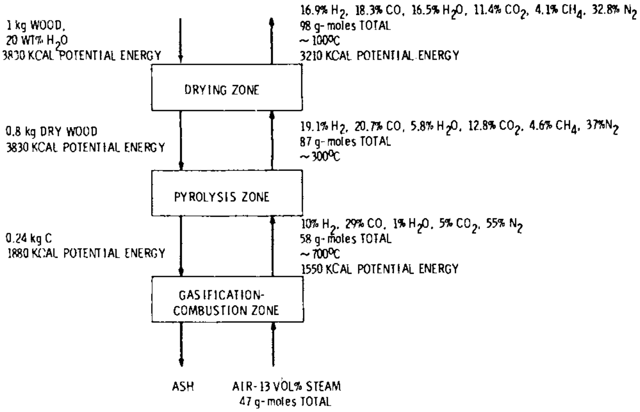

Figure 2. Typical mass and energy flows for wood gasification

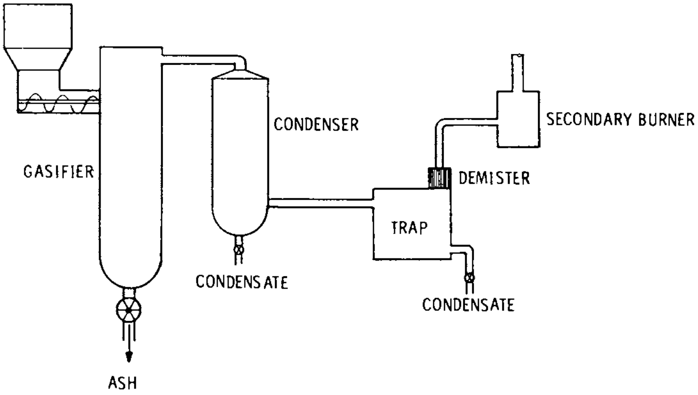

Figure 3. Schematic of experimental equipment

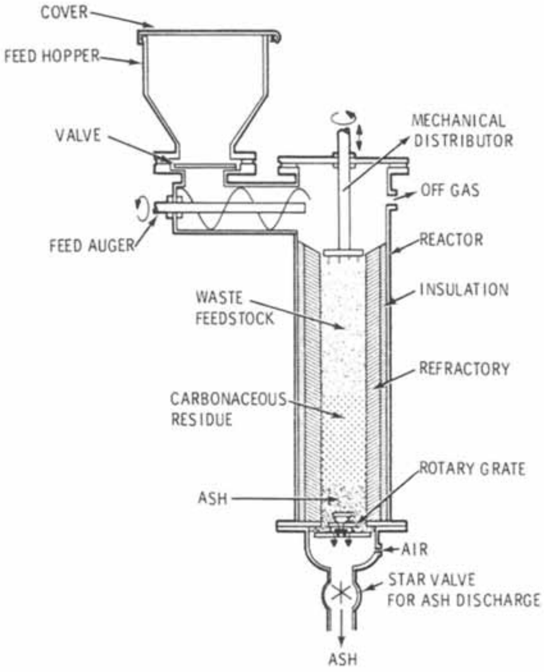

Figure 4. Schematic of small gasifier

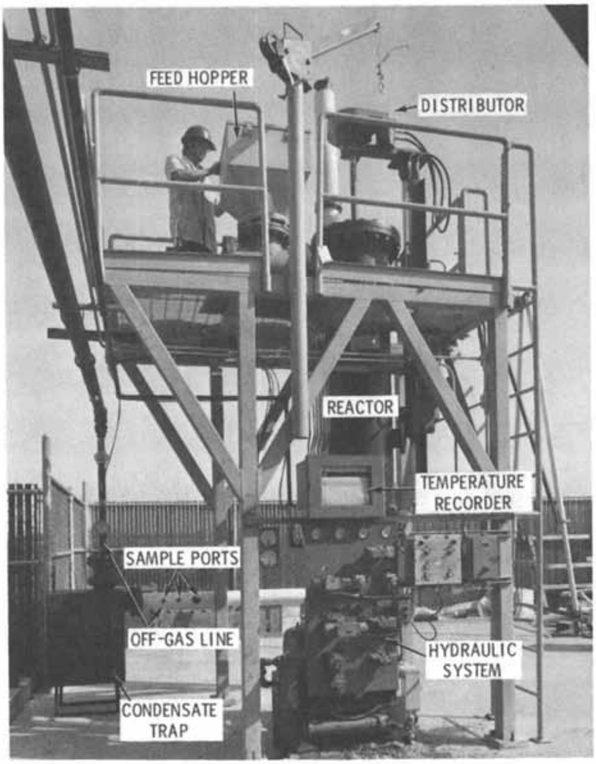

Figure 5. Small gasifier

Figure 6. Three-foot gasifier and hot gas clean up pilot plant

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

TABLE I. Refuse Composition

Component	Weight Percent
Paper	38.5
Fines (Mostly paper and dirt, some glass)	34.2
Rubber and Leather	2.0
Wood	5.3
Metals other than Aluminum	7.0
Aluminum	0.6
Plastic	1.7
Cloth	5.7
Glass	5
Food	Trace

TABLE II. Analysis of Refuse

Ultimate Analysis	% by Weight
Moisture	5.08
Carbon	45.91
Hydrogen	7.06
Oxygen	28.82
Nitrogen	4.19
Sulfur	0.10
Chlorine	0.32
Ash	14.18
Proximate Analysis	
Moisture	5.08
Volatile at 950°C	80.74
Ash	14.18
Btu/lb	7203

Gasification of the wood chips proceeded without difficulty and resulted in a gas product with a heating value of 170 Btu/ft^3, or greater. Processing of municipal refuse was hampered by development of air channels around the bed of refuse in the retort. This is illustrated by the data in Table III. These data were obtained in a run that was initiated by processing wood chips at about 200 lb/hr for 15 hrs and then switching to the municipal waste as feed material at an average rate of 180 lb/hr. An air-steam blast (0.2 lb steam/lb air) preheated to 350°C was used to gasify the waste streams.

As can be seen from examination of the data in Table III, gas quality begins to deteriorate about 5 hrs after switching to the shredded municipal refuse. At this time, none of the wood chips feed remains in the retort. Continued operations with municipal refuse results in further deterioration of gas

TABLE III. Gas Compositions for Run Started with Wood Chips and Completed with Municipal Refuse

Time, Hr	Feed	Gas Composition, vol %								Btu/ft^3
		H_2	CO	CH_4	C_2H_6	C_2H_4	O_2	N_2	CO_2	
5	Wood Chips	22.2	22.0	2.2	0.17	0.15	0.35	41.8	11.8	180
6	Wood Chips	25.2	22.6	2.2	0.16	0.09	0	37.6	37.6	192
10	Wood Chips	23.9	20.2	2.8	0.35	0.27	0	38.5	14.0	191
13	Wood Chips	19.8	21.1	3.0	0.23	0.29	0	41.0	14.6	180
14	Wood Chips	20.3	23.7	2.7	0.25	0.31	0	40.3	12.4	189
14.6	Start municipal refuse feed									
16	Municipal Refuse	21.0	23.5	2.0	0.15	0.13	Trace	42.9	10.3	186
18	Municipal R.	21.6	21.0	1.8	0.15	0.27	Trace	43.3	11.8	172
19	Municipal R.	22.1	20.9	1.6	0.12	0.13	Trace	43.5	11.6	168
20	Municipal R.	17.9	16.4	1.3	0.13	0.35	Trace	50.1	13.8	137
21	Municipal R.	17.8	15.6	1.3	0.12	0.24	Trace	51.1	13.8	135
22	Municipal R.	16.5	15.1	1.5	0.15	0.13	Trace	50.8	15.8	129
27	Municipal R.	11.4	7.3	1.1	0.10	0.15	0.4	61.0	17.8	80
33	Municipal R.	5.1	5.4	1.5	0.18	0.17	0.7	66.5	19.9	58

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

quality. Finally after 33 hours, operations were terminated when the feed rate could not be maintained at 100 lb/hr. Indications of channeling, other than deterioration of gas quality and reduced throughput, include increases in the product gas temperature and in wall temperatures in the upper zone of the retort.

Results of the operations in the three-foot unit made it obvious that non-free flowing feed could not be expected to gravitate through a vertical retort at design conditions. Free flowing feeds such as wood chips were found to be suitable for processing in a vertical retort. Maintaining an open structure in the bed of refuse to allow for gas percolation was found to be essential for satisfactory operation.

Experience with the three-foot diameter unit led to the construction of the unit shown schematically in Figure 4 for processing of simulated wastes from the nuclear industry. Composition of these wastes as simulated is given in Table IV. The high plastics and rubber content in this feed make it particularly troublesome for processing in a fixed bed gasifier.

TABLE IV. Simulated Waste Composition

Component	Weight %
Paper	15
Rags	15
Wood	10
Latex Rubber	5
Hypalon Rubber	10
Neoprene Rubber	10
PVC	20
Polyethylene	15

The simulated waste with the composition shown in Table IV was shredded to less than 4 in. pieces for processing in the small gasifier. Pyrolysis of some of the components in the simulated feed does not yield a carbonaceous residue. For example, polyethylene simply distills when heated and increases the yield of tars from the process. Characteristics of some of the components in the simulated waste from nuclear processing are given in Table V.

TABLE V. Characteristics of Feed Components

Component	wt % Moisture	wt% Ash	Calorific Value Btu/lb	Calorific Value kJ/kg	Decomposition Mode
Paper & Rags	3.4	0.8	7,170	16,700	Endothermic at 320°C
Latex Rubber	---	1.4	18,760	43,600	Exothermic
PVC	---	1.8	8,970	20,900	Endothermic at 260°C
Neoprene	---	28.5	8,690	20,200	Exothermic at 340°C
Polyethylene	---	---	19,790	46,000	Endothermic at 420°C (Melting starts 80°C)

Proper operation of the distributor mechanism permitted processing of the simulated nuclear waste composition shown in Table IV. The distributor mechanism rotated and moved axially to distribute the waste in the refuse bed. Penetration at the center of the bed with an extended prong was found to be beneficial for increasing bed porosity and pushing the refuse outward against the walls of the gasifier. The longest run made with the simulated waste was 50 hours. Gas compositions 24 hours and 48 hours after initiation of this run are shown in Table VI. Average processing rate during this run was 25 lb/hr ft^2.

TABLE VI. Gas Composition from Simulated Waste Run

Time After Startup,	Volume %									
Hour	H_2	CO	C_2H_4	C_2H_6	CH_4	C_3H_8	CO_2	O_2	N_2	Btu/ft^3
24	7.5	21.0	0.4	0.2	1.5	0.1	9.5	0.5	59.3	126
48	8.7	24.8	0.9	0.3	1.9	0.3	8.0	0.4	54.6	162

Yields of tars and oils were high when gasifying the simulated wastes. Condensate collected was about 20% of the weight of the waste gasified. Collection of the condensate in the condenser and tar trap was not complete and tar droplets were carried to the secondary combustion chamber. It is estimated that the actual tar yield is about 30 wt %. These tars had viscosities ranging from 100 cp to 750 cp at 55°C and had a calorific value of 16,900 Btu/lb.

Other feedstocks tested in the small gasifier, Figures 4 and 5, include shredded corn stalks, cubed corn stalks, cubed grass straw, pressed planer shaving, and a mix of pressed planer shavings and shredded rubber tires. Characteristics of these feedstocks are given in Table VII.

Shredded corn stalks could not be satisfactorily processed even with use of the mechanical distributor. Finely divided residue, similar to lampblack, was formed and deposited in the offgas lines. This material had an ash content of about 30 wt% and a calorific value of about 8400 Btu/lb. Satisfactory operation could not be achieved with 100% shredded rubber either. In this case, the rubber melted, hung to the walls, and could not be pushed through the retort. All cubed material, corn stalks, grass straw, and pressed planer shavings gave highly satisfactory operations. A mix of shredded tires (20% by wt) with pressed planer shaving also gave highly satisfactory results.

Gas compositions obtained from processing the feedstocks shown in Table VII are presented in Table VIII. These compositions were obtained about 12 hours after initiation of operations. An air-steam blast (about 0.1 lb steam/lb air) preheated

TABLE VII. Feedstock Characteristics

Feedstock	Calorific Value[(a)] Btu/lb	Calorific Value[(a)] kj/kg	wt % Moisture	wt % Ash[(a)]	Size
Corn Stalks					
Shredded	7085	16,500	---	---	3 in. to 1 in. cuts
Cubed	7085	16,500	7.5	8.8	1 3/4 in. sq, 2 to 6 in. long
Cubed Grass Straw	5980	13,900	25	23[(b)]	1 3/4 in. sq, 2 to 6 in. long
Pressed Planer Shavings	8920	20,700	5.1	0.1	1 1/2 in. right cylinders of pressed wood
Shredded Tires	(c)	---	---	---	2 in. to 6 in. strips of shredded passenger tires

(a) Dry basis
(b) Impregnated with NaOH
(c) Calorimeter tests not conducted

TABLE VIII. Gas Compositions for Different Feedstocks

Feedstock	Volume % H_2	CO	C_2H_4	C_2H_6	CH_4	C_3H_8	CO_2	O_2	N_2	Btu/ft^3
Shredded Corn Stalks	7.0	4.6	0.2	0.1	1.6	0.1	21.3	0.9	64.2	60
Cubed Corn Stalks	16.6	24.5	0.3	0.2	3.1	---	11.4	0.6	43.3	180
Cubed Grass Straw	14.9	21.9	0.3	0.4	2.9	0.1	13.3	0.4	45.8	170
Pressed Planer Shavings	14.4	18.4	0.8	0.4	5.2	---	15.2	0.6	45.0	190
Shredded Tires	5.7	7.0	1.0	0.3	1.9	---	17.0	1.0	66.1	90
20% Shredded Tires/ 80% Pressed Wood	17.9	19.7	0.7	0.3	4.3	0.4	13.5	0.7	42.5	200

to 150°C was used as the gasifying medium.

Yields and processing rates for the different feedstocks are shown in Table IX.

Condensate yields vary with time over the duration of the run. Initially yields are high until the gasifier has reached steady state thermal conditions. Yields also can be expected to vary with the depth of the beds of waste in the gasifier. Operation with a shallow bed depth is preferred to reduce the quantity of condensate generated. Proximity of the high temperature zone results in cracking of the high molecular weight organics. Alkali metal catalysts present with the cubed grass straw were observed to be effective in reducing condensate yields. Alkali metals are reported to be effective gasification catalysts for coal by Cox et al. (8)

The condensate obtained during gasification of pressed planer shavings and cubed agricultural residue was a single phase aqueous solution containing about 20 wt% dissolved organics. The condensate is similar to pyroligneous acids obtained during wood distillation.

An important consideration in determining the suitability of a feedstock for processing in a fixed bed gasifier is the ash properties. The residues from the cubed grass straw was primarily Na_2CO_3 which melts at 850°C. After about 8 hours of operation, the rotary grate siezed up and became inoperable. No ash could be discharged during the run. Examination of the gasifier after the run showed that the residue has frozen into a large clump around the grate impeding its rotation. Residue from other feedstocks did not interfere with operation of the grate. Ash content of the other waste feeds was usually so low that the grate was used infrequently.

Process Economics

Economics were determined for the processing of municipal refuse to generate a low Btu gas for firing a boiler. (6) The flowsheet for the process is shown schematically in Figure 7.

Capital and operating cost estimates for the system shown in Figure 7 are presented in Figure 8 for systems with a capacity of 50 to 1000 tons/day. Costs have been revised to reflect mid year 1977 dollars. If a gas price of $\$1.10/10^6$ Btu could be obtained, plants with a capacity of 400 tons/day would operate at a profit. The costs assume that value obtained for recovered scrap just offsets costs for disposal of ash residue. The operating costs assume sale of municipal bonds with a 25 year payback at 6% interest. Revenue from the low Btu gas assumes that the refuse has a calorific value of 7200 Btu/lb so that 10×10^6 Btu of gas can be obtained from each ton of refuse.

TABLE IX. Yields and Processing Rates Obtained with Different Feedstocks

Feedstock	Conversion Efficiency(a)	Condensate, wt% of Feed	Processing Rate,(d) lb/hr ft^2	Air Required, lb air/lb feed(d)
Cubed Grass Straw	69	--	110	0.9(b)
Cubed Corn Stalks	69	13	58	1.3(b)
Pressed Planer Shavings	70	24	63	1.8(e)
20% Shredded Tires/ 80% Pressed Planer Shavings	60	35	63	1.8(e)
Simulated Nuclear Waste	42	20-35	23	2.3(f)
Shredded Tires	(c)	32	(c)	(c)

(a) Energy in low Btu gas divided by energy input to process

(b) Blast is air-steam with 0.1 lb steam/lb air

(c) Steady operation not achieved

(d) As received wt basis

(e) Blast contained 0.2 lb steam/lb air

(f) No steam in blast

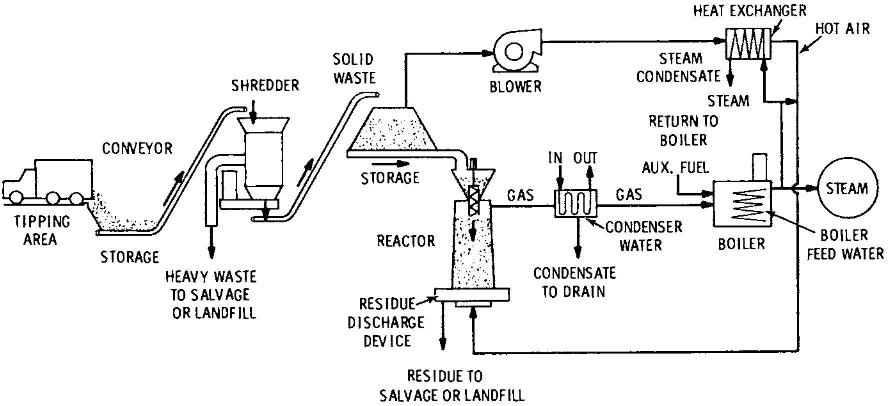

Figure 7. Gasification process to provide fuel for boiler

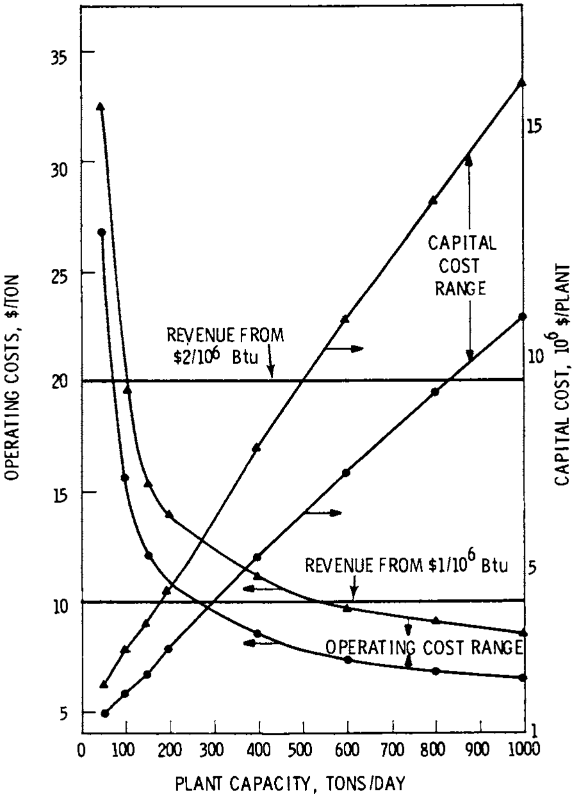

Figure 8. Capital and operating costs for gasification plant to provide boiler fuel

If a market exists for low Btu gas, a region with a population of more than 100,000 people could dispose of their soild waste by gasification at costs less than landfill. Natural gas costs of \$1.50 to \$2.00/10^6 Btu are currently quite common. Similar prices for low Btu gas as a replacement for natural gas appear to be reasonable.

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MARCH 3, 1978.

8

Development of Pilot Plant Gasification Systems for the Conversion of Crop and Wood Residues to Thermal and Electrical Energy

R. O. WILLIAMS[1], J. R. GOSS, J. J. MEHLSCHAU, B. JENKINS, and J. RAMMING
University of California, Davis Agricultural Engineering Department, Davis, CA 95616

A noteworthy observation on the energy crisis, such as it existed in America in 1906, can be found in "A Treatise on Producer-Gas and Gas Producers" by Samuel S. Wyer, M.E.(1). According to Mr. Wyer, "One of the most serious problems of the near future will be that of fuel supply. We are even now returning to mines that were abandoned several years ago and are beginning to rework them with an economy unknown before; and coal seams, formerly regarded as being too thin or too poor to work, are being investigated and purchased in large areas. Brown coal, lignite, peat, sawdust and other refuse are now being used in European Gas Producers, and these low-grade fuels must soon be used extensively in this country."

On the future of natural gas Mr. Wyer stated that "While natural gas is now piped long distances and utilized in manufacturing industries where the use of a gaseous fuel is imperative, yet the states that have gas wells will soon enact laws that will prevent piping the gas outside the state or using it for anything except domestic service." And on the use of gas producers, Mr. Wyer continued, "On account of its high fuel economy, producer gas will be one of the best means of curtailing the useless waste of fuel, utilizing low-grade fuels, and presenting a substitute for natural gas."

This statement appropriately describes the motivation behind the research discussed here. Futhermore, with renewed interest being shown in coal, the price of indigenous natural gas in dispute and general acceptance of residue-derived fuel as a potential contributor to our energy needs, Mr. Wyer's opinions are equally applicable today - 70 years after the publication of his Treatise.

Agriculture and related industries account for 12 to 20 percent of the national energy consumption - depending on the defini-

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[1] Current Address: Biomass Fuel Conversion Associates, Inc., Yuba City, CA

0-8412-0434-9/78/47-076-142$05.25/0

tion of the food chain (2). At the same time, large amounts of energy are available in the form of crop, food processing and forest industry residues (biomass). To facilitate the successful utilization of this energy at the farm, processing plant or lumber mill a simple, low-cost method of recovering thermal and electrical energy from these residues is required.

Downdraft gasification is one such method. This concept has been under investigation at the Agricultural Engineering Department, University of California, Davis for 2 1/2 years. The work is supported by private and government funding. The downdraft producer is a moving, packed bed, vertical-flow reactor employing cocurrent gas flow. A batch-filled, laboratory-scale producer (capacity of 40 to 90 fuel $lb\text{-}hr^{-1}$) has been used to study the gasification characteristics of residues and to test different designs of grate and firebox. The results of this work are available elsewhere (3) (4). A 10 Kw engine-generator set, fueled with producer gas, has been demonstrated, running continuously, for periods up to 4 hours.

Two pilot plant producers systems were designed and constructed to demonstrate gasification in actual application. One unit is supplying 10 million $Btu\text{-}hr^{-1}$ as hot, particulate-free gas, to the burner in a steam boiler, replacing natural gas. A second unit generates fuel gas from corn cobs which is cleaned, cooled and supplied to a 180 Hp dual-fuel, turbo-charged, intercooled diesel engine that is direct-coupled to a 100 Kw generator.

Gasification Reactions

The term "gasification", as used here, is the reaction of solid fuels with air and steam to yield a low-Btu gas which is suitable for use as a source of energy. The principle components of the product gas are formed by a combination of reactions between the carbon in the fuel, oxygen in the air blast, and steam derived from the fuel.

In an ordinary solid fuel combustion process, due to the comparatively shallow layer of incandescent carbon and a relatively high air velocity, carbon dioxide is formed and the greatest number of available heat units is obtained from the combustion of the fuel.

$$C + O_2 + (\text{nitrogen}) \longrightarrow CO_2 + (\text{nitrogen}) + 174{,}000 \text{ Btu} \quad \ldots 1$$

A greater depth of fuel gives rise to the formation of carbon monoxide with corresponding loss of the available sensible heat units. With a sufficient thickness of highly heated carbon, relative to the air velocity, carbon monoxide alone may be the final product, with only traces of carbon dioxide.

$$2C + O_2 + (\text{nitrogen}) \longrightarrow 2CO + (\text{nitrogen}) + 103{,}500 \text{ Btu} \ \ldots 2$$

The action of oxygen on carbon results in the simultaneous formation of carbon dioxide and carbon monoxide but, with a sufficient thickness of fuel, the carbon dioxide reacts with excess carbon producing carbon monoxide (5).

Reactions 1 and 2 are exothermic and the producer temperature will continue to rise for given air input rate until counter-balanced by heat losses in the gas, solid residue, undecomposed steam, and by radiation. In practice a limit is reached by the liability of the solid residue to form clinker and by the maximum temperature which the producer walls can withstand.

Biomass fuels contain substantial quantities of moisture and elemental oxygen. Both the cellulose and the lignin that make up the majority of the biomass fuels contain in their structure hydrogen and hydroxyl (OH) groups. Under the conditions in the producer, as the fuel is fed down into the reaction zone, when the temperature reaches about 150°F (65°C) free water or moisture is rapidly expelled into the product gas stream. Closer to the reaction zone, at about 450°F (232°C), the chemically combined oxygen, hydroxyl groups, and the hydrogen react, producing water. This reation is exothermic; however, the full heat value of stochiometric combustion of hydrogen is not available as much of the hydrogen is combined into the hydroxyl group.

In a downdraft system air is forced radially into a packed bed of fuel through tuyeres which are located near the top of the firebox. Producer gas exits the bed through the grate below the firebox.

The approximate locations of the reaction zones are shown in Figure 1. Neither the location of the various zones nor the temperatures shown are rigidly defined. The diagram should only be used as a guide to the reactions in a producer.

Chemically produced steam along with the steam from free moisture moves down thru the hot fuel bed. There are traces of steam introduced as water vapor in the air blast. The steam reacts with the incandescent carbon according to the following equations:

$$C + H_2O \longrightarrow CO + H_2 - 71{,}200 \text{ Btu} \qquad \ldots 3$$

$$C + 2H_2O \longrightarrow CO_2 + 2H_2 - 76{,}600 \text{ Btu} \qquad \ldots 4$$

Reaction 3 requires temperatures of at least 1650°F (900°C); reaction 4 predominates at lower temperatures (930°F) (500°C).

By the simultaneous action of air and fuel-derived steam, a thermal balance is set up between the exothermic air-carbon reaction and the endothermic steam-carbon reaction. Depending on the relative proportions of air and steam, a constant temperature is maintained and, corresponding with this temperature, a definite composition of the mixed gas is obtained which, in turn depends on the relative parts played by reactions 2, 3 and 4.

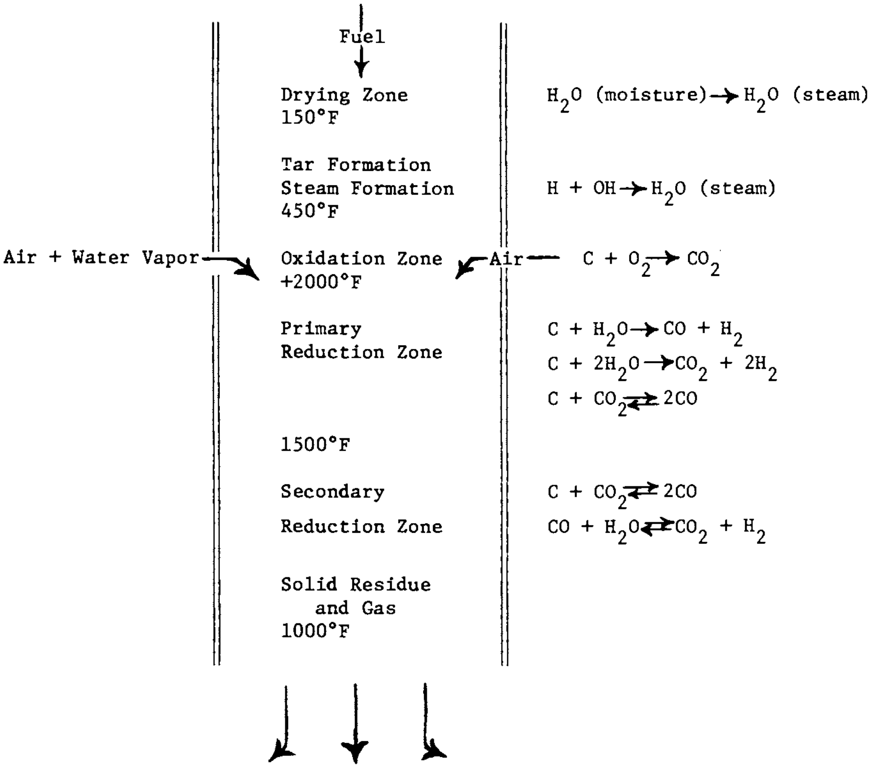

Figure 1. Reaction zones in a downdraft producer

The endothermic nature of reactions 3 and 4 constitute another limit on the maximum obtainable producer temperature. Furthermore, the presence of too much steam chills the combustion zone to a point where carbon dioxide increases, carbon monoxide decreases, and hydrogen increases to a peak value and then decreases.

The volatile matter in the fuel is decomposed during preheating; producing methane, higher hydrocarbons, pyroligneous acids, and other related compounds. Biomass fuels consist mostly of cellulose and lignin and the thermal decompostion of these compounds is very complicated. Large volumes of higher hydrocarbons of tars are produced. As in the case of steam, in a downdraft system the tars are exposed to incandescent carbon before they exit from the producer. A significant portion of the tar is broken down into gaseous hydrocarbons in these hot zones.

A possible analysis of the gas obtained from a downdraft residue-fueled producer would be 10% CO_2, 20% CO, 5% C_xHy, 18% H_2, 1% O_2, and 46% N_2. Carbon dioxide and oxygen are present due to reactions not being completed during the process, and nitrogen is introduced in the combustion air. The gas may also contain from 10 to 20 percent by volume of undecomposed steam, depending on the moisture content of the fuel.

Hydrocarbons are present in the gas over a wide range of homologues. The 5% C_xH_y included in the analysis above consists of methane, ethane, and other permanent gases. Some higher hydrocarbons or tars escape into the gas stream where they form heavy vapors and aerosols. Their dew point covers a wide temperature range. Although tars add to the heating value of the produced gas, their presence is undesirable since they deposit on cold surfaces in piping, gas clean-up and combustion systems.

The following additional reactions remain to be considered.

In the air-carbon reaction, if all the oxygen has combined with carbon, hot carbon, carbon dioxide and carbon monoxide exist simultaneously. Their proportions are controlled by the following reversible reaction:

$$C + CO_2 \rightleftharpoons 2CO \qquad ...5$$

High temperatures and maximum contact time favor the formation of carbon monoxide.

Since hydrogen and steam are also present, the following reaction occurs:

$$CO + H_2O \rightleftharpoons CO_2 + H_2 \qquad ...6$$

This is considered to be a fuel surface phenomenon, with very little reaction in the gas phase. It approaches equilibrium generally as a function of steam decompostion and fuel bed thickness at a given temperature although fuel reactivity and catalytic activity of the fuel ash also play a part (6).

Gasifier Design, Specific Rate of Gasification and Process Variables

A section of the pilot plant gas producer is shown in Figure 2. Excluding the fuel feed and ash removal systems, its net weight is about 4 tons and its overall height is 17 feet. Air is distributed to the fixed fuel bed with an internal cylindrical plenum and eight nozzles or tuyeres. The plenum consists of two concentric steel cylinders, the inside cylinder being the firebox wall. Gas exits through the base and sides of the grate. Designs are discussed later. A variable head of fuel is maintained in the fuel hopper. It passes down into the reaction zones in the firebox under the influence of gravity.

The rate of burning fuel or "specific rate of gasification" is expressed in pounds of dry fuel per square foot and hour, lb $(ft^2\text{-}hr)^{-1}$. When considering the installation of, or the working results of a gas producer plant, specific rate has an important bearing on the technical and commercial success of the plant in question. For a given thermal efficiency, specific rate can be discussed in terms of other process design and operating variables. Specific rate is directly affected by both fuel consumption rate and air input rate, since both of these operating variables are stoichiometrically related. Fuel consumption rate is mechanically controlled by movement of the grate. Grate action controls solids-flow through the vertical reactor and specific rate can be increased until this control is lost.

Grate configuration is a significant design variable. In addition to supporting the fuel bed and enabling solid refuse to be continuously removed from the bottom of the bed, the grate must also present sufficient open area for the gas to escape through and, when necessary, provide suitable bed-agitation. If solid refuse removal is too rapid, the reaction zones are subjected to undue disturbance. Raw fuel replaces char in the reduction zone producing large volumes of tar in the gas and char streams.

The rotating eccentric grate shown in Figure 3, which has seen widespread use in updraft producers, has been used successfully with many fuels. It consists of flat, circular, steel-plate rings mounted one above the other with their edges overlapping. Solid refuse must pass horizontally between the plates and this motion is imparted to the refuse by rotation of the grate. The grate plates are eccentric with the center support so rotation forces refuse between them and also imparts a grinding action to any clinkers which may form. These are crushed between the edges of the plates and the sides of the firebox.

Another grate design shown in Figure 2 involves the use of short, horizontal lengths of stainless-steel tube-augers mounted inside a truncated conical basket which extends below the firebox. The auger flights are mounted in opposing sets on each of three shafts and discharge on opposite sides of the basket where solid refuse drops down to the ash pit. The auger shafts extend outside the producer through packing glands, so the drive mechanism

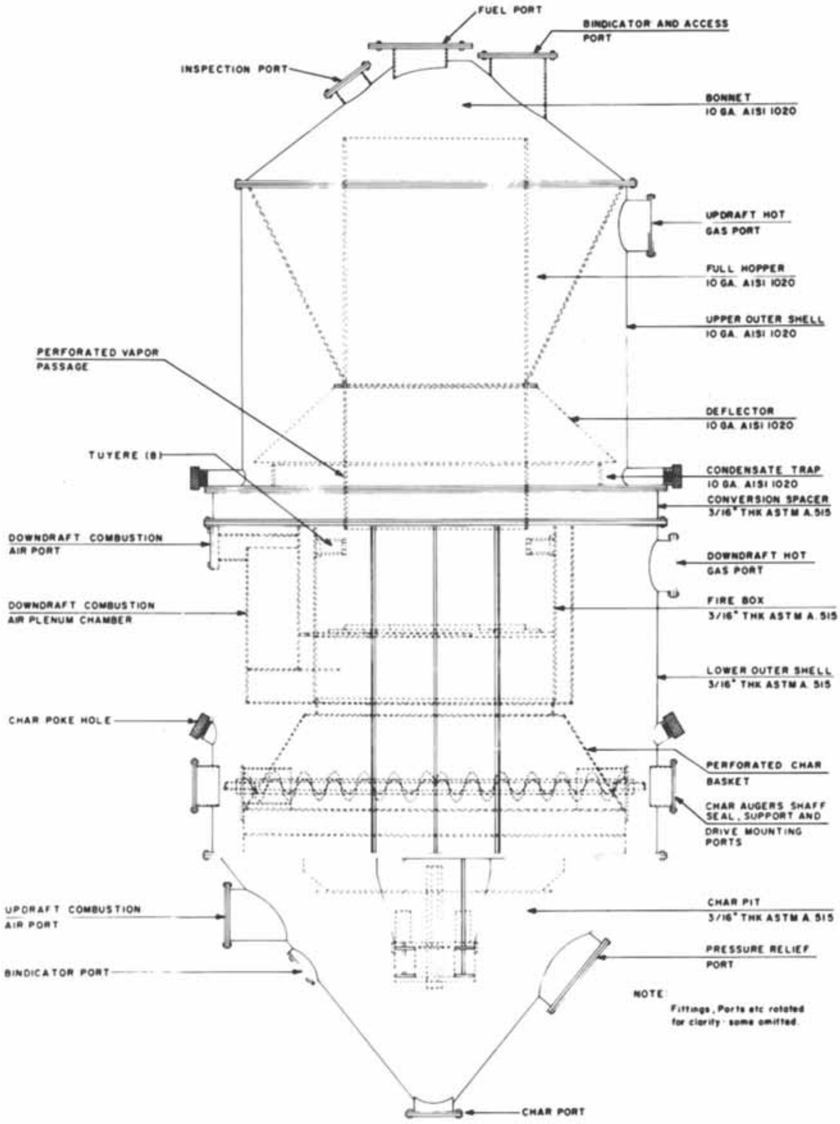

Figure 2. Cross section of downdraft pilot plant gas producer—January, 1978

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

is in a clean atmosphere.

Specific rate is limited by an acceptable air input rate and consequent gas flow rate. If the velocity of the produced gas, immediately below the grate, is greater than the setting velocity of the solid refuse, particles are entrained, resulting in an overload on the downstream gas clean-up system. If the air input rate is too rapid, sufficient residence time may not be available to the gas in the reduction zone, resulting in incomplete formation of CO. In practice, however, this is rarely a limiting factor. The firebox should be designed to give maximum fuel depth to provide for adequate residence time.

Specific rate can be limited by the physical and chemical properties of the feedstock. Moisture and total ash content can limit feedstock from being gasified without suitable preprocessing, but the effect of these two parameters on specific rate is not as pronounced as that of fuel particle size and fusion properties.

Fuel particle size and distribution determine the pressure differential across the bed for a fixed rate of air input. Fine materials, such as sawdust and rice husks, present a high resistance to the passage of air and gas due to the low voidance of fixed beds composed of these materials. A Δp of 5" w.g. was recorded over a bed of sawdust one foot deep for a superficial velocity of 110 ft-min^{-1}. With similar beds of broken corn cobs, wood chips, and bark, a Δp of 0.5" w.g. was obtained for the same air velocity. Fine feedstocks must be densified (cubed or pelletized), to increase individual particle size, before they can be gasified in a fixed bed downdraft system. The high pressures required to force the combustion air into the bed, tend to blow material through the grate.

Densification also improves the material handling properties of feedstocks such as cotton gin waste and cereal straw. Furthermore, without densification these materials bridge in the fuel hopper section of the gasifier.

Fusion temperature of the inorganic matter (ash) in the feedstock determines the formation of clinker from the ash in the oxidation zone (+2000°F). The ash components form eutectic mixtures of lowest melting point which melt out from the whole mass. Cotton ginning waste is particularly prone to clinker formation and must be cleaned to remove as much fine inorganic matter as possible prior to gasification. Careful temperature control in the reactor is essential with clinker-prone feedstocks.

Total ash content in the feedstock is important, since it is usually present at the expense of carbon. Most biomass fuels have ash contents within the range of 1 to 5 percent by weight. Rice husks, with up to 21% ash, mostly silica, are an exception. Gasification of rice husks results in unavoidably low thermal efficiency due to sensible heat losses in the solid refuse.

Water, from moisture in the fuel and water of reaction, is converted to steam which, in a downdraft reactor, must then pass through the reaction zones. Too much steam, from very wet feedstocks, chills the reaction zone, reducing gas quality and thermal efficiency. Wet fuel (>30% by weight moisture) must be dried. Where the feedstocks also requires densification, these two requirements can be met in one operation.

Temperature resistance properties of the firebox wall materials would be expected to limit the specific rate. In the downdraft producer, a layer of char builds up around the inside of the firebox, insulating the wall material, tuyeres, and choke plate from the high temperatures in the oxidation zone. Formation of this layer is not affected by specific rate.

Ancillary Systems

In addition to the gasifier, a gas producer system includes a fuel feed system with a storage bin, solid refuse removal devices and disposal means and a suitable gas clean-up train. Where it is desired to replace natural gas for heated air drying, a producer gas burner must be furnished and where the desired end product is electrical power, an engine-generator (or turbine-generator) has to be provided.

Fuel feed and ash removal have to be accomplished through air locks since the producer is operated either at a low pressure (1 to 5 psig) or under a slight vacuum. For fuel feed, a rotary valve mounted below either a knife gate or butterfly valve (to provide positive seal when the system is not cycling) has been found to give satisfactory performance. Fuel is conveyed from the storage bin to the top of the feed valve using tube augers for materials such as corn cobs, or a belt conveyor for wood chips. A rotary valve can be used to discharge solid refuse from the base of the gasifier. Level indicators must be provided to control the operation of the fuel feed and solid refuse discharge. Totally automatic operation of both systems is easily achieved.

Coarse particulates (solid refuse carry-over) should be removed from the produced raw gas as close as possible to the gasifier. Cyclones are well suited to this task. Accurate cyclone design is made difficult by a fluctuating raw gas flow so high efficiencies are not achieved. The cyclone should be designed to accommodate maximum flow; the objective should be to relieve the load on down-stream clean-up components. A conventional baghouse, employing fiberglass bags has been used to remove the remaining fine particulates. A cluster of fiberglass bags is shown in Figure 4. To avoid the deposition of either tar or water in the cyclone and baghouse, the gas must be maintained above the dew point (about 225°F) of the condensable vapors. Heat losses should be minimized by insulating the cyclone, baghouse, and connecting pipework (Figure 5).

After removal of coarse particulates, the gas is suitable for combustion in a steam boiler. A diverting valve should be provided to route the gas to a flare either in an emergency shut-down or when low fire is called for in the boiler combustion chamber.

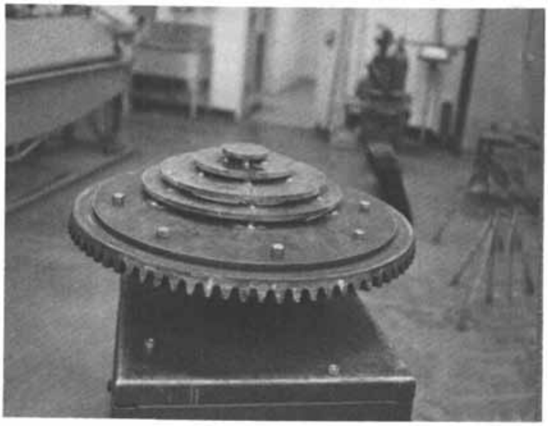

Figure 3. Rotating eccentric grate

Figure 4. Fiberglass filters

A center-fire type gun is generally proposed for use with producer gas in a boiler firebox. It should have a large gas orifice as insurance against tar build-up and high pressure drop.

If the gas is to be used as fuel in an internal combustion engine, it requires further cleaning and also cooling. A finned-tube heat exchanger has been used to remove condensables from the gas and to increase its density. A fan is used to cool the tubes with ambient air. Although direct contact between the gas and a cooling fluid (generally water) is a more efficient method of cooling, disposal of the cooling water is usually an environmental problem.

A constant speed, positive displacement blower was placed in the system between the gas clean-up equipment and the diesel-engine gas throttle valve. The blower is rated at 300 cfm and with a set of valves and pressure regulators performs two functions. During producer start-up and flare operation, the blower supplies pressurized air to the producer. In this mode the producer and cyclone are pressureized to around 4 psig. When producer gas is to be supplied to the diesel engine, the blower is valved such that below atmospheric pressure exists in the producer and gas clean-up system. Air is drawn into the producer from ports valved to the atmosphere. The blower draws producer gas from the producer, through the gas clean-up system, and provides a pressurized, regulated gas supply to the throttle valve on the diesel engine air intake stack.

A complete trailer-mounted energy conversion plant, comprising gasifier fuel bin, gasifier, clean-up train, and 100 KVA engine-generator is shown in Figures 5 and 6. A schematic of the process is shown in Figure 7.

Dual Fueling and Performance of a Standard Turbo-charged Inter-Cooled Diesel Engine

A diesel engine with a 100 kw 12 wire alternator mounted on the trailer was used to supply the necessary electrical power to the onboard equipment and to generate electricity using producer gas as the main source of energy. The engine is a 6 cylinder, 4 stroke, direct injection diesel engine with 531 in^3 displacement and a compression ratio of 14.5 to 1. It is turbo-charged, intercooled, has a power rating of 202 bhp intermittent, 172 bhp continuous at 1800 rpm, and is equipped with a Bosch inline injection pump and Bosch variable speed governor (type RSV).

Considerations for Dual Fuel Conversion. In order that diesel engines can operate successfully on producer gas, the following characteristics are considered important. A minimum of diesel fuel should be used to maintain ignition of the producer gas. This will indicate the ability of a diesel-producer gas system to economically operate in a situation where abundant cellulosic matter is present for gasification but diesel fuel is in short supply

Figure 5. Gas producer system showing insulated cyclone and filters

Figure 6. Complete energy conversion system (100 KVA engine–generator is in foreground)

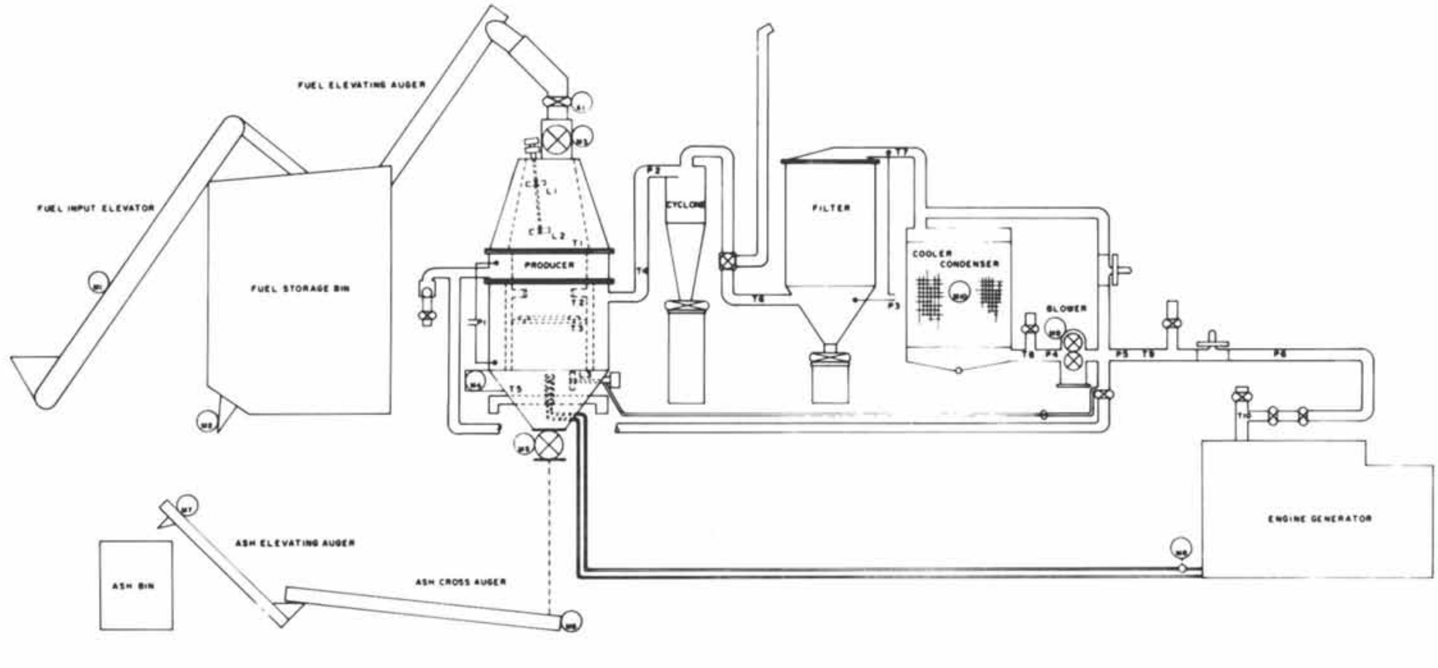

Figure 7. Schematic of portable 100 w farm power plant

and/or costly.

It is desirable to retain the accuracy and fast response to load changes that are achieved with diesel engine governors on new engines. This will allow diesel-producer gas generator sets to supply regulated electrical power if required. Present day dual-fuel engines normally are supplied with a pressurized gas supply. Gas input response is dependent solely on the response of the gas throttle valve to governor signals. With the gas coming from a local gas producer, gas input response will be dependent on the operating characteristics of the producer and supporting equipment upstream of the engine. Poor response will result in large speed variations during changes in engine load. Therefore, it is unacceptable to fix the diesel injection quantity and allow the engine governor to only operate the gas throttle valve, as is currently done with many dual-fuel systems.

Safety considerations are of prime importance with respect to both the personnel operating the equipment and to the equipment itself. Normal safety features are included in the engine operating system. The main safety problem resulting from using producer gas to dual-fuel the engine, is one of an explosion hazard within the exhaust system if unburned producer gas is exhausted from the combustion chamber. The solution is to ensure that complete combustion occurs under all operation conditions and that the producer gas is purged from the intake and exhaust system, prior to engine shut-down.

Modifications and Additions to Standard Diesel-Engine. The only alterations made to the production model diesel engine were to the Bosch governor and were minor in nature. The governor housing was modified to allow the addition of an adjustable, minimum stop for the diesel fuel rack, and a sensor which indicated the absolute position of the fuel rack. The fuel rack stop dictated the minimum quantity of fuel that was injected from the fuel pump. This fuel rack stop is solenoid controlled and when disengaged, the operation of the governor and fuel pump to engine speed variations is no different from the unmodified diesel engine. The fuel stop is set at the minimum quantity of diesel fuel required to maintain continuous ignition of the producer gas. This quantity is about half that consumed at no load idle (1800 rpm) on full diesel. The stop ensures that complete combustion will be maintained under all operating conditions, thus, reducing the explosion hazard.

The fuel rack position sensor has a resolution of .001 inch and is used in the gas throttle valve control circuit. The sensor is a wire-wound linear motion potentiometer and has little effect on the governor's performance.

The fact that no other alterations to the diesel engine were required is a good indication that minimal capital will be needed to develop successful producer gas/dual-fuel diesel engines manufactured under mass production techniques. Equipment added to the diesel engine consisted of a gas throttle valve, its actuator and electronic control and logic circuit and the following safety equipment: 1) emergency intake air shut-off valve, 2) over-speed sensors, 3) emergency kill button, and 4) producer gas snap-off valve upstream of the throttle valve (low oil pressure and high water temperature sensors were standard equipment).

Engine Performance Results of Dual Fuel Operation on Producer Gas. After the gas producer system had been brought up to operating temperature, the diesel engine was supplied with producer gas generated from broken corn cobs for dual fuel running. Producer gas composition was assumed to be 20.1% CO, 16.3% H_2, 5.0% THC with a heat content of 162 Btu/ft^3 at STP. These average values are from gas analyses data obtained with dry broken corn cobs at another location. Resistive heater panels provided an electric load up to 100 Kw (150 Bhp) in 6 Kw increments. For most of the testing the quantity of diesel fuel injected for ignition purposes was maintained at that quantity required for 1800 rpm, no load idle. Load was added or subtracted in 6 Kw increments and the gas throttle valve manually adjusted to maintain the idle quantity of diesel fuel. Zero to 60 Kw load was delivered by producer gas with no problems being experienced. In fact less exhaust smoke and quieter running were experienced when producer gas was being used. This is attributed to more complete combustion of the outer surface of the injection spray and to the slower combustion characteristics of the low-Btu gas. Producer gas could not pick up more than 60 Kw of load without detonation appearing. By increasing the diesel fuel being injected, alternator output could be increased beyond 60 Kw. One hundred Kw output was easily achieved with 60 Kw being supplied from producer gas.

This demonstrated that air had not become a limiting factor to increasing the load picked up with producer gas. The effect of low amounts of injected diesel fuel was investigated. The diesel engine has an indicated demand of 40 hp at 1800 rpm no load idle. The alternator output was set at 46 Kw (70 Bhp) and the gas throttle valve slowly opened. As producer gas flow rate increased, the governor reduced the quantity of diesel fuel being injected. As the position of the minimum fuel rack setting was approached, detonation appeared, and about 60 Kw of energy was being supplied by producer gas. Up to this point it has been the quantity of producer gas that caused detonation, because enough diesel was being injected to ensure consistent ignition. This procedure was repeated with less than 40 Kw load on the alternator. The fuel rack stop was disengaged and the injection quantity of diesel fuel was allowed to pass below this setting. Just below the minimum setting severe misfiring began to occur. Approximately 6 1/2 lb/hr of deisel fuel was being injected when misfire occurred. Producer gas quantity was then varied between conditions of misfire and detonation over all 6 Kw incremental settings of alternator load. The engine operating performance was observed with no increase in the exhaust smoke noted at anytime.

It was observed that the only factor contributing to detonation was the quantity of producer gas being used in the engine. Whenever the producer gas contributed more than 60 Kw of load, detonation would appear. Except for the quiet running of the engine when burning low-Btu gas, it was impossible to tell that it was running in a dual fuel mode. Ratios of gas to diesel energy up to 4.5 to 1 were experienced. Detonation or misfire prevented higher ratios from occuring.

The engine intercooler was supplied with engine coolant and resulted in an induction mixture temperature of about 170°F. Detonation occurred mainly because of the hydrogen content of the producer gas and the relatively high induction mixture temperature. Injection pump timing was not set at optimum for dual fuel operation and could have added possibly 5 Kw prior to onset of detonation. The use of tap water in the intercooler would lower the tendency of the producer gas to detonate. Then possibly 100 Kw of the load could be supplied by producer gas.

Calculations indicate that sufficient combustion air is available for full load operation. Unless detonation appears prior to 100 Kw loads, the diesel engine should not be derated for 100 Kw (150 Bhp) for producer gas operation. Thus, clean cool gas is necessary for reliable continuous dual-fuel operation. Although in these tests detonation limited producer gas output to 60 Kw (90 Bhp), minor modifications will allow the producer gas power to increase and likely to the full rated output of 100 Kw output for the alternator.

Pilot Plant Gasifier Performance. The pilot plant produced low-Btu gas from kiln dry hog wood chips to fuel a low-pressure (10 psi) natural-draft boiler of about 151 boiler horsepower capacity (4370 lb steam/hr @ 1157 Btu/lb enthalpy). The boiler steam was used primarily for space heating in a single story concrete construction building housing the State of California printing facility. The wood chip fuel had a calorific heating value of 8814 Btu/lb dry fuel basis and contained about 13.7 percent moisture on a wet basis. With nighttime ambient temperatures between 52 and 56°F, a gasification rate of 1105 lb dry wood chips/hour was adequate to meet the boiler heating load. The average combustible gas contents by volume at ambient temperature and pressure were: CO - 23.4%, H_2 - 13.2%, and THC (total hydro carbon) - 9.2%. The net total energy per cubic foot of low-Btu gas for these volume percentages was 205.7 Btu based on Btu/ft^3 heating values of 321.8 for CO, 275.0 for H_2, and 985.9 for THC (90% CH_4 @ 913.1 and 10% C_2H_6 @ 1641.0). Assuming an 85% fuel thermal efficiency for the gas producer, 8.278 million Btu of low-Btu gas was produced hourly with an average off-gas temperature of 780°F. Excess gas not required by the boiler was burned at flares installed in the heat gas line.

After steady-state gas production was achieved, there was negligible condensate collected (less than 1 qt/hr). The average dry char productive rate was 8.27 percent of the dry fuel input rate. For the 4-foot diameter pilot plant producer firebox, the specific dry fuel rate was 87.9 lb/ft^2 hour. Proximate analyses of char from the same fuel at a considerably higher moisture content but gasified at a different time contained 5.0% volatile combustible matter, 92.28% fixed carbon and 2.70% ash (% are by weight dry basis of the char).

Cost Projections

Cost projections for gasifier applications are made difficult by their site-specific nature. Projected fuel costs, fuel transportation, and residue availability are particularly difficult to predict. In the following hypothetical situation a suitable residue with a heating value of 7,500 Btu-lb^{-1} (moisture content of 10%) is assumed to be available at a cost of $15 per ton delivered to the producer site. The following example describes a water pumping application.

In some sections of the U.S. the cost and availability of energy for pumping irrigation water has increased to seriously limit the profitability of intensive cropping. In parts of the world experts are now recommending a draw-down of irrigation wells of as much as 98 feet (30 meters) to obtain the water required for crops to support the local population (7). Every year, in the Central Valley of California and in other western areas that use underground water for irrigation, the water table is lower and the required pumping lift that much more to provide the water.

For the purpose of this study, a Central Valley ranch has been selected not as a statistical average but as an illustrative example. The ranch has 900 acres in cultivation with crops that need 30 inches of water per year of soil moisture which is about 42 inches of irrigation water, allowing for losses and inefficiencies. The irrigation water comes from four wells and the depth to water table requires a total dynamic lift by the pumps of 120 feet. The pumps work 200 days per year, 12 hours per day, to pump the required irrigation water of 3150 acre-feet.

The five year average cost of natural gas has been projected in the following manner:

Present cost, per million Btu (10 therms)	$2.20
1978 season cost, with indicated 30 percent increase	2.86
1979 season cost, estimated 15 percent increase	3.29
1980 season cost, estimated 15 percent increase	3.78
1981 season cost, estimated 10 percent increase	4.16
1982 season cost, estimated 10 percent increase	4.58

Average cost, five years, of $3.73 per million Btu

Quantity of water required per day and well:

$$\frac{\text{900 acres x 3.5 ft. of water}}{\text{200 days x 4 wells}} = 3.937 \text{ acre feet}$$

Volumetric demand on each well:

$$\frac{3.937 \text{ acre feet} \times 3.259 \times 10^{5} \text{ gallons per acre ft.}}{12 \times 60}$$

$$= 1782 \text{ gpm/well}$$

Horsepower requirements of each well:

$$\frac{1782 \text{ gpm} \times 8.34 \text{ lbs/gal.} \times 120 \text{ ft. of lift}}{33{,}000} = 54 \text{ hp}$$

The thermal efficiencies of the pump and engine are assumed to be 0.5 and 0.2 respectively.

Required energy input to each installation:

$$\frac{54 \text{ hp} \times 42.42 \text{ Btu/min per hp}}{0.5 \times 0.2} = 22{,}907 \text{ Btu/min}$$

It is assumed that a natural gas engine has been available to drive the pump. The natural gas will be replaced by residue-derived producer gas.

Daily natural gas requirements of engine:

$$22{,}907 \text{ Btu/min} \times 60 \times 12 = 16.49 \times 10^{6} \text{ Btu/day}$$

Natural gas costs per day for each well:

$$16.49 \times 10^{6} \text{ Btu/day} \times \$3.73/\text{million Btu} = \$61.5$$

Quantity of residue required to fuel gasifier:

$$\frac{22{,}907 \text{ Btu/min} \times 60}{0.65 \times 7500} = 282 \text{ lb/hour}$$

(The thermal efficiency of the gasifier is assumed to be 0.65).
Hourly cost of residue fuel:

$$\frac{282 \times 15}{2000} = \$2.11/\text{hour}$$

A gas producer with a 24 inch diameter firebox and grate is required to gasify the residue at a fuel rate of 282 lb/hour.

The specific rate of this producer is 90 lb/ft^2 hour which is an acceptable design for a downdraft producer.

The purchase price (PV) of a 24 inch producer system is assumed to be \$25,000. This includes the fuel feed, solid refuse removal and gas clean-up equipment.

The PV funds are assumed to be available as proceeds from a self-amortising loan payable in monthly increments over 5 years at

9% rate of interest.

The monthly payments are calculated to be $518.96. Total cost of producer:

$$\$418.96/\text{mo.} \times 60 = \$31,137.60$$

Hourly cost of producer:

$$\frac{31,137.60}{5 \times 200 \times 12} = \$2.59/\text{hour}$$

Total operating costs of the gasifier system:

$$\$2.59 + 2.11 = \$4.7/\text{hour}$$

Savings: $\$\,\frac{61.5}{12} - \$4.7 = \$0.42$ per operating hour

$$\$0.42 \times 12 \times 200 \times 4 \text{ (wells)} = \$4,032 \text{ per year.}$$

The following example considers the use of producer gas as replacement for natural gas at a hypothetical rice mill.

The gasifier has to supply the following loads:

24 hrs/day and 360 days per year
One 200 Bhp Steam Boiler
Two 250 Bhp Steam Boiler
One 350 Bhp Steam Boiler
Boiler usage = 0.65 (65%)
Boiler efficiency = 0.75

Gas is to be generated from rice hulls. These are assumed to have a heating value of 7,200 Btu/lb. A gasifier thermal efficiency of 0.65 is also assumed.

To meet peak demand of the boilers the following fuel consumption rate is required:

$$\frac{1050 \text{ Bhp} \times 3.352 \times 10^{4} \text{ Btu/hr/Bhp}}{0.75 \times 0.65 \times 7,200 \text{ Btu/lb}} = 10,027 \text{ lb/hr}$$

Five gasifiers, each with 60 inch firebox and grate are required. Specific rate of gasification:

$$\frac{10,027}{5 \times 6.25\pi} = 100 \text{ lb. per ft}^{2} \text{ hour}$$

Cost of gas producer system:

Five 60 inch downdraft gasifiers with controls and hot cyclones	$500,000
Feedstock preparation (2 pelletizers with material handling equipment)	125,000
Feedstock bunkerage	50,000
Modifications to existing burners and controls	40,000

Total equipment costs	715,000
Installation (30% of equipment cost)	214,500
Total capitalized cost	929,500

Projected Costs of Natural Gas. The current cost of natural gas to the rice mill is \$2.68 per million Btu. This is increasing at a rate of \$0.03 per month.

Since the gasification system is to be paid for over a 10 year period, the predicted cost five years from now will be used.

Total increase in 5 years = 5 x 12 x 0.03 = \$1.80
Predicted cost in 5 year time = \$4.48 per 10^6 Btu

Natural gas consumption in terms of Btu:

$$\frac{1050 \times 3.352 \times 10^4}{0.75} \text{ Btu/hr}$$

At 65% boiler usage, annual energy consumption:

$$\frac{1050 \times 3.352 \times 10^4}{0.75} \times 24 \times 360 \times 0.65 = 2.6 \times 10^{11} \text{ Btu/yr.}$$

Annual cost of natural gas:

$$\frac{2.6 \times 10^{11} \times 4.48}{10^6} = \$1{,}180{,}700 \text{ per year}$$

Cost of capital:

9.75% self-amortizing 10 year loan, monthly payments, expressed as an annual cost to the producer.

Monthly payments: \$12,155.09
Annual cost of gasifier: \$145,861

Other Operating Costs:	\$ Per year
Two men to operate feedstock preparation system, single shift, 5 days (2 x \$17,000)	\$34,000
Three men to operate gasifier, three shift 7 days (equivalent, 3 x \$20,000)	60,000
Maintenance and parts (5% of capital cost)	46,475
Insurance and taxes (5% of capital cost)	46,475
Pelletizing at \$5/ton	216,000
Grinding at \$2/ton	86,400
Gasifier System	145,861
Total operating costs	\$635,211

Savings: \$1,180,700 - 635,211 = \$545,489/yr.

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9

Stagewise Gasification in a Multiple-Hearth Furnace

ROBERT D. SHELTON
BSP/Envirotech, One Davis Drive, Belmont, CA 94002

The thermal reformation of solids is one of the oldest technologies in the world. One of the earliest advanced tools available to people of the world was the reformation of the physical and chemical characteristics of solids via the application of heat. Today, ore roasting, incineration, gasification and carbon production and activation are common examples of thermal reformation processes. Gasification is a thermal reformation technology with one of the most prominent historical roles. The thermal gasification of wood was known to the Pharaohs of Egypt and it produced charcoal in Europe through the Middle Ages. It supplied the majority of wood alcohols and acetone for the Western Hemisphere until the early 1900's and it harnessed the energy of coal, wood and peat for industry and transporation in Europe from the 1930's until the 1950's. Today, gasification is emerging as a prominent technology for the conversion of carbonaceous wastes to energy.

There are a huge assortment of carbonaceous solids in todays world, by-products of people and industries, which are classically identified as wastes. Municipal sewage sludge, garbage, industrial sludges, agricultural residues and food processing wastes are examples of such solids. In the past, these have generally been disposed of at the lowest possible cost. Today, stringent environmental requirements, increasing disposal costs, and limitations on the availability of natural resources have encouraged the utilization of these carbonaceous solid wastes as renewable resources.

Municipal sewage sludge provides a good example of the changing waste disposal situation in this country. Most communities are faced with the problem of sludge disposal. Sewage sludges are becoming an especially difficult problem because of their escalating quantities and decreasing disposal options. It is estimated that disposal methods will have to deal with 23,000 dry tons of sewage sludge per day by 1985 in the United States. In addition, methods of disposal available today will not be legal or practical in the years ahead. The

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EPA has assured the end of ocean dumping, a technique which presently deals with approximately 15% of the sludge produced and the escalation of fuel costs has made the economic future of sludge incineration untenable for some communities and uncertain for others. Already, sludge incinerators are sitting unused in many communities, even though they are an environmentally acceptable disposal method. Incinerative disposal presently handles 35% of the present sludge outfall, but this fraction will certainly decrease as communities become unable or unwilling to pay the escalating fuel costs.

The landfilling of sludge and the land application of sludge are disposal options that pose an interesting contradiction. The present Resource Conservation and Recovery Act (RCRA) mandate is that sewage sludge, when landfilled, must be treated as a hazardous waste. This is because sewage sludges may contain heavy metals, pathenogens and sometimes pesticides and herbicides. The presence of these elements makes the landfilling of sludges a potential hazard for plants, animals and the land and water quality. According to RCRA, these hazardous properties of sludge make it imperative that the release of the harmful substances into the environment be controlled. The EPA, on the other hand, has endorsed land application of sludges as a soil conditioner and fertilizer. The extent of the danger that this land application of sludges may pose, both in the long and short term, has yet to be determined, but the present EPA recommendations are in stark contract to the RCRA designation of sludge as a hazardous material. Obviously, there are constituents in sludge which could aid in soil conditioning but their benefits will have to be weighed against the potential dangers of allowing significant quantities of the hazardous constitutents into plants and animals, specifically, and the environment, in general.

The disposal and utilization situation with sludge is analogous to the other municipal and industrial solid waste problems of today. The chief considerations in all wastes are the costs associated with discharging or disposing of the solids and the liability that is assumed because of that disposal. The peculiar toxicity of certain industrial sludges is a constant liability for manufacturing and process industries. Although thermal reformation via incineration of these toxic solids reduces the environmental liability, it does so by exchanging liability for high costs. Municipal solid wastes such as garbage are not nearly as toxic as some industrial counterparts but the options for disposal are decreasing while the costs are escalating.

For all solid wastes the problem is one of implementing disposal technologies to deal with the increased quantities of waste in the face of fluctuating disposal options and a general uncertainty as to their value and place in the environment. The greatest potential for the handling of these wastes lies in the capture of their value rather than in their destruction. Across the full range of wastes, from municipal garbage to industrial sludge, efforts are being mounted to recapture some of the inherent worth contained in these solids. Not only does this recapturing provide an increased materials utilization efficiency, it can also result in a nulification of some or all of the detrimental attributes of the waste. The reformation of wastes into usable forms can be a difficult task, though, when faced with present environmental, technical and economic constraints. However, the gasification of these carbonaceous solids into energy has been demonstrated to provide a direct, cost-effective, available, and environmentally sound method of disposal. Gasification provides a technology which transforms the discarded and often dangerous components of society and industry into renewable energy supplies.

The multi-hearth furnace can gasify carbonaceous wastes into utility energy, and the technology is well proven. The development of the multiple hearth gasification system (MHGS) grew out of a long history of solids thermal reformation experience, beginning in 1900. To date the list of its activities include incineration, ore roasting, char production, char activation, and more recently, waste conversion to energy. The design and operation of the multiple hearth furnace has shown it to be capable of handling a wide range of solid feedstocks and for controlling process conditions so that specific product qualities can be attained. The same technological capabilities which make it possible to groom the thermal reformation of wood, char, and activated carbon make it possible to reform the carbonaceous wastes of industry or municipality into usable energy, with a high degree of efficiency and quality control. For the multiple hearth it is a rather short hop from producing charcoal and gas from wood wastes to producing char and fuel gas from sludges and solid wastes.

There are immense compositional ranges of solids which are amenable to thermal reformation. Some of these solids have a composition which can sustain gasification with a net energy gain while others require a fuel input to gasify the solid. Envirotech has experimentally and commercially gasified a wide range of feedstocks and has developed a general correlation between the feedstock properties and the energy which can be derived from a given feedstock composition.

The best characterization of a feedstock's energy potential utilizes the ratio of its calorific value to its water content, expressed as the available BTU per pound of water in the feedstock. The ratio automatically takes into account the inherent feedstock calorific value and the diluting effect which feedstock moisture has on energy production.

Figure 1 depicts the relationship between a feedstock's BTU/lb water ratio and its energy potential. 3500 BTU/lb water is the autothermic point where the inherent calorific value of the solids is just sufficient to offset the water dilution. At lower BTU/lb water ratios the calorific value of the feedstock is insufficient to evaporate the contained water and fuel must be utilized to augment the evaporation and gasification. Above 3500 BTU/lb water is the autothermic region where the calorific content of the feedstock is sufficiently high to evaporate the feedstock moisture and produce energy. As the feedstock's BTU content increases with respect to its moisture content there is more energy available which can be converted to energy.

A typical municipal sewage sludge will contain 20% solids. 65% of these solids are combustibles with a heating value of 10,000 BTU/lb. Figure 2 depicts this typical sludge and shows that it contains only 1625 BTU/lb water, considerably below the autothermic region. Traditionally, sewage sludge has been disposed of via incineration and its compositional BTU/water ratio has required the input of fuel.

In contrast, typical municipal refuse contains over 20,000 BTU/lb water, high enough to generate significant amounts of energy. The problem with municipal refuse is its heterogeneous composition, including large quantities of metals and glass. The composition of a typical refuse is shown in Figure 3. However, with available technology, the combustible fraction of the refuse, refuse derived fuel (RDF), can be separated. Figure 4 indicates the compostion of RDF and shows that it contains 27,200 BTU/lb water. This places RDF well into the autothermic process region and marks it as a good fuel producer.

The RDF and sludge provide an example of the range of waste properties which must be dealt with. In the past, some of the municipal refuse and sludge have been disposed of via separate incineration without any heat recovery. While it is theoretically possible to capture the heat of incineration from these wastes, practical considerations, such as solids clinkering, limit this possibility.

The major restriction on heat recovery from incineration is the low gas temperatures of the process. Incineration is capable of producing such high temperatures (2000°F) that the solids can fuse together. To prohibit this, large amounts of excess air are used to quench the reaction. By comparison gasification of these solids liberates the fuel potential of the feedstock as a gas with solids temperatures at 1500°F or below. The gas is combusted in a separate chamber from the solids. This allows for high gas temperatures and high heat recovery efficiencies without solids agglomeration problems (see Figure 5).

Figure 6 is a graph which represents the possible compositional makeup of all combustible solid waste feedstocks. The graph shows that the composition can be characterized by three basic elements, non combustibles, water and combustibles. This tripartite representation of the solid's composition allows for a comparison of the full range of solid wastes amenable to thermal reformation as well as a discussion of the role and effect of the three compositional components. The shaded portions of the graph is the autothermic region and it represents range of compositions, which inherently contain enough energy in their combustibles

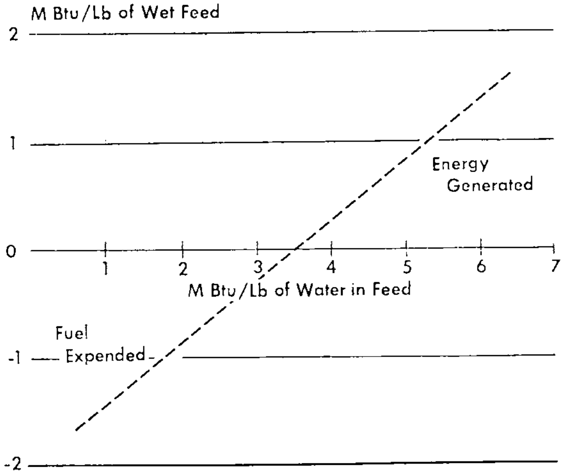


Figure 1. Rule of thumb

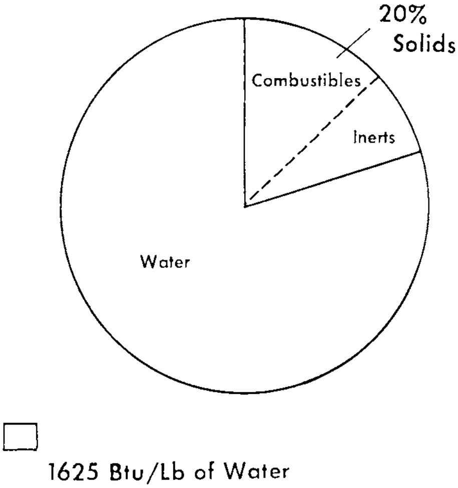


Figure 2. Sludge

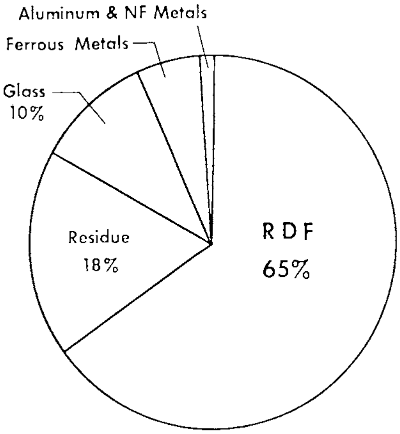


Figure 3. Municipal refuse

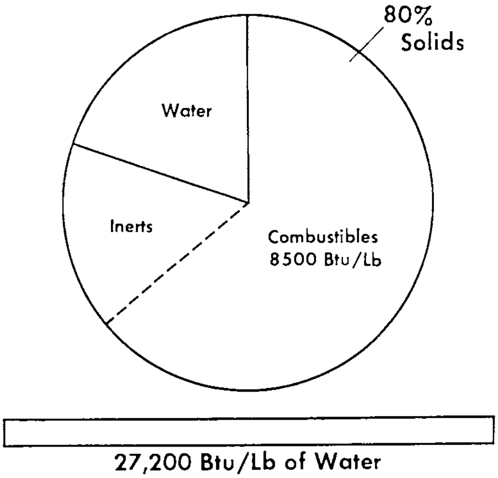


Figure 4. RDF

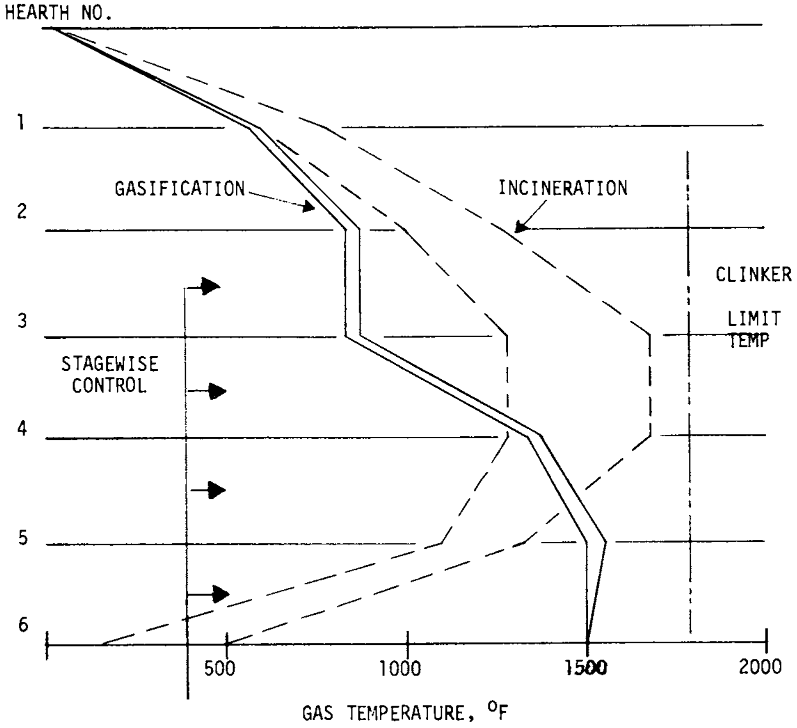


Figure 5. Operating range

fraction to offset the diluting effects of the moisture and inert fraction. This figure is analogous to Figure 1 but in this case the noncombustible fraction of the feedstock is represented as well as the water fraction. The autothermic region represented here appears to be constant for all combustible feedstocks near 9000 BTU/lb combustibles, and has been established by testing a wide range of solids including RDF, sewage sludge, walnut shells, wood, ferro-manganese solids and onion skins. The compositional limits of autothermicity run between 70% water and 60% non combustibles. In the region bounded by the curve between these two points, the solids are capable of sustaining gasification without fuel addition. At the higher combustible fraction and BTU contents a net energy gain can be produced from the solids.

In Figure 7 the thermal reformation of the 40% water, 40% combustibles, 20% non combustibles feedstock is represented by the pathway between the feedstock composition and the solid product. The pathway represents the sequential changes in feedstock composition that are brought about by the gasification of the solid. The feedstock, on initial exposure to the hot environment of the thermal reformer, increases in temperature until the water fraction of the feedstock begins to evaporate. The solids remain at approximately 100°C throughout the greater portion of the drying stage and do not increase until the largest portion of the water has been evaporated. The loss of feedstock moisture is represented by the segment of the pathway labeled 'DRYING'. In all feedstocks the initial solid transformation is the evaporation of the feedstock moisture.

As the last of the moisture is driven from the solid, the solids temperature increased to 300°driving off the lower molecular weight volatiles and the last remaining water. As the solids temperature increases even further to 400°C the heavier weight volatiles are driven off and the increasing thermal fluxes liberates some of the fixed carbon from the solid. This second reformation phenomenon is represented by the 'VOLATILIZATION' segment of the reaction pathway.

The final stage of the thermal reformation is the reduction of the last remaining combustibles in the solids. Here the fixed carbons are reduced to produce a gaseous product which is reclaimed for its sensible and latent heats of combustion in the downstream process combustor. The solid may be reacted to completion which would result in a 100% non combustibles product. However, the volatilization of low levels of fixed carbons is subject to deminishing returns for energy recovery and some carbon is usually left in the solid product. This carbon is not wasted because it helps retain certain components (i.e. heavy metals) in the solid fraction which would be detrimental to the environment if they were allowed to escape.

These three sub phenomena of gasification strongly affect the efficacy of various gasification reactor technologies. The multiple hearth gasification system developed by Envirotech

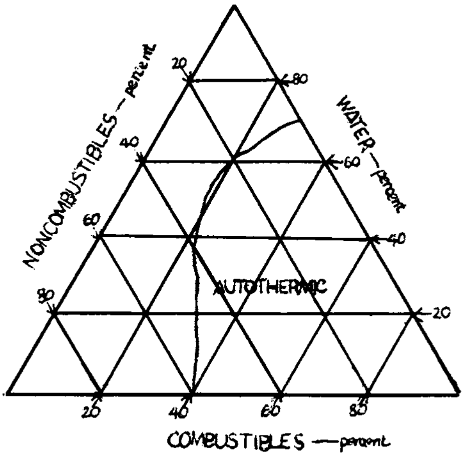


Figure 6. Compositional triangle

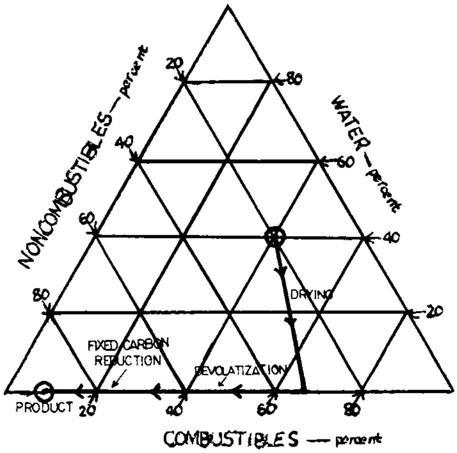


Figure 7. Reaction pathway

provides the greatest accommodation of the gasification phenomena for carbonaceous solids. Figure 8 depicts the operation of the multiple hearth gasifier. Feedstock solids are introduced into the top section of the furnace. The furnace is segmented into discrete process stages, called hearths, where the sequential thermal reformations are carried out. The solids are constantly mixed and transported across each hearth by air cooled mixing arms located on a central shaft. The solids drop from one hearth to the lower hearth after they have been mixed and reformed for the specified residence time on the hearth. The gases evolved in the reformation process flow counter currently to the down flowing solids. This counter current flow provides for efficient heat transfer from the hot gas to the solids. The solids exit the reactor at the bottom, essentially free from combustibles and the low calorific gases exit at the top of the reactor where they are transported to an afterburner and heat recovery system. The hearth staging provides compartmentalized regions for each of the three gasification subphenomena to occur. Each hearth provides a platform for the solids to undergo the sequential thermal reformations. The constant solids stirring on each hearth overcomes the problems of stagnant reaction areas and the low velocity, countercurrent gas contacting provides efficient heat transfer without solids entrainment. The importance of constant agitation of the solids should not be overlooked. In a commercial demonstration facility it was shown that a cessation of mixing decreased the gas production rate by nearly an order of magnitude. Not only does this mixing prohibit stagnant zones but it promotes the solids particle breakdown which exposes fresh reaction surfaces and facilitates the solids transfer from one stage to another.

A temperature profile increasing from top to bottom is maintained in the multiple hearth gasifier. This profile accommodates the reaction requirements of each of the three subphenomena. The heat for drying, volatilization and fixed carbon reduction is generated in the gasification zone of the lower hearths where air is admitted in sub stoichiometric quantities. This allows for combustion of some of the fixed carbon and the associated heat release volatilizatizes the remaining fixed carbon. In addition to air, steam can be used to gasify the fixed carbon via the water-gas shift reaction (see Figure 9).

The hot gases from the gasification zone flow upward into the pyrolysis zone. The sub-stoichiometric quantities of air are consumed by the fixed carbon in the lower stages so the rising gases are reductive. In the pyrolysis zone the volatiles and some fixed carbon are thermally reformed in the reductive environment.

The drying zone utilizes the heat remaining in the liberated gases to drive off the moisture. As can be seen, the sequential drying, volatilization and reduction phenomena of gasification are accommodated and individually controlled in the integrated,

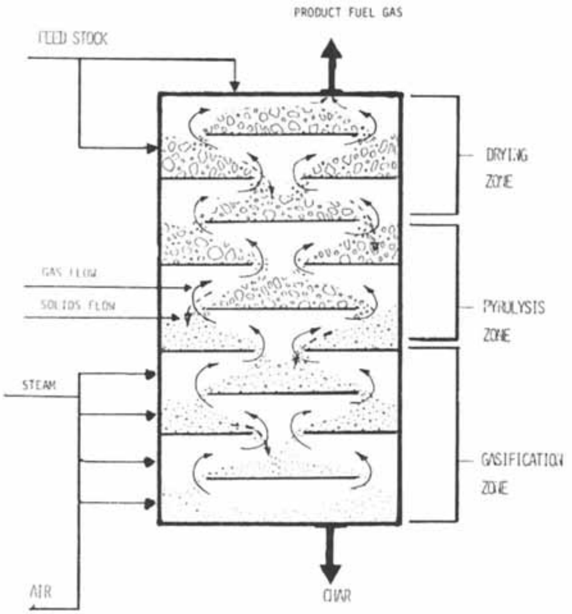


Figure 8. Multi-hearth gasifier

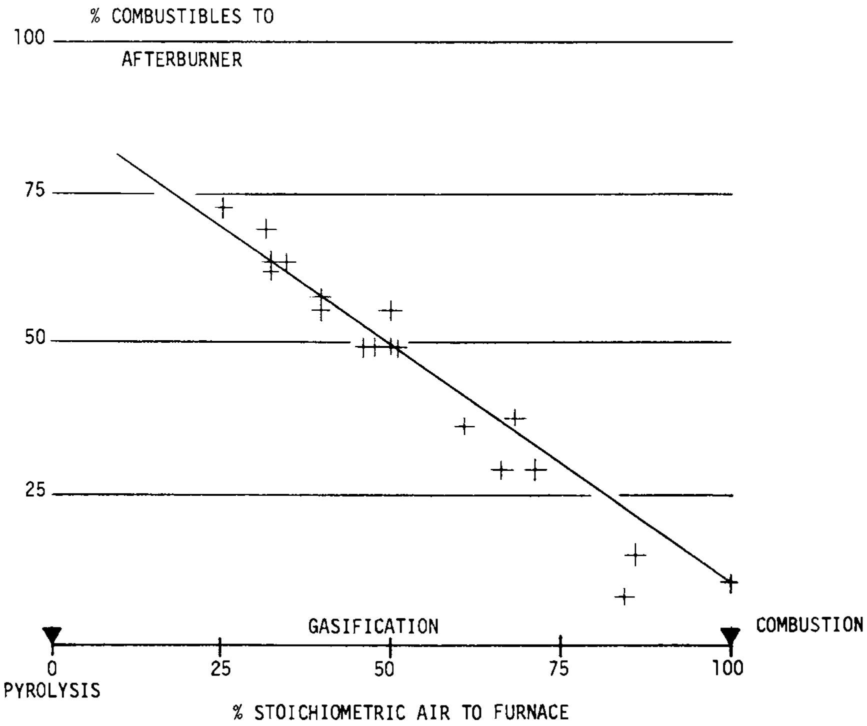


Figure 9. Air requirements

staged and countercurrent multi hearth gasifier. The low BTU gases evolved from the solids are normally 130 BTU/SDCF but strongly depend on the feedstock. The gas is burned in an external afterburner which maximizes the utilization of the fuel value of the gas via an Envirotech temperature/oxygen control system. The control system maximizes the afterburner combustion temperature thereby increasing the heat recovery efficiency without producing solids agglomeration problems. The Envirotech control system also provides for control of the temperature/oxygen profile throughout the gasification reactor. This maximizes the rate of reaction as well as the calorific value of the reformed gases.

Gasification has been termed "destructive distillation" because of its irreversible separation phenomena. In fact, there is some merit in the distillation analogy based on experience gained by Envirotech with the MHGS. The stage wise, countercurrent, mixed, and temperature profiled characteristics of the Envirotech MHGS bear a striking resemblence to liquid distillation columns. The difficulty of transporting and contacting solids makes gasification significantly different from liquid distillation but the analogy is generally valid. What has been found to affect control, operation, cost and product quality for liquid distillation has its counterpart in the "destructive distillation" of solids.

The overall process control variables for gasification are listed in Table 1. The process control system uses these variables to maximize energy generation, protect the equipment and provide stable operation even with fluctuating feedstock characteristics. In addition, the patented temperature/air control system provides the special zonal control which efficient gasification requires. The total MHGS is depicted in Figures 10,11. Both sludge and RDF are shown as feedstocks in what is described as co-gasification. The high moisture sludge is fed into the uppermost hearth to take advantage of gas drying while the drier RDF is introduced at a lower hearth. Combination of feedstocks in this manner has definite process advantages. Figure 12 shows the combined BTU/lb water ratio for RDF and sludge (3:1). The three to one RDF to sludge ratio was chosen because it represents typical municipal generation rates of RDF and sludge. The mixture is significantly above the autothermic region for gasification, as is even a 1:1 ratio mixture. Thus, both wastes can be disposed of with a net energy gain via co-gasification. Incineration does not offer this option since moisture induced control problems in the burning region limit sludge additions to small quantities.

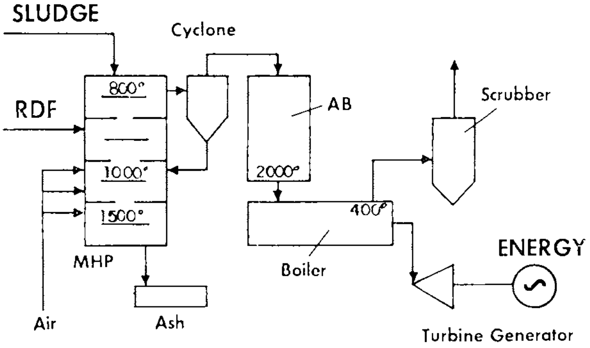


Figure 10. The MHP system

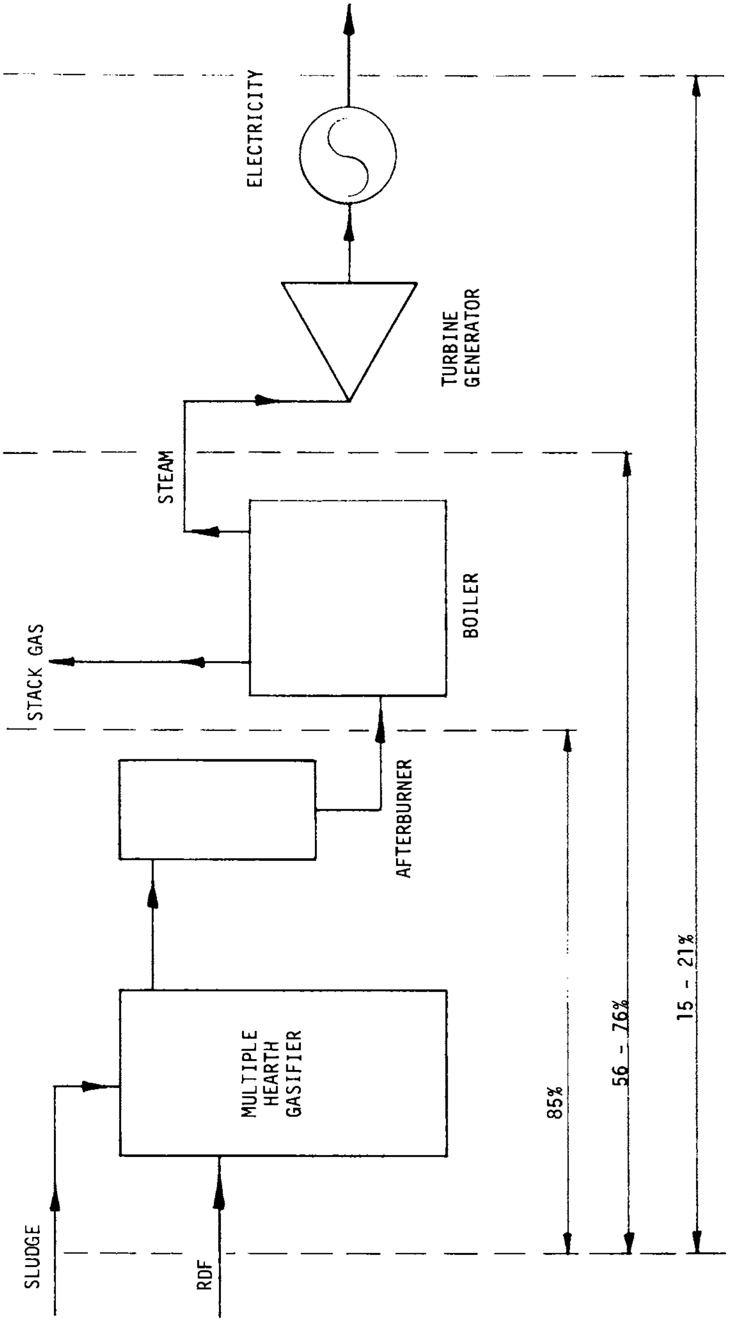


Figure 11. Process efficiencies

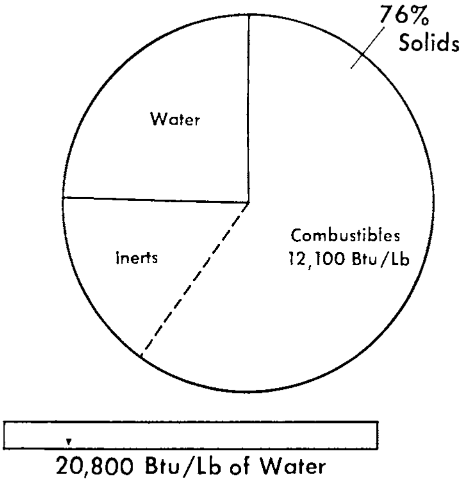


Figure 12. RDF and sludge (ratio 3:1)

TABLE I

CONTROL VARIABLES

FEED RATE

FEEDSTOCK BTU/POUND WATER

COMBUSTIBLES/AIR RATIO

TEMPERATURE PROFILE

MAXIMUM PROCESS TEMPERATURE

PARTICLE SIZE AND DENSITY

RESIDENCE TIME

AFTERBURNER AIR/FUEL GAS RATIO

The commercial scale co-gasification of sludge and RDF has been accomplished by Envirotech in an EPA sponsored demonstration at Concord, California. Varying ratios of the wastes were gasified to produce energy in a 144 TPD, 16' diameter six hearth furnace. The system, shown in Figure 13, was constructed around an existing multi-hearth sludge incinerator which was modified to process specifications. The system processed a wide range of sludge/RDF mixtures as well as 100% sludge and RDF. The data showed that the energy recovery, pollutant emissions and operating stability of the co-gasification was constantly superior to incineration.

Table 2 presents some of the test data from the Concord tests. As can be seen, a wide range of feedstock compositions were evaluated. Even non-autothermic sludge was gasified utilizing auxilliary fuel. In the other instances, autothermic mixtures were gasified without auxillary fuel. Figure 14 presents the material balance for a typical feedstock evaluation and Figure 15 presents the analogous heat balance. These demonstrations prove the capability of the Envirotech MHGS to process a wide range of feedstocks as well as the capability to accommodate instantaneous changes in feedstock composition without process upsets. In fact, the gasification process appeared less sensitive to moisture increases in the solids than did incinerative disposal.

The environmental consequences of co-gasifying sludge and refuse in a MHGS were shown in the demonstration at Concord, California to be less than those of incineration. In fact, the Envirotech multiple hearth system surpassed both EPA and San Francisco Bay Area Pollution Control Board (SFBAPC) standards for gasification process emissions (see Figure 16). The strict control of the temperatures and the oxygen concentrations not only

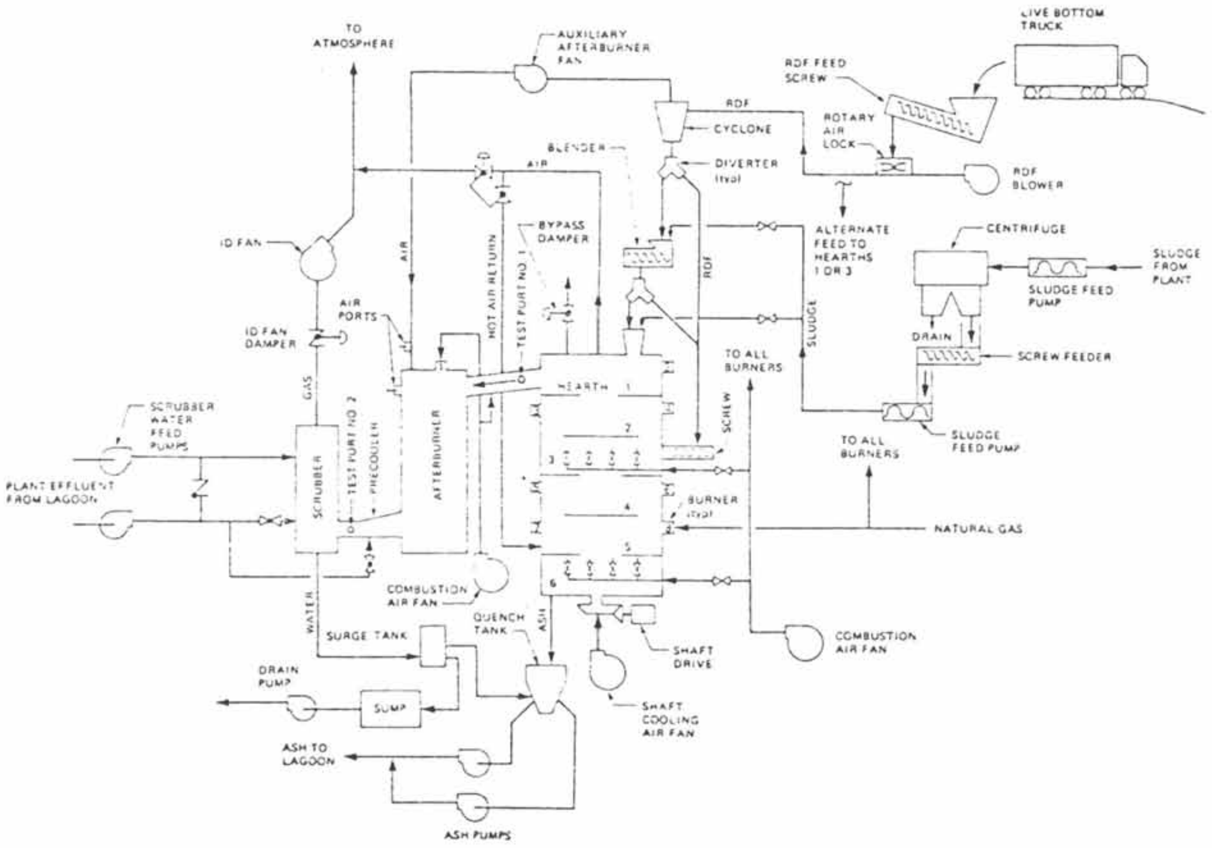


Figure 13. Concord commercial demonstration system

Table II. Furnace Operating Data for

Test number	RDT/ sludge (wet basis)	Total wet feed, lb/hr	Percent solids in total feed	Total feed high heat value[b], Btu/lb of wet solids
19H1[g]	2.1/1	5,408	56.4	4,249
26L	1.2/1	5,562	51.7	3,862
27M	1/0	3,822	80.9	5,865
29N	1.1/1	5,441	45.8	3,401
30 0	0/1	4,110	24.0	1,956
31P	1.5/1	6,217	48.2	3,386
33	2.3/1	6,189	59.2	4,310
34	1.1/1	5,605	49.7	3,682
35	1/2.1	3,981	42.8	3,384
36	2.9/1	6,818	51.8	3,462
37	1/1.3	7,304	40.2	2,028
38	2.5/1	8,716	58.5	3,881
39	1.1/1	7,457	43.7	3,068

[a] Vista=Vista Chemical and Fiber Products. CRS/UCB=Cal Recovery Systems/University of California at Berkeley.

[b] High heat value (gross heat of combustion) determined by ASTM Standard 0240-64 (burning weighed sample in an oxygen bomb calorimeter).

[c] Includes any natural gas added to either the furnace of the afterburner.

[d] Includes latent heat, sensible heat and heat value of unburned combustibles.

[e] Includes latent heat and sensible heat.

[f] Computed as the ratio of Btu's/hr into furnace divided by Btu's/hr out of afterburner.

Selected Material and Heat Balances

Auxiliary fuel[c], Btu/hrx10^6	Furnace off gas[d], Btu/hr x10^6	Afterburner off gas[e], Btu/hr x 10^6	Process efficiency[f] percent
None	13.50	13.36	58
None	17.46	17.4	81
None	18.44	18.24	81
None	15.01	14.84	80
1.83	7.05	8.89	90
None	15.59	15.49	74
None	19.58	19.48	73
None	18.36	18.24	88
None	10.67	10.51	78
None	21.16	21.09	89
None	18.51	18.35	89
None	24.18	24.12	71
None	16.99	16.81	73

[g] RDF feed rate for test numbers 19H and 19I (see Table 4-1) averaged for heat and material balances.

[h] Preprocessed Vista RDF used.

Note 1: Latent heat refers to the latent heat of vaporization of water, or that heat absorbed when water is evaporated (or released when evaporated water is condensed). It is normally taken at 60°F. and 14.7 psia and is about 1000 Btu/lb at these conditions.

Note 2: Sensible heat is defined as the specific heat in a gas or liquid or all heat available above the latent heat of vaporization of water in the gas.

Note 3: Mean of pyrolysis mode efficiencies is 79 percent.

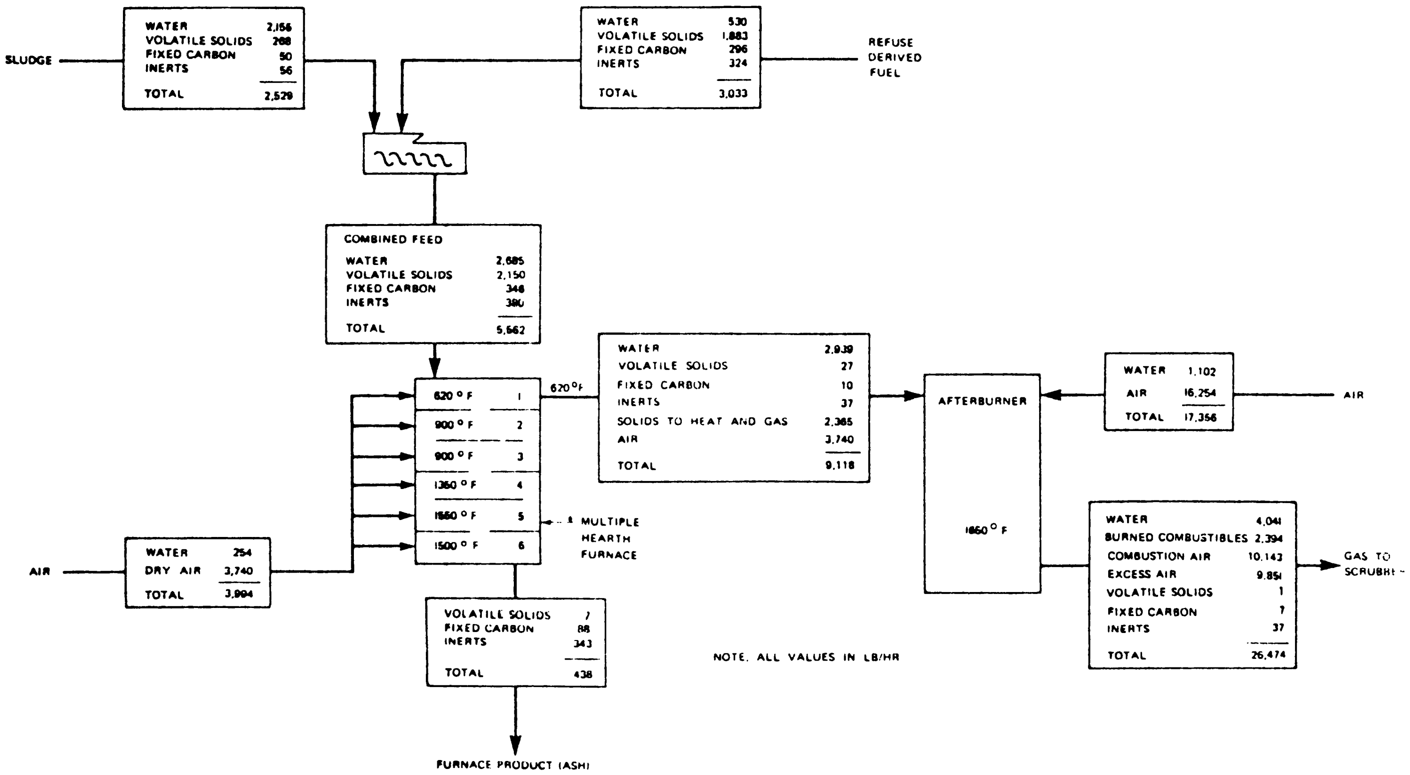
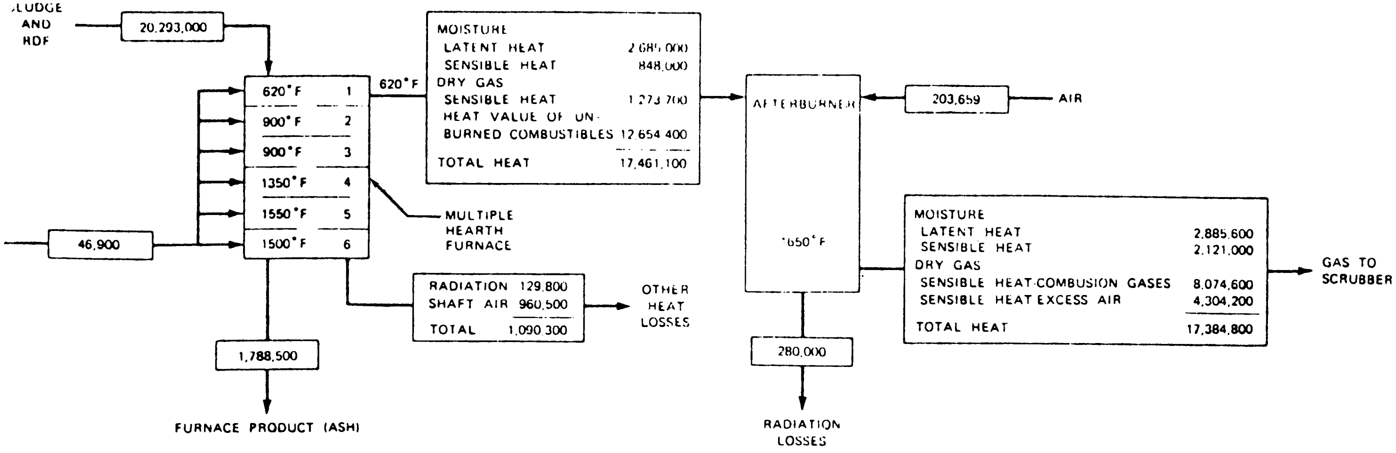


Figure 14. Typical material balance



NOTE 1. ALL VALUES IN BTU/HR
NOTE 2: MOISTURE INCLUDES FREE WATER IN FEED AIR, COMBINED WATER IN FEED, WATER PRODUCED BY COMBUSTION OF REMAINING HYDROGEN IN FEED.
NOTE 3: EXCESS AIR IS AIR ADDED IN ADDITION TO STOICHIOMETRIC AIR REQUIRED FOR COMPLETE COMBUSTION

Figure 15. Typical material balance

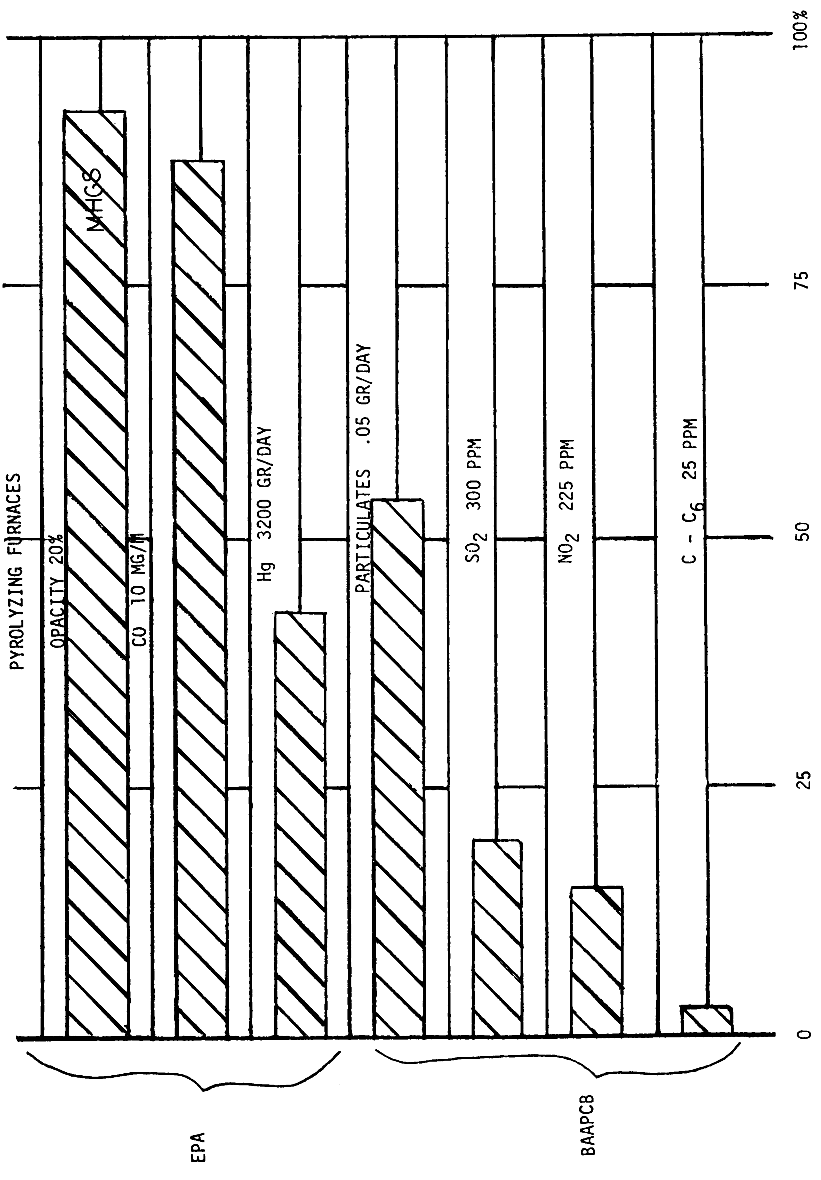


Figure 16. Emission standards

provides efficient energy recovery from the solids but it also minimizes their detrimental effects on the environment. In general, solid wastes make good renewable fuel sources but their composition may include significant amounts of detrimental substances such as heavy metals. The low solids temperatures of the MHGS minimize the metal volatilization rates and promote their fixation in the carbon matrices of the char product. This retention is characterized by the gas emission of less than 1600 grams per day of mercury into the atmosphere.

The low velocity gas contacting, made possible and efficient by the stirring of the solids, promotes a low particulate loading of the gas stream. A reduction in the particulates is effected in the afterburner and in downstream gas scrubbing.

Sulfur dioxide and hydrocarbon emissions were less than one quarter of the standards and obviously pose no environmental threat. The exceptionally low nitrogen dioxide emission levels are unique to gasification. The solid waste feedstocks may contain significant amounts of nitrogen compounds. In the MHGS these nitrogen compounds are almost all reduced to ammonia. The presence of ammonia in the afterburner, in conjunction with any nitrous oxides and oxygen, promotes the formation of diatomic nitrogen, N_2 and limits NO_x production. Thus, a potentially noxious pollutant is transformed into an inert gas by the reductive/oxidative staged environment of the MHGS.

The economics of waste gasification can be quite attractive as an environmentally sound disposal option and as means of developing renewable energy source. The projected economics of sludge/RDF co-gasification are presented in Table 3 as an example. The RDF/sludge ratio is varied from 3:1 down to 100% sludge to provide a comparison of feedstocks. These figures are derived from the Concord commerical demonstration studies and based on the data from that project a new disposal facility is being constructed at the Central Contra Costa Sanitation District in California. The design is based on a combined RDF/sludge rate of 888 TPD at 60% solids. The superheated steam produced will drive the in-house machinery and generate electricity.

One of the important realizations to emerge from the studies was the economic benefit which could be accrued from combining wastes for co-gasification. In particular, the low BTU/lb water ratio of sludge could be offset with the high BTU/lb water ratio of RDF with very profitable results. In essence, the RDF is acting as a very inexpensive auxillary fuel to aid in the gasification of sludge and the energy potential of both feeds are captured. The economic benefits of combining feedstocks for co-gasification are attainable with a multiple hearth gasification system. The gasification of carbonaceous wastes in an Envirotech MHGS has several advantages over other disposal systems. It is immediately available and it is not an environmental threat. In addition, it is an economically attractive disposal technology because its major product is

Table III.

Case	I	II	III	IV
Description	Sludge only	RDF/Sludge	RDF/Sludge	RDF/Sludge
Ratio	1	1/1	2/1	3/1
Wet Ton/Day	125	125/125	250/125	375/125
% Solids	20	50	60	65
Capital Cost $/yr	131,000	209,000	244,000	349,000
Maintenance $/yr	4,000	6,300	7,300	10,500
Labor $/yr	128,000	128,000	192,000	192,000
Fuel $/yr	383,000	---	---	---
Power $/yr	35,000	42,000	49,000	70,000
RDF $/yr	---	228,000	456,250	684,375
Heat Recovery + $/yr	129,210	876,000	1,989,250	2,974,750
-Cost Total +Earned	-732,790	+262,700	+1,040,700	+1,968,875
$/T Wet	- 16.00	+3	+8	+11
$/T Dry	- 80.00	+6	+13	+17

Note: Capital cost annualized 6%/20 yrs for RDF handling/MHF/AB, WHRB/Scrubber installed, Fuel $3/mm BTU, RDF $5/T, Power 4 ¢/KWH.

Flowsheet includes RDF storage and handling, multiple hearth gasifier, afterburner, waste heat recovery boiler and scrubber. RDF costs based on $5/ton and steam revenue on $4/1000 lb steam.

inexpensive energy, either for in-house use in place of expensive fuels or for sale. Also, the co-gasification of wastes, possible with MHGS, has the added advantage of being an integrated solution to the solid waste disposal dilemma. The hearth by hearth control, countercurrent flows and stirred characteristics of the Envirotech multiple hearth gasification system provide the best control and optimum process conditions for the conversion of carbonaceous wastes to energy. The Envirotech MHGS is capable of meeting today's solids disposal and energy supply needs with a technology that is commercially proven and cost effective.

March 3, 1978.

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

10

Pyrolyzer Design Alternatives and Economic Factors for Pyrolyzing Sewage Sludge in Multiple-Hearth Furnaces

CHARLES VON DREUSCHE and JOHN S. NEGRA

Nichols Engineering and Research Corporation, Homestead and Willow Roads,
Belle Mead, NJ 08502

While multiple hearth furnaces have been used commercially in ten large plants for pyrolysis of bark and sawmill waste to charcoal for briquette manufacture, used commercially for pyrolysis of paper and plastic laminates on aluminum foil to recover the foil, (1) and demonstrated experimentally for pyrolysis of RDF alone (2) and for co-pyrolysis of RDF and sludge,(3) this paper will be limited to considering pyrolysis of sewage sludge alone.

Many sewage treatment plants are equipped with multiple hearth furnaces for sludge incineration. This transforms an offensive and dangerous waste into a safe and inoffensive ash, and satisfies various purposes listed in Table I.

The need for this transformation is likely to continue since sewage sludge is the one most likely place to find, perhaps as an unrecognized by-product of organic synthesis, organic substances that are found to be dangerous and non-biodegradable, perhaps long after their initial appearance. Thermal conversion of the complex organics to simple inorganic form is the only way to insure that a dangerous substance is not widely distributed as a sewage plant product.

Most of the many multiple hearth incinerators now use oil or gas to supply the heat needed to evaporate 2 to 4 pounds of water received with each pound of dry sewage sludge solids and heat this water along with 100% excess air from combustion of the organics to a legally required 1400 ^{o}F afterburner exhaust gas temperature.

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TABLE I

Reasons for Thermal Destruction of Sewage Sludge

A. Change Physical Properties
  1. 10:1 weight reduction from 30% solids, 33% ash dry basis to 100% ash.
  2. 5:1 volume reduction
  3. Sticky slime to powder or granules

B. Public and Employee Safety and Aesthetics Promoted by Eliminating:
  1. Odor
  2. Potential odor by decomposition
  3. Pathogens
  4. Potential pathogen multiplication if contaminated.
  5. Support for larvae, rodents, etc.
  6. Organic Toxins - including those which may be unidentified or whose toxicity may be currently unrecognized.

C. Energy Conserved
  1. Less trucking and landfill operating fuel
    (a) because of volume and weight reduction
    (b) because nearer site likely to be acceptable.

D. Reduced Handling and Disposal Cost
  1. Acceptable land disposal may be much nearer
  2. Reduced transport quantity and distance from plant
  3. No cover dirt required
  4. Leachate control may not be required
  5. Smaller land fill area to buy or longer life of existing landfill

By considering simultaneous application of pyrolysis and modern sludge dewatering practices, it is possible to accomplish the destruction of sludge organics without the need of any auxiliary fuel, and substantial energy will be available for recovery.

Substitution of refuse derived fuel is another means of conserving gas and oil. However, RDF costs money to prepare from refuse. The cost depends on volume through-put. Large scale uses,needed to justify RDF preparation plants, will no doubt govern plant location and market price.

A high priority fuel requirement that justifies pyrolyzing a large amount of RDF for its fuel value in conjunction with the sludge may on the other hand make co-pyrolysis very attractive in some situations.

A population of 400,000 could be expected to generate about 800 tons of refuse per day. The corresponding 40 MGD wastewater treatment plant would require about 24 tons RDF per day to make up the fuel deficiency for pyrolysis of their sludge if they are able to dewater the sludge to only 25% solids. The fair price of the RDF, plus transportation, plus storage and metering costs for this bulky, odoriferous bacterially decomposable fuel plus incremental furnace size required to burn the RDF may make this unattractive in some circumstances.

How typical the 40 MGD plant is in size, since it will be carried through this paper as an example, may be judged by reference to Table II. The magnitude of energy available, energy demand and energy dollar worth may also be judged from this Table.

It is foreseen that energy may be recovered in 3 ways. The hot gas exiting an afterburner on the pyrolysis operation can be used to generate steam. Energy can be recovered from a 25% carbon/75% ash char produced by pyrolysis. The char so produced need not be consumed in the sewage treatment plant, and if it is, it need not be consumed at the instantaneous rate that it is generated from sludge. A third possibility for heat recovery is the generation of methane from digesters, made economical by the opportunity created by improved dewatering plus pyrolysis, to safely destroy the energy depleted sludge produced by digesters without purchased fuel and to return to the digesters

TABLE II

ESTIMATED ENERGY - SEWAGE SLUDGE (10)

EXISTING PLANTS

PLANT CAPACITY mgd	NUMBER PLANTS	MILLION BTU/YEAR (IN THOUSANDS) REQUIRED PER WTP	MILLION BTU/YEAR (IN THOUSANDS) AVAILABLE IN * SLUDGE/WTP (NET)	MAXIMUM ** $/YR.
5	12761	5	7	21,000
5 - 10	494	28	28	84,000
10 - 20	267	59	67	201,000
20 - 50	180	125	133	399,000
50 &	83	289	175	525,000

*BASIS OF EXPECTATION - SLUDGE AT 40% SOLIDS (SEE FIG. 2)

** BASIS $3 PER MILLION BTU.

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

heat they require. Digester gas is relatively simple to produce, and to clean for transmission, and it is a high BTU gas.

It is not the purpose here to examine in depth the many alternative uses for the surplus energy generated by improved efficiency in thermal destruction of sludge, nor to question how likely the actual utilization of such energy may be. It must be noted however that in many plants where digester gas is produced, the gas is wasted. The exhaust gas of incinerator afterburners is also at 1400 °F and in larger volume than will result under the proposed scheme of improved dewatering plus conversion to pyrolysis; yet only about 20 plants, including 12 yet to come on stream, actually practice this recovery (4). It is believed that pyrolysis, by providing more flexibility as to the form of recovery, plus the opportunity to store energy as carbon and recover heat from that, at times of peak energy demand, will make such recovery more practical.

Pyrolysis of sludge alone without auxiliary fuel has been demonstrated and explored over a total of 400 hours operation in a 3 ton per day pilot operation utilizing two different raw sludges and one digested sludge. The pilot furnace is shown in Figure 1. The program was carried out in the pilot plant of Nichols Engineering and Research Corporation for the Interstate Sanitation Commission of New York, New Jersey and Connecticut as part of a program funded by the U.S. Envirommental Protection Agency. Release of the report of that program is expected soon.

Meanwhile one full scale unit for the pyrolysis of sludge and the recovery of carbon from a 40 MGD industrial wastewater treatment plant is now coming on stream and the first municipal contract for two units pyrolyzing sludge alone has been bid and Nichols expects award of the contract momentarily.

In focusing on the pyrolysis of sludge alone, it is the viewpoint of this paper that in-plant elimination of sludge organics is the primary aim of the operation. Eliminating the purchase of valuable fuels such as RDF, oil, gas or coal, is the first step to economy and conservation. The recovery of heat from use of the char will follow in an increasing number of

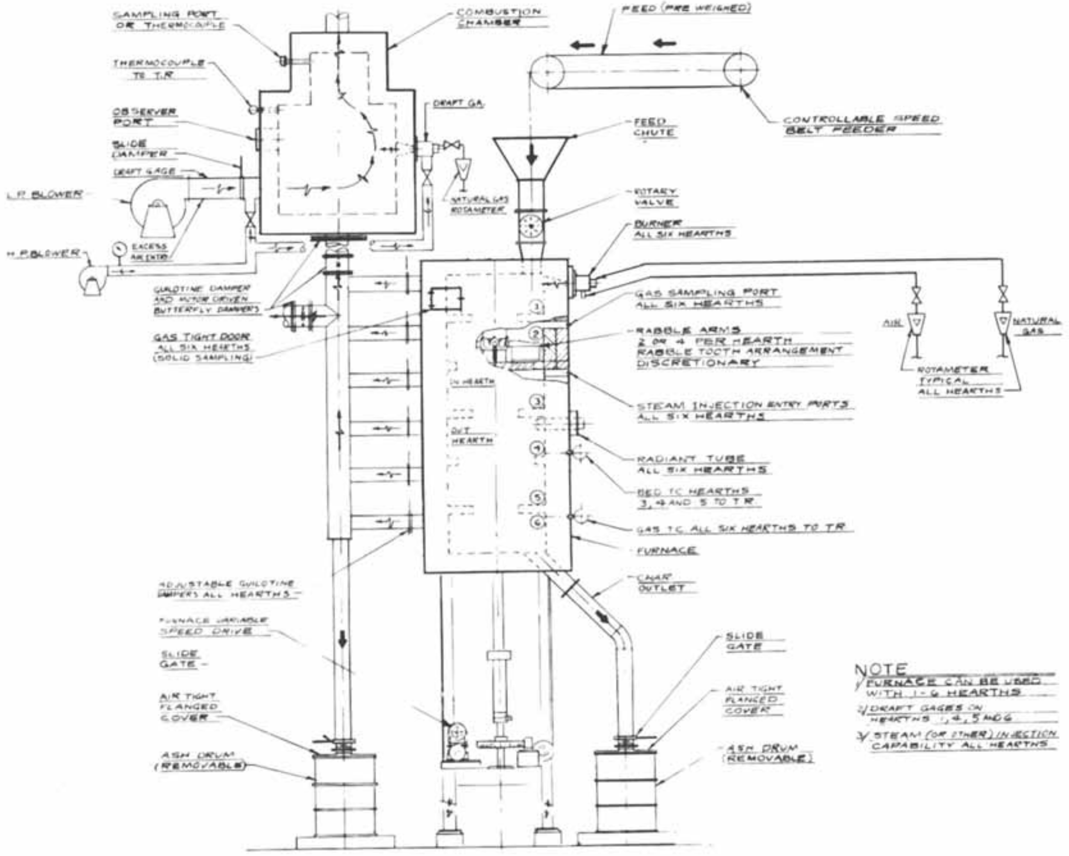


Figure 1. Pilot plant

instances.

The conversion of existing units to pyrolysis seems eminently practical but first some choices must be made between process design alternatives leading to different products. These choices depend on the individual plants requirements, opportunities for energy recovery and the nature of its sludge.

Pyrolysis Process Description and Operating Modes

The Pyrolysis Process. The thermal destruction of sludge by pyrolysis in a multiple hearth furnace is a process by which solids are raked along the floor of successive compartments while a larger vapor space exists above the solids. When limited amounts of air are properly introduced into the vapor space the oxygen reacts with the organic vapors present and there is starved air combustion (not pyrolysis) of these vapors.

The oxygen does not, in the proposed designs, persist through the vapor space and reach the solids. The solids receive heat by radiation. They are shielded from contact with oxygen by a stagnant film of organic vapor or water or both leaving the stirred bed of solids. Adding to the protection is the fact that the limited amount of air admitted has its oxygen consumed rapidly in the organics rich vapor phase. This is pyrolysis with respect to the solids.

Hot products of starved air combustion from the middle hearths provide the required heat for drying the wet incoming sludge on the upper hearths. Combustion of the gaseous products is completed in an afterburner.

The essential point of the pyrolysis process in reducing or eliminating fuel requirements is the elimination of large excess air requirements common to incinerators. The middle hearths of the multiple hearth incinerator receive and burn sludge that has been predried on the upper hearths. Approximately 100% excess air is required to dilute the temperature of combustion of this sludge to the 1800 ^{o}F upper limit for which normal multiple hearth construction is suitable, as can be shown by a heat balance around this portion of an incinerator. Pyrolysis eliminates this need for excess air by postponing the completion of combustion until after partially burned organic vapors have passed over the wet incoming sludge where they will cool, pick up water vapor and may pick up additional organic vapors, steam distilled toxins, or odors. The organic vapors provide fuel in the after burner to raise the temperature of water evaporated in the upper hearths to the required 1400 °F final combustion temperature.

Low Temperature Char Operating Mode. The organics in the sludge have undergone destructive distillation giving off almost all the organic vapors that can be released, when the solids have been heated to 1200 to 1300 °F. The process can be called complete at that point and the char discharged.

This would be low temperature char (LTC operating mode). The product is a sterile, black, charcoal-like free flowing mixture of powder and granules having a bulk density of approximately 16 pounds per cubic foot and containing 75% ash. It contains all of the ash of the feed sludge plus amorphous carbon and some non-putrescible unleachable material akin to the volatile content of coal. It is believed that this material could be safely landfilled satisfying the objectives of Table I.

This material has been observed to have retained the lead, zinc and cadmium content of sludge to an extent not usual in other modes of pyrolysis operation nor in incineration. Allowing the solids temperature to exceed 1200 to 1300 °F seems to initiate volatilization of these substances to a fume which is difficult or impossible to capture in most air pollution abatement equipment.

Under development are methods to recover these metals from the low temperature char. One half percent total lead, zinc, cadmium content in the char is not uncommon for sludge in industrial areas (5). Recovery is the one way to avoid putting the 80,000 pounds per year (more or less) of these metals, that a 40 MGD plant might have, into either the land, the water, or the air.

High Temperature Char Operating Mode. For high temperature char, the char can be brought to temperatures of 1600 °F or higher either by allowing a high gas temperature in the last hearth of the pyrolysis zone, or by very limited contact of the char with air on a hearth at the end of the pyrolysis zone.

The product is about 25 pounds per cubic foot and contains only ash and carbon. The carbon content can be estimated from the data in Table III. It ranges from 7 to 18% of the organics in the sludge and will vary from about 10% to about 40% of the product depending on the kind of sludge and the ratio of ash to organics in the sludge.

Lack of restriction in applying high gas phase temperatures, if one need not carefully avoid heating the char above 1300 °F, allows a greater heat transfer rate throughout the furnace. Thus the furnace for operation in this mode can be of minimum size for a given throughput capacity. Also raising the temperature to this level does reduce the heating value of the char discharged, making more heat available in the furnace or afterburner.

Char Burning Operating Mode. The char resulting from pyrolysis in the middle hearths can be converted to ash on the lower hearths, (CBA operating mode).An oxygen free atmosphere can be maintained as in the pyrolysis hearths and the carbon reacted with CO_2 and H_2O to produce CO and H_2. This requires a temperature of at least 1600 °F to obtain practical reaction rates (6). Alternatively air, or air plus steam (7), can be circulated across these lower hearths and substantially vented from the side of these lower hearths to avoid an excess of air entering the middle pyrolysis hearths. An excessive amount of oxygen entering the hearths where maximum organic vapor release occurs, risks development of temperatures there in excess of the 1800 °F that normal multiple hearth furnace construction can withstand. The oxygen containing gases vented from the side of the lower hearths are brought by a separate external duct to the afterburner where these gases serve as preheated air for combustion, thus utilizing the heat obtained from the carbon to raise the afterburner temperature.

TABLE III

SOME COMPOSITIONS OF VARIOUS SEWAGE SLUDGE SOLIDS TESTED

SAMPLE NUMBER	COMBUSTIBLES - %		% ASH	TYPE SOLIDS
	VOLATILES	FIXED CARBON		
1	63	18	19	AEROBIC DIGEST
2, 3, 4	48-66	6-12	28-40	ANAEROBIC DIGEST
5, 6, 7	59-67	11-17	18-24	WAS + PRIMARY
8	71	8	21	AERATED PRI.
AVERAGE	62	12	26	

The advantage of reacting away the carbon resulting from the pyrolysis zone of the furnace is to avoid a loss of heat from the process in the form of unburned fuel value as carbon in the material produced. The material produced consists entirely of ash and has a bulk density of approximately 33 pounds per cubic foot.

Comparison of modes. Table IV shows the reasons for selecting one mode of operation as compared to another.

It should be pointed out that a single furnace sized large enough to react away all of the fixed carbon, can be operated in any of the three modes, LTC, HTC, or CBA, by simple variation in control settings. This is important later when considering the possibility of storing and refeeding char product.

Carbon Yields and Reaction Rates. It should however be pointed out that maximum energy production from a given quantity of sludge, by reacting away all of the fixed carbon formed in pyrolysis, does require additional hearth area. The extra hearth area required can be found by estimating the pounds per hour of fixed carbon likely to be formed from the organics in the sludge, and dividing by a rate of reaction.

Some typical carbon reaction rates are shown in Table V. Typical fixed carbon yields found for various sludges are indicated in Table III.

The importance of this factor may best be shown by example. The 40 MGD plant carried through this paper as an example, pyrolyzing 5000 pounds of sludge solids (dry basis) per hour at 25% solids (alternately 40% solids) requires 1,333 square feet (alternately 833 square feet) to reach the HTC stage of product. This area requirement can be calculated from the usual heat transfer equations in the form of q (heat transfer to solids)=U(a co-efficient of transfer) X Area X gas to Solids Temperature Difference. Based on a typical 12% fixed carbon yield and 0.6 pounds reacted per square foot, the oxidation of the 600 pounds of fixed carbon formed requires that an additional 1000 square feet be added

TABLE IV

Comparison of Pyrolysis Modes of Operation

	LTC	HTC	CBA
Loading rate (approx.)	10 lbs./Hr-SF.	13 lbs/Hr-SF.	9 lbs/Hr-SF
Heavy Metals Retention	Maximum	Less	Less
Chrome as Cr+3	All as Cr+3	All as Cr+3	A little may remain oxidized still under study
Heat Recovery	Minimum	Intermediate	Maximum
Landfill Volume	Maximum	Intermediate	Minimum
Bulk Density	16 lbs/ft^3	25 lbs/ft^3	33 lbs/ft^3

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

TABLE V

CARBON GASIFICATION RATE

TEMPERATURE GAS - ^{o}F	CONCENTRATION IN GAS* O_2	CONCENTRATION IN GAS* CO_2 + STEAM	RATE CARBON GASIFICATION LB/HR. - FT^2
1823	0	HIGH	0.86
1742	MODERATE	MODERATE	0.67
1596	0	HIGH	0.30
1608	MODERATE	MODERATE	0.30
1588	MODERATE	MODERATE	0.25
1599	HIGH	LOW	0.40

*CONCENTRATION IN GAS

	O_2	CO_2 + STEAM (VOL.%)
LOW	0	16
MODERATE	4	25 - 30
HIGH	9	42 - 47

to the furnace size.

Similar consideration of hearth area to eliminate fixed carbon from solids accounts for the earlier statement that RDF is a fuel that needs hearth area in a multiple hearth furnace while gas or oil burners do not.

Heat Balance Results

Presentation of Fuel Requirements and Energy Recovery Potential. In Figure 2, the fuel requirements (where any exist) and the heat available for recovery, are both plotted against the sludge characteristics expressed in terms of higher heating value per unit of water content. It is water that consumes heat in the furnace and fuel value of the organics that must provide the heat or be supplemented with purchased fuel. The ash hardly matters except that it carries water with it at any given feed solids content.

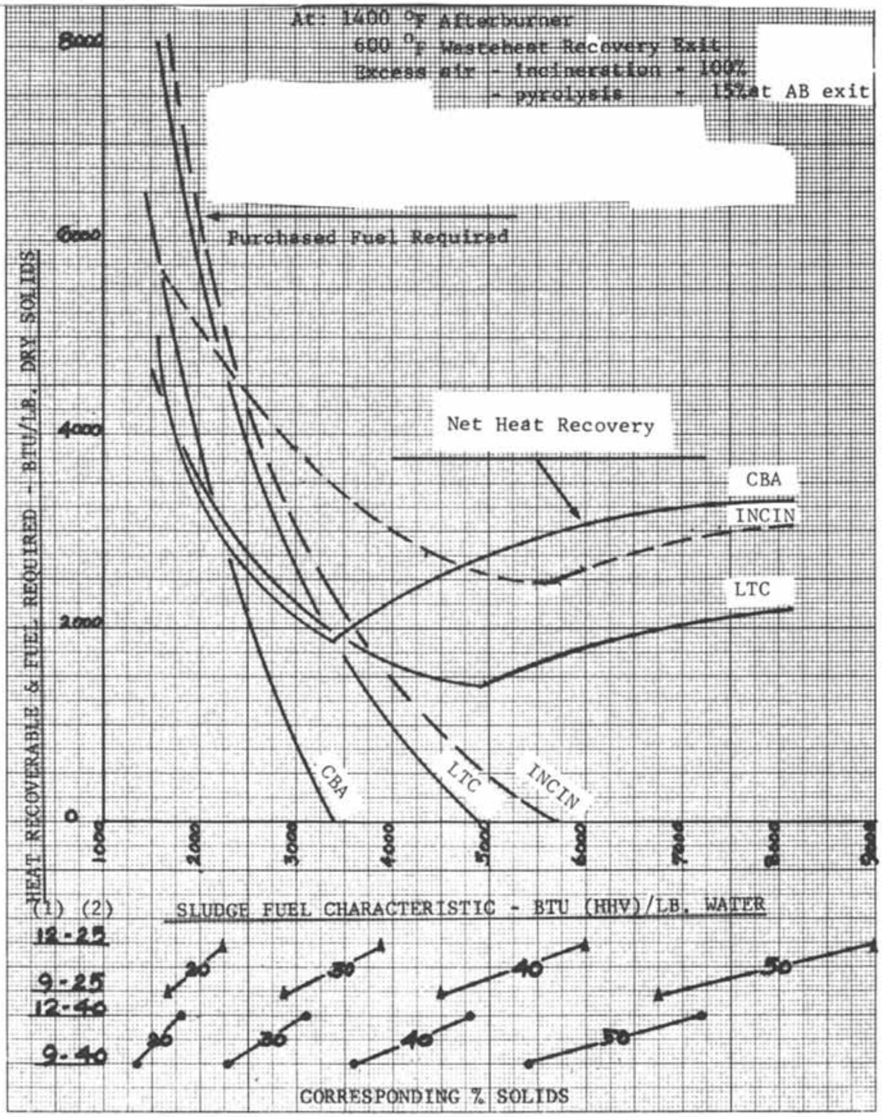
It is not possible to simply relate fuel requirement to filter cake moisture content without assuming a figure for sludge ash content and for heating value of the sludge organics. Ash content of sludge varies from 20 to 50%. Heating value per pound of organic content varies from 8,000 to 14,000 BTU. Variations over about half as wide a range as that just stated, occur within many plants from season to season if not from week to week.

The two supplementary scales on Figure 2 show the solids content to which sludge must be dewatered to provide the corresponding fuel characteristic for two different sludges. The range has been intentionally narrowed to keep the scale on the page.

Sludge at 40% solids shown to range from 3600 BTU/lb. water (if it has 40% ash content and the non-ash portion has a heating value of 9000 BTU/lb) to 6000 BTU/lb water (if it has 25% ash content and the non-ash portion has a heating value of 12,000 BTU/lb.

Impact of Cake Ash Content, Heating Value of Organics and Water Content. These two supplementary scales certainly emphasize the importance of considering sludge ash content and heating value of

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.



(1) This column BTU/lb. combustible X 1000 HHV
(2) Percent Ash

Figure 2. Heat recoverable and required with respect to sludge characteristics

sludge organic content when comparing different filter cake moisture contents as feed to a furnace. Oily or greasy sludges are frequently difficult to dewater to high solids content, but the high BTU per pound of organics may well result in the wetter cake being equally acceptable for pyrolysis or incineration. High dosages of lime and ferric salts for coagulation before filtering sometimes gives a drier cake than polymer flocculation but if the cake is not a lot drier, may increase cake ash content as to result in no improvement in "burnability".

The graph also makes it clear that small fluctuations in moisture content plus fluctuations in ash content and heating value of organics will normally cause considerable variation in the fuel characteristics of what is being fed to the pyrolyzer or incinerator. This will become important when considering storage and return of char to the furnace.

Conclusions from Figure 2, re: Pyrolysis. The graph makes it clear that:

A. Pyrolysis in the CBA mode increases recoverable heat by 10% as compared to incineration, only when considering sludge feed of more than 5,000 BTU higher heating value per pound of water. Otherwise recoverable heat is reduced 33%, because with lower heating value in the sludge feed the incineration base of comparison uses purchased fuels.

B. The desirable aspect of pyrolysis from an energy economy viewpoint, is reduction or elimination of requirement for auxiliary fuel.

C. Pyrolysis can be fuelless (including fuelless afterburner) with sludge dewatered to the range of 29 to 39% solids for most sludge compositions as compared to 42 to 52% solids required for incineration to be fuelless and provide a 1400 °F exhaust temperature.

Being able to dewater sludge to 42 to 52% solids range is quite uncommon, in fact the authors know of no commercial piece of equipment that could be depended upon to do this regularly on typical sludges, although perhaps the new membrane plate filter presses will get well into this range on some sludges. A range of 29 to 39% solids for dewatering both sludges is something one would very confidently expect from 3 kinds of dewatering equipment: recess tray filter presses, belt filter presses equipped with squeezing devices and most confidently membrane plate filter presses.

It seems apparent that raising the sludge fuel content per pound of water content from the results usually obtained by dewatering to 20 to 30% solids content is necessary to make pyrolysis a fuelless operation; and that a shift from incineration to pyrolysis is necessary to keep the extent of dewatering required down to a level that can be practically depended upon from normal equipment.

Raising sludge fuel content per pound of water content depends primarily on improved dewatering of the sludge since one has no control over the heating value of the organic content and little or no control over the ash content.

Improved Dewatering

Much is relatively new in this important area.

Flocculation. Polymer flocculants have become cost competitive with ferric-lime coagulation. Some sludges coagulate and dewater best with ferric and lime, some with one polymer, some with another. It is not easy to be sure the choice of flocculant made during a test period is still the best flocculant choice during subsequent weeks of operation when sludge characteristics may have varied. But increasingly systems are being designed for easy change from one flocculant to another.

The increased flexibility can result in better fuel characteristics in the sludge filter cake if due allowance is made for the effect of ferric and lime on sludge ash content, instead of seeking simply the highest solids content.

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

Filtering Equipment. The recessed plate filter press and belt filters that squeeze the cake, both introduced to the sewage treatment field within the past few years, have raised cake solids content from the 20 to 30% solids range to the 35 to 45% solids range.

Now interest is shifting to expandable membrane filter presses. This is a concept patented some years ago by Ciba-Geigy (8) and licensed to Moseley Rubber Company Limited of England. That company after several years of developing and improving fabrication methods for such plates is now introducing them to sewage works in England and in this country (9).

In these the compartments containing a partly or fully formed cake is reduced in volume, by expanding the face of the trays as shown in Figure 3. This squeezes the cake transversely whereas in the older recessed tray plates pressure is applied to the cake only by the force of new sludge coming in through the feed entry port. The older fixed plate presses therefore apply pressure only edgewise to the center of the cake and this pressure is soon lost to friction of the cake against the two adjacent filter cloths. The cake can be squeezed by the expandable membrane tray for any desired time period and a thin cake is squeezed as hard or harder than a thick cake.

The expandable membrane feature can be used to reduce initial capital cost of filter presses by reducing total cycle time. Typically, for the standard recess plate, 75% of the sludge volume enters the press in the first 25% of the press cycle after which sludge input rates drastically slow. By expanding the membranes with compressed air at this point, a given size press produces 75% of the normal cake weight and with squeezing time allowed may require only 30 to 35% of the normal cycle time. This in essence doubles press capacity.

Besides producing a drier cake in normal operation due to the transverse direction of the squeezing forces and the time for which squeezing forces can be applied, and the thinner cake that is produced, the membrane plate eliminates production of occasional "wetter than normal" cake. An incompletely filled press has full squeezing pressure

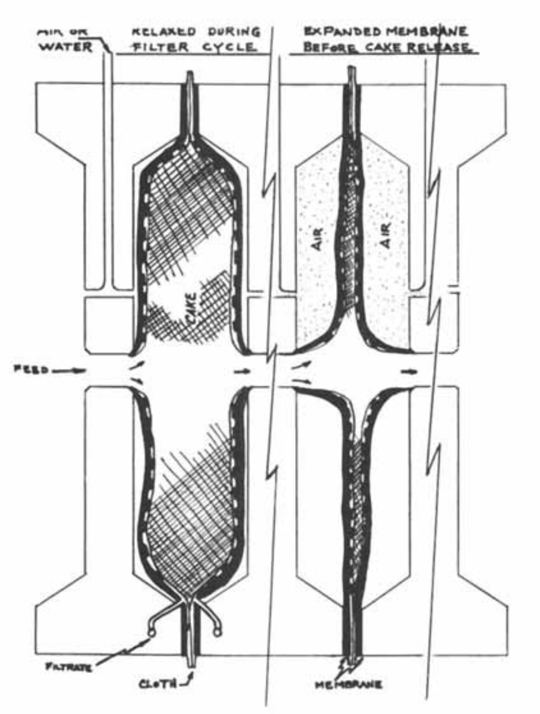


Figure 3. Flow scheme expandable membrane filter press

applied to each cake. Also an occasional upset in coagulation may be largely offset by slow initial filtration giving a thinner than normal cake during the time allowance for press filling but that thin cake will be squeezed dry while correction is made to coagulation. The cake can be profitably left to squeeze for as much cycle time as safety factor in the initial installation of the presses and any under-capacity of actual operation compared to planned operation may allow in the filter cycle.

Filtration Results with Membrane Plates. The performance of such membrane plates is described in a British publication (9) from which Table VI has been borrowed. Conditioned sludge from a single tank was pumped to a standard recessed plate press and simultaneously to a membrane plate press. Cakes were formed, squeezed and then additional sludge pumped in and squeezed, a procedure which accounts for there being a first and second feed time and a first and second squeeze time. Press capacity seems to be about doubled and comparative cake solids content increased from an average of 26.2% solids to 34.3% - a change from 2.8 lbs. water/lbs. solids to 1.9 lbs. water/lbs. solids.

Such membrane plates were purchased for the Nichols pilot plant press which uses 4 ft. by 4 ft. plates. The main object of their use was to facilitate cake preparation for pyrolysis tests. But when only 1 tray was available for the first such campaign, it was installed in a stack of regular rigid recessed trays and tried. When the diaphrams were expanded they squeezed the cake on either side of this one tray. Though some slight sludge movement out of the two squeezed compartments occurred, perhaps preventing full pressure development, the results in Table VII were obtained.

It is believed that it is now possible to depend on dewatering most sludges to 35 to 50% solids in a cost competitive system and obtain sludge cake that will generally exceed 5000 BTU available per pound of water contained.

This is an important part of the system for pyrolysis of sludge alone since the first aim after

TABLE VI

Table A — Large test press. 1 No. Recess Chamber 1220mm x 1220mm x 32mm Per chamber

				Output comparison — 2 shift system				
				Membrane Press		Recess Press		
Cake Dry Solids %	Dry Solids Applied kg	Feed Time mins	Maximum Feed Pressure kPa	Possible No. of Cycles per day	Total Dry Solids per day kg	Possible No. of Cycles per day	Total Dry Solids per day kg	Increase obtained by use of Membrane %
25.8	11.6	180	621	8	88	5	58	52
28.0	12.7	180	621	10	86	5	63	37
28.3	10.6	180	621	10	108	5	53	104
28.3	10.6	180	621	10	102	5	53	92
22.0	8.6	180	621	10	82	5	43	91
22.0	8.6	180	621	10	100	5	43	133
31.2	13.0	180	621	10	100	5	65	54
31.2	13.0	180	621	10	104	5	65	60
21.4	10.0	150	621	10	106	5	50	112
21.4	10.0	150	621	10	115	5	50	130
26.0	11.7	150	621	10	109	5	58	88
28.4	11.9	150	621	10	110	5	59	86

Table A — Brockhurst sludge. 1 No. Membrane Chamber 1130mm x 1075mm x 32mm

Test No	Cake Dry Solids %	Dry Solids Applied kg	Total Time mins	Feed Time mins		Squeeze Time mins		Maximum Feed Pressure kPa		Maximum Squeeze Pressure kPa		Cake Thickness mm
				1st	2nd	1st	2nd	1st	2nd	1st	2nd	
27	36.7	11.0	80	45	10	15	10	317	310	517	517	22
29	31.2	8.6	55	20	15	10	10	310	310	483	414	25
30	35.4	10.8	55	20	15	10	10	310	310	379	434	25
31	30.9	10.2	55	20	15	10	10	310	310	414	414	25
32	29.7	8.2	55	20	15	10	10	310	317	448	448	25
33	34.1	10.0	55	20	15	10	10	310	310	386	483	25
34	34.3	10.0	60	30	10	10	10	310	310	485	503	22
35	33.4	10.4	60	30	10	10	10	310	310	414	448	22
36	34.3	10.6	60	30	10	10	10	310	310	455	455	25
37	36.3	11.5	60	30	10	10	10	310	310	448	414	25
38	34.9	10.9	60	30	10	10	10	310	310	414	414	25
39	40.9	11.0	60	30	10	10	10	310	310	621	586	19

The output consideration is based on a day's operation at the Brockhurst Works ie a shift system with attendance hours 07.30 to 22.00h Monday to Friday.

621 kPa = 90 psi

TABLE VI-A

FILTER COMPARISON SUMMARY

STANDARD RECESS PLATES

VS.

MEMBRANE PLATES

TYPE	% SOLIDS	PER DAY	DRY SOLIDS (AVERAGE) PER DAY - KG
RECESS PLATES (STD.)	26.2	5	55
MEMBRANE PLATES	34.3	10	103

TABLE VII
NICHOLS PILOT PLANT
TEST RESULTS ON FILTER PRESS

PRESS RUN	CAKE SOLIDS - % UNCOMPRESSED CAKE	CAKE SOLIDS - % COMPRESSED CAKE
POLYMER CONDITIONED SLUDGE		
1	31.9	34.4
2	35.8	40.2
3	34.3	37.4
LIME-FERRIC CHLORIDE CONDITIONED SLUDGE		
19	29.8	45.7
20	43.5	46.8

thermal destruction of the organics must be to eliminate the use of purchased fuel.

Char as Supplemental Fuel.

Variations in sewage sludge dewaterability, in ash content and BTU per pound of organics, and occasional upsets in coagulant choice and application must be expected. Supplementation of the heating value of the sludge itself, by return of previously produced char is one way to overcome this problem.

For a 40 MGD wastewater treatment plant producing approximately 5000 pounds of sludge solids per hour, fuelless operation in the LTC or HTC mode to produce char would be recommended whenever the sludge has more than 5000 BTU per pound of water contained. This will be most if not all of the time, judging from the conversion scales at the bottom of the graph in Figure 2 and from the performance expected of the Moseley membrane plates. Assuming 30% ash in the feed and 12% yield of fixed carbon from organics, the char produced would consist of 600 pounds per hour of carbon plus 1500 pounds per hour of ash and be about 140 cubic feet per hour.

A 14,000 cubic foot storage bin for example would therefore hold about 100 hours production, or 60,000 pounds of carbon - 900 million BTU of heat value in easy to store and handleable form.

The furnace for pyrolysis could be sized at 1833 square feet of hearth area as indicated earlier in this paper, to burn away all of the fixed carbon from the normal feed rate. Self-sustaining operation would then be maintained through any deviations in sludge ash content, or heating value per pound of organics contained, or cake moisture content that reduced the fuel characteristic of the sludge to as little as 3400 BTU per pound of water contained.

Extreme high sludge ash content or low heating value of sludge organic content, reducing sludge heating value below 3900 BTU/lb., could be offset by the return of char from the storage bin. Low sludge availability in relation to heat demand could be met in the same way.

A temporary excessive sludge moisture content, for example change from 40% moisture to 25% moisture for 8 hours would require feeding the furnace half the usual solids rate. This 10,000 lbs/hr. wet cake feed would be pyrolyzed and the carbon burned away in 1200 square feet leaving 600 square feet for burning of supplemental char added from the storage bin, 360 pounds per hour of carbon or 5,000,000 BTU/hr. or 2,000 BTU/lb. of dry sludge solids fed could be added in this way; maintaining fuelless operation although reducing energy recovery from exhaust gases from 2750 BTU recovered per ton sludge solids to 2500 BTU recovered. These numbers are quickly worked out from the graphs and tables given earlier in this paper.

The char storage bin could also of course be used for stand-by fuel to keep the multiple hearth furnace hot when sludge is not available, or to provide start-up fuel.

Conclusions

Pyrolysis of sludge alone can be relied upon to eliminate all use of purchased fuels.

This requires the use of modern dewatering equipment, which now has become commercially available.

Heat recovery in the range of 2,000 to 3,000 BTU per pound of dry sludge solids seems easily feasible.

Energy recovery in the form of increased use of digesters to generate methane, while still avoiding a fuel cost to thermally destroy the organics in the resulting energy depleted sludge may in some cases be a preferred use.

Energy recovery as a char ash mixture to be treated for the elimination of heavy metals, converted to active carbon, or used as a fuel at other locations than the sewage treatment plant, may also develop as the preferred form.

It is believed that much work remains to be done in finding valuable uses, preferably in-plant, for the carbon ash mixture that can be produced from sewage sludge and for the heat available from the gases exhausting from the afterburner. Steady improvement in dewatering equipment and in thermal destruction energy efficiency will increase the availability of these products and thus bring about increasing efforts to utilize them.

The pyrolysis of sludge alone in conjunction with better dewatering practices, will eliminate all fuel cost; and generate enough steam from the exhaust gas to run the exhaust gas fan (the only major power consumer on the pyrolysis unit)and aeration blowers for the waste activated sludge process, by steam turbine. It will at the same time provide for the production of char, thus laying a base for eventual heavy metals recovery.

This should make it much more cost effective than heretofore to carry out the essential function of converting offensive and dangerous sludge organics to safe and inoffensive inorganics.

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April 18, 1978.

11

A Thermal Process for Energy Recovery from Agricultural Residues

RITCHIE D. MIKESELL, DONALD E. GARRETT, and DINH CO HOANG
Garrett Energy Research and Engineering Co., Inc., 911 Bryant Place, Ojai, CA 93023

In May, 1976, a one-year exploratory contract, E(04-3)-1241, was let by the U. S. Energy Research and Development Administration to Garrett Energy Research and Engineering Co., Inc. to study the thermal decomposition of biomass materials in a bench scale pilot plant of the multiple hearth reactor type. In October, 1977, this contract was extended by DOE for 18 months for operation in a larger, process development unit (PDU) pilot plant. This unit will have a six ton/day, four-foot diameter, four hearth reactor. The program to date has consisted of vacuum drying, direct contact drying, pyrolysis, the char-steam reaction, and combustion experiments. The results of these studies are reported in the following paper.

GERE Process

The proprietary Garrett Energy Research and Engineering Company biomass process is based upon a multiple-hearth furnace. This type of equipment has been successfully employed for continuous high temperature processing in the chemical and metallurgical industries for over 100 years. There are several vertically stacked compartments with a common central shaft. Rabble teeth are mounted on arms attached to the central shaft whose slow rotation imparts a positive mechanical motion to the solid material on each hearth. Downcomers through which the solid drops onto the hearth below are located alternately near the inner and outer periphery. They may have star valves for good sealing, and the conditions on each hearth may be optimized for its desired function. Since the solid is spread in a thin layer with constant raking and tumbling action, the residence time is adjustable, and efficient solid-gas heat- and mass-transfer are achieved. The multiple-hearth furnace is well suited for processing materials that are moist, fibrous, sticky, or susceptible to ash fusion at high temperature.

In the GERE process, Figure 1, the raw material may be first chopped and mechanically dewatered if necessary, and then conveyed

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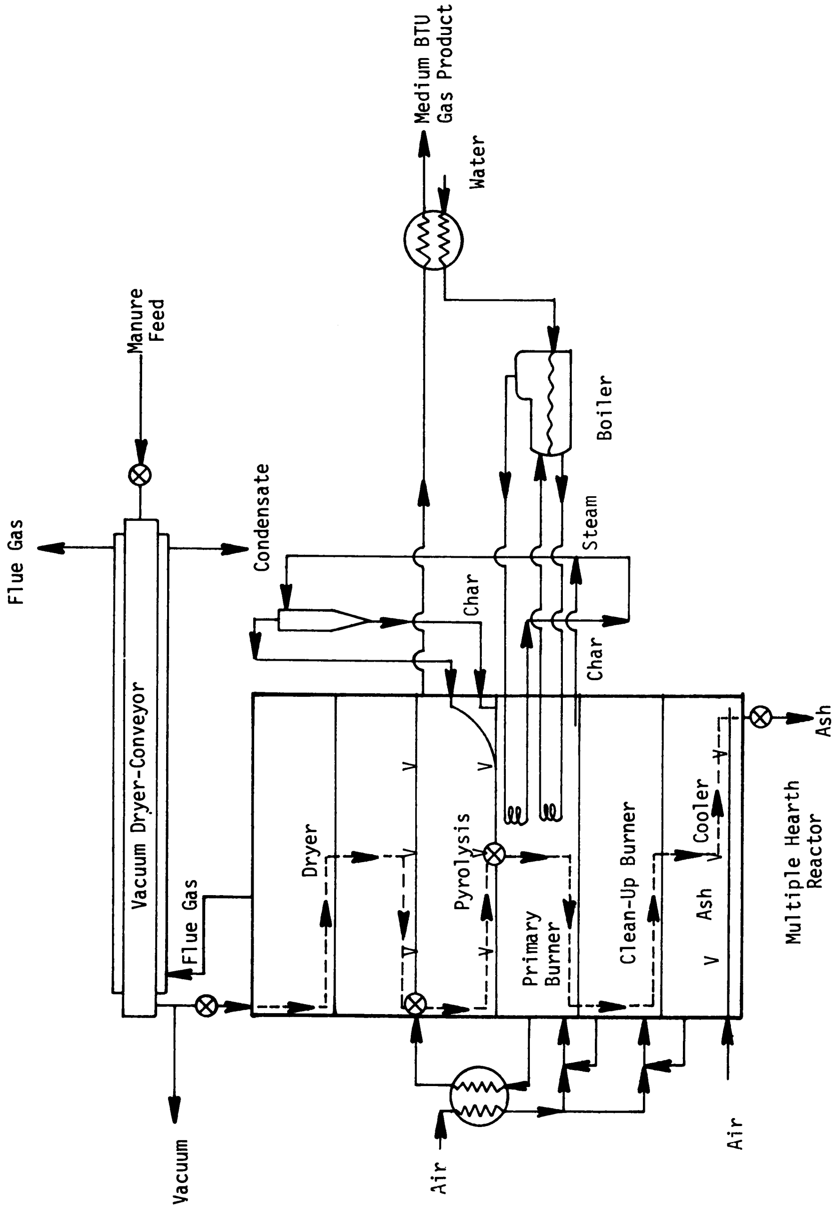

Figure 1. GERE multiple hearth pyrolysis process

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

to the top of the multiple-hearth reactor by means of a jacketed screw conveyor. The material is partially dried under vacuum by moist flue gas flowing countercurrently through the conveyor jacket and screw, and condensing inside. The conveyor has star valves at each end to maintain a good seal. Following the vacuum drying the remaining approximately half the water is removed by direct contact with hot flue gas on the first, or drying hearth. This water is then partially condensed in the vacuum screw feeder to provide multiple effect economy. Because biomass usually contains a great deal of moisture, it is very important that this moisture be removed as efficiently as possible.

The biomass, after drying to the desired moisture content, then falls to the pyrolysis hearth. Here it is heated by hot char from the steam-char reaction, and also by the hot gases accompanying the char. The char is introduced through an external steam lift in which synthesis gas is formed as the char reacts with superheated steam. The mixture of this synthesis gas and the gaseous products of pyrolysis (H_2, H_2O, CO, CO_2, CH_4, CnHm, H_2S, NH_3, COS) are cooled and condensed. The condensed tar is recycled to the pyrolysis zone.

Char from the pyrolysis hearth drops onto the combustion hearth where it is partially burned to generate steam and process heat. The hot flue gas preheats the incoming air in an external heat exchanger, and then goes to the drying section. Some of the hot char is reacted with steam in the steam lift and then sent to the pyrolysis zone, while an ash-bleed stream drops to the cleanup burner where it is completely burned. The ash then drops to the lowest hearth where it is cooled by incoming air and discharged from the reactor.

The gas leaving the condensate collector may either be utilized directly or sent to a scrubber in which H_2S, CO_2 and moisture are removed. This product gas may be piped for use at a nearby industrial plant, or blended into a natural gas pipeline.

Experimental

Biomass Feed. The GERE pilot plant work is done in Hanford, California, an agricultural community, with steer manure as the first pilot plant feed. The moisture content of fresh manure is about 80%, although the moisture content of piled manure ranges from about 10 wt % (1) to about 50 wt % (2) depending on the climate and length of storage. A typical ultimate analysis of bovine manure is given by Schlesinger, Sanner, and Wolfson (3) in Table I. For the pilot plant work, the moisture content of the feed manure was adjusted to 50 wt %.

Table I. Bovine Manure, Typical Ultimate Analysis

C	41.2 wt %
H	5.7
O	33.3
N	2.3
S	0.3
Ash	17.2
Total dry	100.0 wt %

Pilot Plant Equipment. During the first phase of this work a single-hearth reactor was used. Each process--direct contact drying, pyrolysis, water gas char gasification, and combustion--was studied in turn in this hearth. Figure 2 shows a schematic of the reactor; detailed description of the equipment is given in Reference (4).

Laboratory Experimentation. Thermogravimetric analysis of manure pyrolysis showed that over the temperature range 25-700°C, the reaction is slightly exothermic and that the use of Na_2CO_3 had little if any effect on the reaction rate. It was found that flue gas as hot as 180°C could be used to dry manure without burning it. Manure can be decomposed (pyrolyzed) at a temperature as low as 250°C, though the product gas at this temperature is mostly CO_2. If the tar produced in pyrolysis is itself pyrolyzed, more gas is produced. In particular, the hydrocarbon gas production is enhanced. Manure ash begins to fuse at about 950°C.

Manure drying experiments were first conducted in a laboratory oven. The data were correlated using the relationship

$$-\frac{\delta M}{\delta \tau} = 4.27\cdot M\cdot \exp -\left(\frac{2474}{T_S}\right) \qquad (1)$$

According to Reid, Prausnitz, and Sherwood (5), the exponential term is associated with the moisture diffusivity.

Hot air was passed through a small, well insulated, fixed bed of manure in order to dry it. Since the drying rate ought to depend upon the superficial gas velocity to the 0.8 power, and on the moisture diffusivity to the 0.56 power, the data were correlated as

$$-\frac{1}{M}\cdot\frac{\delta M}{\delta \tau} = 1.27\cdot\left(\frac{V_G}{150.8}\right)^{0.8}\cdot\exp -\left(\frac{1385}{T_S}\right) \qquad (2)$$

This relationship is consistent with Equation (1) and represents the laboratory data to within ± 20%. Using bed thicknesses comparable to those used in the pilot plant (about 3 cm) and superficial gas velocities close to those used in the pilot plant (about 150 cm/min), it was found that the gas and solids temperatures very rapidly came together. This was taken to mean that the product of the heat transfer coefficient and the heat transfer area,

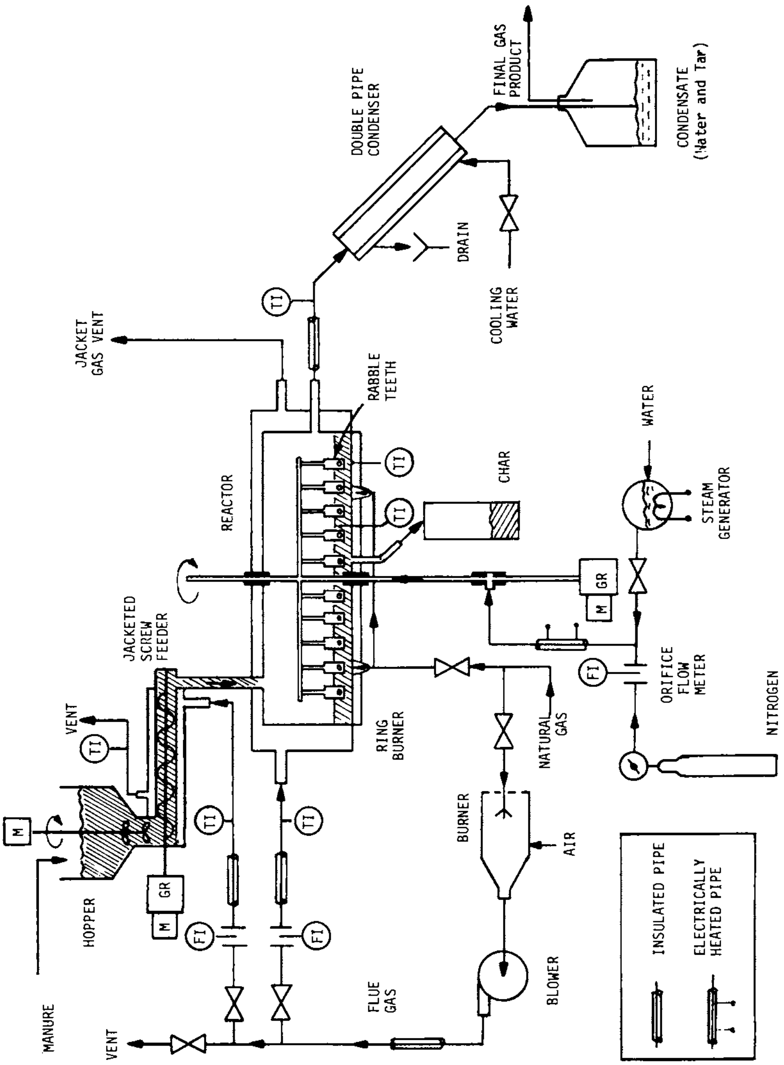

Figure 2. GERE single hearth bench-scale pilot plant unit

UA, for the bed is essentially infinite. Equation (2) together with

$$\text{Thermal Equilibrium} \qquad T_G = T_S \qquad \rule{3cm}{0.4pt} \tag{3}$$

$$\text{Heat Balance} \qquad (C_P)_S \cdot \frac{\delta T_S}{\delta\tau} - \lambda_V \cdot \frac{\delta M}{\delta\tau} = (C_P)_G \cdot \{(T_G)_{in} - T_G\} \tag{4}$$

$$\text{Moisture Balance} \qquad -M_S \cdot \frac{\delta M}{\delta\tau} = W_G\{H - H_{in}\} \qquad \rule{3cm}{0.4pt} \tag{5}$$

make a system of relationships that describe the drying process. The problem was programmed on an HP-67 and numerically integrated. The model predicts drying rates about 20% higher than experienced in the laboratory. This probably reflects the fact that the model does not consider the heat losses.

Pilot Plant Experimentation

Vacuum Drying. The small jacketed screw feeder was used to measure heat transfer coefficients. The screw in this feeder was not hollow, no vacuum was used, and saturated steam was fed to the jacket. The data are shown in Table II.

Table II. Screw Conveyor Heat Transfer Data for Vacuum Drying Manure

Feed Moisture wt %	Screw Speed RPM	Solids Residence Time, τ_R min.	Overall Heat Transfer Co-efficent, U cal/min cm^2 ^oC	$\frac{U.\sqrt{\pi\alpha\tau_R}}{k}$ dimensionless
37.34	1	25.7	$1.01(10^{-2})$	1.23
37.34	2	15.0	2.00	1.86
37.34	3	9.4	2.55	1.87
50.0	1	50.4	0.55	0.50
50.0	2	27.5	1.61	1.08
50.0	3	15.1	1.81	0.90

Direct Contact Drying. Moist manure was fed to the single hearth through the screw feeder. The rabble teeth swept the material across the baseplate from the center to the outside where it fell through a single hole into a receiver. The baseplate contained 1/8 in. holes spaced about 2-1/2 in. apart, and hot flue gas was passed upward through these holes and then through the moving bed of manure. Long solids residence times were obtained by using low rabble arm speeds and by turning the rabble teeth almost parallel to their direction of motion. The moisture contents of the feed and product, the solids feed rate, the holdup volume, the inlet and outlet solids temperatures, the rabble arm rotation rate, the rabble tooth angle, and the hot flue gas inlet and outlet temperatures and humidities and its flow rate were measured for each run.

Equations (2) through (5) describe direct contact drying in the pilot plant as well as in a fixed bed, so long as it is understood that time, τ, in the model corresponds to $\int dr/u_r$ in the pilot plant and that the outlet gas temperature in the pilot plant is the same as

$$\frac{1}{\tau_R} \cdot \int_0^{\tau_R} T_G \, d\tau$$

in the model. Pilot plant data for three experimental runs are compared with results of numerically integrating Equations (2) through (5) in Table III.

The model does not consider heat losses, and so the predicted outlet solids and gas temperatures are higher than measured, and the predicted outlet solids moisture content is lower. The measured outlet solids temperatures were obtained by immersing a thermometer in the solids after they had fallen from the reactor. As the solids accumulated in the receiving bag, they probably cooled down. Thus, the measured outlet solids temperatures in Table III seem unreasonably low. In the future this temperature will be measured inside the reactor by a thermocouple attached to one of the rabble teeth.

Pyrolysis. In these experiments, manure samples having different moisture contents were pyrolyzed at moderate (600-750°C) temperatures. The data show that a substantial part of the carbon was gasified by the steam. Here, then, pyrolysis and steam gasification occurred together in the same reactor at only moderate temperatures.

Initially the reactor was empty while the baseplate was heated from below by natural gas burners. Argon was used to purge the reactor. At time zero the feeder was turned on, and after a few minutes the argon flow was turned off and the temperatures stabilized. At 10-minute intervals, the pyrolysis gas flow rate was measured, pyrolysis gas samples were taken, the baseplate temperature, the solids temperature near the inlet, and the solids temperature near the outlet were recorded. The solids temperatures were measured by thermocouples attached to the rabble teeth. These temperature measurements showed that the solids are very nearly isothermal as they are swept across the reactor floor. The pyrolysis gas samples were analyzed by a Beckman Gas Chromatograph 2A. At the end of the experiment the remaining char was cooled under argon. The amount of char remaining that can be swept out by the rabble teeth (the dynamic holdup), and finally the rest of the char in the reactor (the static holdup) was removed and weighed. The moisture content, the volatile content (750°C), the residual carbon, and the residual ash of the product char, the

Table III. Direct Contact Drying of Manure: Comparison of Model With Pilot Plant Data

M_S	gm dry solids	1112		1452		589	
V_G	cm/min	138.3		138.3		138.3	
W_G/M_S	gm dry gas/(min)(gm dry solids)	.4021		.3079		.7592	
$(T_G)_{in}$	$^{\circ}$K	417°		421°		423°	
H_{in}	gm H_2O/gm dry gas	.025		.033		.025	
M_{in}	gm H_2O/gm dry solids	.6234		.6234		.6234	
τ_R	min	27.84		19.04		7.39	
		Calc.	Meas.	Calc.	Meas.	Calc.	Meas.
T_S	$^{\circ}$K	386°	320°	353°	320°	380°	319°
M	gm H_2O/gm dry solids	.310	.390	.416	.507	.500	.545
$\frac{1}{\tau_R} \cdot \int_0^{\tau_R} T_G d\tau$	$^{\circ}$K	358°	348°	344°	332°	376°	340°

dynamic holdup char, and the static holdup char were measured using the laboratory muffle furnace. The steam and tar in the pyrolysis gas were condensed and weighed. Later these samples of condensate will be analyzed for their water contents. The condensed tar can, of course, be re-pyrolyzed to produce a hydrocarbon-rich gas.

Some representative data are shown in Table IV. These data are for the pyrolysis of two manure samples at about 650°C. The moisture content of the first sample was 43.2 wt %; the moisture content of the second was 5.2 wt %.

In these pyrolysis experiments, Reactions (6), (7), and (8) also occurred.

$$C + H_2O \rightarrow H_2 + CO \qquad (6)$$

$$CO + H_2O \rightleftarrows H_2 + CO_2 \qquad (7)$$

$$CH_4 + H_2O \rightleftarrows 3H_2 + CO \qquad (8)$$

Carbon and methane can also be oxidized by CO_2 at these temperatures, but probably at slower rates. In the pyrolysis of Sample 1, the steam content of the pyrolysis gas in the reactor was quite high, and so the reactions above were driven further to the right. Reactions (7) and (8) are far from equilibrium for the data of Table IV. These reactions as well as Reaction (6) are probably almost, but not quite, frozen at these temperatures. Even if the feed manure were completely dry, Reactions (6) through (8) would occur to some extent, because H_2O is one of the products of pyrolysis.

Water Gas Reactions. Nitrogen was used to purge the reactor as the baseplate was heated by natural gas burners. Pyrolysis char and superheated steam were fed to the reactor and the nitrogen flow was turned off. At 15-minute intervals gas samples were taken. The condensate, which contained very little tar, was collected and weighed after each run. At the end of the run, nitrogen was passed through the reactor until it had cooled. The reactor holdup, dynamic and static together, was weighed, but this char was not analyzed. The ultimate analyses of the inlet and product chars were determined, but the volatiles (750°C) and residual carbon were not. The baseplate temperature, the gas temperature, and the temperature of the char as it fell from the reactor were measured. In the future the solids temperature will be measured by a thermocouple attached to one of the rabble teeth. In Table V the solids temperatures were probably in the 600-700° range.

These runs lasted only about as long as one solids residence time, so the char remaining in the reactor at the end of each experiment is important to the material balance. In the next set of experiments the dynamic and static holdup chars will be weighed and analyzed.

Table IV. Pyrolysis of Moist Manure: Pilot Plant Data

Sample 1. Solids 635°C; Pyrolysis gas 437°C; 45 min solids residence time.

	Feed 3170 gm/hr	Product Char 390 gm/hr
H_2O	.4316 wt. fract.	.018 wt. fract.
Volatile (750°C)	.2353	.159
Carbon	.0954	.183
Ash	.2377	.640
	1.000 wt. fract.	1.000 wt. fract.

	Dynamic Holdup 295 gm	Static Holdup 343 gm
H_2O	.016 wt. fract.	.015 wt. fract.
Volatile (750°C)	.107	.090
Carbon	.174	.146
Ash	.703	.749
	1.000 wt. fract.	1.000 wt. fract.

Condensate (H_2O + Tar) 1304 gm/hr

Gas 772 gm/hr

H_2	.471 vol. fract.	37.1 gm/hr	.035 gm/gm DAF
CO	.112	123.6	.118
CO_2	.308	533.1	.509
CH_4	.089	56.1	.054
C_2H_4	.015	16.5	.016
C_2H_6	.005	5.9	.006
	1.000 vol. fract.	772.3 gm/hr	.738 gm/gm DAF

Overall mass balance closure	97.9%
Overall ash balance closure	94.7%
Carbon (750°C) gasified	42.9%
Pyrolysis gas heating value (low) (including CO_2 but not H_2O)	272.5 Btu/SCF

Table IV. Pyrolysis of Moist Manure: Pilot Plant Data (Cont.)

Sample 2. Solids 657°C; Pyrolysis gas 460°C; 23 min solids residence time.

	Feed 2700 gm/hr	Product Char 600 gm/hr
H_2O	.052 wt. fract.	.011 wt. fract.
Volatile (750°C)	.385	.113
Carbon	.2262	.304
Ash	.3368	.572
	1.0000 wt. fract.	1.000 wt. fract.

	Dynamic Holdup 230 gm	Static Holdup 595 gm
H_2O	.012 wt. fract.	.009 wt. fract.
Volatile (750°C)	.073	.063
Carbon	.270	.233
Ash	.645	.695
	1.000 wt. fract.	1.000 wt. fract.

Condensate (H_2O + Tar) 466 gm/hr

Gas 875 gm/hr

H_2	.352 vol. fract.	27.6 gm/hr	.017 gm/gm DAF
CO	.177	194.3	.118
CO_2	.309	533.1	.323
CH_4	.126	79.2	.048
C_2H_4	.021	23.4	.014
C_2H_6	.015	17.7	.011
	1.000 vol. fract.	875.3 gm/hr	.531 gm/gm DAF

Overall mass balance closure	97.6%
Overall ash balance closure	99.5%
Carbon (750°C) gasified	37.3%
Pyrolysis gas heating value (low) (including CO_2 but not H_2O)	319.2 Btu/SCF

Table V. Steam-Carbon Reaction: Pilot Plant Data

Run 6. Baseplate 810°C; Gas outlet 425°C; 115 min solids residence time.

	C	H	O	Ash	Total
Char In, kg/hr	.250	.025	.148	.497	.920
Steam Consumed, kg/hr	-	.038	.302	-	.340
Total, kg/hr	.250	.063	.450	.497	1.260
Output Gas					
CO_2, kg/hr	.074	-	.197	-	.271
H_2, kg/hr	-	.026	-	-	.026
CO, kg/hr	.013	-	.018	-	.031
CH_4, kg/hr	.003	.001	-	-	.004
Char Out, kg/hr	.127	.014	.024	.497	.662
Total, kg/hr	.217	.041	.239	.497	.994

Overall mass balance closure	78.9%
Overall carbon balance closure	86.8%
Carbon gasified	36.0%

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

Table V. Steam-Carbon Reaction: Pilot Plant Data (Cont.)

Run 7. Baseplate 860°C; Gas outlet 415°C; 90 min solids residence time.

	C	H	O	Ash	Total
Char In, kg/hr	.344	.035	.204	.685	1.268
Steam Consumed, kg/hr	-	.052	.418	-	.470
Total, kg/hr	.344	.087	.622	.685	1.738
Output Gas					
CO_2, kg/hr	.094	-	.250	-	.344
H_2, kg/hr	-	.023	-	-	.023
CO, kg/hr	.022	-	.029	-	.051
CH_4, kg/hr	.003	.001	-	-	.004
Char Out, kg/hr	.148	.010	.042	.685	.885
Total, kg/hr	.267	.034	.321	.685	1.307

Overall mass balance closure	75.2%
Overall carbon balance closure	77.6%
Carbon gasified	34.6%

Char Combustion. Char from the water gas experiments was fed to the reactor. Combustion air was pushed through the base-plate and through the hollow rabble teeth. A 2 cm ash layer was maintained on the reactor floor. The reactor was not lined with fire brick. Typical data are shown in Table VI.

Table VI. Char Combustion: Pilot Plant Data

Char feed rate	67.7 gm/min
Carbon feed rate	2.10 gm mol/min
Air flow	11 SCFM
O_2 flow	3.70 gm mol/min
O_2/C ratio in feed	1.76
Carbon burned	63.1%

The heat losses from the reactor were very large. In the work to follow, the combustion chamber will be well insulated and the combustion efficiencies should be higher.

Commercial Biomass Processing Plant Based Upon the GERE Process

Plant Sizing. It is assumed that each pyrolysis plant would serve an individual feedlot (or perhaps a group of feedlots concentrated in a sufficiently small area that they can conveniently use the same manure repository site). To factually determine a reasonable plant size it is necessary to review statistical data on feedlot sizes, as reported in Reference 4. Based upon this data, 2 ton/day (dry basis) plant for small feedlots (730 head @ 5.5 lb. per head), 100 ton/day (dry basis) plants for large feedlots (36,000 head), or very large plants of 250,000 t/yr. serving several feedlots could all be considered. However, the 2 ton/day plant proved to be prohibitively expensive in a preliminary analysis, so detailed cost analyses were not considered justified and only the larger plant sizes were reviewed. In each case a medium Btu gas was the only product considered.

Economic Analysis. A detailed process design and economic analysis based upon the data from this study was then made. It indicated that the capital cost for a 100 t/d (on a dry feed basis) manure processing plant would be about $1.9 million, and the cost of gas $2.84/MM Btu after depreciation and a 10% pretax profit. 1.7 MM SCF/day of about 500 Btu/cu.ft. gas would be generated. The break-even gas price would be $2.1/MMBtu. The plant would have a small ash disposal problem, but the ash may have some fertilizer value so no charge was made against it. The plant would generate all of its own heat and power, and have a net energy recovery of greater than 70% from the original feed material. The capital investment summary is given in Table VII, and the operating cost in Table VIII.

The proposed process flow-sheet is shown in Figure 3. It assumes that the biomass, in this case manure, is delivered to the plant fairly regularly, and dumped by the producer on a concrete storage pad. As described in an earlier section (GERE process), the manure is dried, pyrolyzed, and then burned to produce a medium-Btu gas. The operating conditions in each hearth of the reactor are indicated in Figure 3.

The pyrolysis and water gas product passes to a quench tank and then to CO_2, H_2S, and water removal equipment. A clean, dry, medium Btu gas (400-500 Btu/cu.ft.) is produced for direct sale or use as a synthesis gas. Any condensate from the quench tank is reinjected into the pyrolysis section to be cracked into gas. Very thorough heat exchange is practiced on all entering and exiting streams. The electricity required for the plant is generated internally from the high pressure steam.

Literature sources are used for all cost estimates shown in Table VII, except the multiple hearth furnace, which is a quotation from a manufacturer based upon past installations of essentially turn-key plants for sewage sludge incinerators, including gas cleanup equipment, elevators, feeders, instrumentation, etc.

Some of the assumptions used in sizing the equipment are: 1) the heating value for dry manure is 6240 Btu/lb; 2) the plant operates for 300 days/yr; 3) the total net energy recovery is 70%. The yearly energy production for the 100 ton/day plant is: (6240 Btu/lb) x (2000 lb/ton) x (100 tons/day) x (300 operating days/yr) x (0.7) = 2.62×10^5 MMBtu/yr, or 1.7 MM SCF/day of 500 Btu gas.

A complete amine-glycol gas purification system was estimated to cost $150,000 as a skid-mounted package, which includes an absorption tower, stripper-regenerator column, reboiler, two coolers and pumps, instrumentation and controls. If the presence of carbonyl sulfide (COS) or carbon disulfide (CS_2) becomes a problem, diethanolamine could replace monoethanolamine. If the moisture content of the product gas exceeds the specified levels, an additional drying column would be required.

An additional cost analysis has been made comparing the GERE process with a similar pyrolysis process studied by others. Their "best" pyrolysis case (Purox) was used, and the GERE process scaled from 100 t/d to 250,000 t/yr by means of the 0.6 power ratio. It is seen that the GERE process produces gas for $2.06/MM Btu, compared to their $6 or $7/MM Btu. The difference is due primarily to their much higher plant capital investment (their oxygen plant, front end drying facility, much more complex reactor, etc.), and the purchase of outside utilities (power). At either plant size the GERE process looks to be especially attractive.

Conclusion

Biomass can obviously be processed to recover energy, but

Table VII. Capital Investment, 100 T/day (dry basis) Manure Biomass Plant

Flow Chart Symbol		Direct Equipment Cost
CP-1	Concrete Storage Pad, 625m^2 (25m x 25m), 15cm thick with 15cm sub-base. 20cm x 2m high walls on 3 sides.	$ 25,000
FEL	Front End Loader	40,000
R-1	Ramp with Guard Walls	15,000
H-1	Feed Hopper, 12m^3 with Grizzly; stainless steel or type 4 concrete	15,000
LB-1	Lump Breaker	20,000
SV-1	Rotary Valve, 10" x 48", teflon coated rotor	5,000
C-1	Vacuum Dryer-Screw Conveyor, 14" dia. x 60 ft., jacketed shell and flights	66,000
SV-2	Rotary Valve	5,000
E-1	Condenser, 250 sq.ft., carbon steel	5,000
E-2	Heat Exchanger, Cooler for quench fluid, 100 ft^2, carbon steel	4,000
P-1	Vacuum Pump, 50m^3/hr at 350mm Hg absolute	5,000
QV-1	Quench Column, 4' dia. and settling drum, carbon steel construction	10,000
	Total Pyrolysis Equipment Cost	$215,000
	Factored Costs of Installed Equipment; 215%, including service facilities, instrumentation and controls, piping, electrical, land, yard improvements, engineering and construction overhead, contingencies	462,000
MHF	Multiple Hearth Furnace, 18' o.d. - 12 hearths (cost includes installation and all other costs)	960,000

Table VII. Capital Investment, 100 T/day (dry basis) Manure Biomass Plant (Cont.)

Total Capital Investment for Pyrolysis (Gasification) Plant	$ 1,637,000
Gas Purification Equipment	
Reciprocating compressor, 2000 SCFM, 100 psig discharge pressure, complete with 250 HP motor drive, intercooler, other auxiliary equipment, installation	100,000
Amine Glycol gas purification system for H_2S, CO_2, and H_2O removal, complete as a package on skids with an absorption column, stripper-regenerator column, reboiler, two coolers, instrumentation and control, carbon steel construction	150,000
Total	$ 250,000
TOTAL INITIAL INVESTMENT	$ 1,890,000

Table VIII. Annual Operating Costs, 100 t/d (dry basis) Manure Biomass Plant

Capital Cost: $1.89MM

	$/yr	$/MMBtu
Labor		
8 operators (2 per shift), 24 hrs/day, 365 days/yr, $8/hr	140,000	
2 mechanics at $20,000/yr	40,000	0.802
1 superintendent	40,000	
Plant overhead cost, 25% of labor and supervision	52,000	0.198
Outside maintenance materials, 4% of initial investment	76,000	
Outside maintenance labor, 1.5% of total initial investment	28,000	0.397
Insurance, 2% of total initial investment	37,000	0.141
Return on investment (10%) and depreciation (7%), 17% of total initial investment	320,000	1.221
Utilities (other than self-generated), power, water, etc.	20,000	0.076
Total if manure is free	$743,000/yr	2.835
Manure @ $7/ton (dry basis)		0.801
TOTAL		$ 3.636/MMBtu

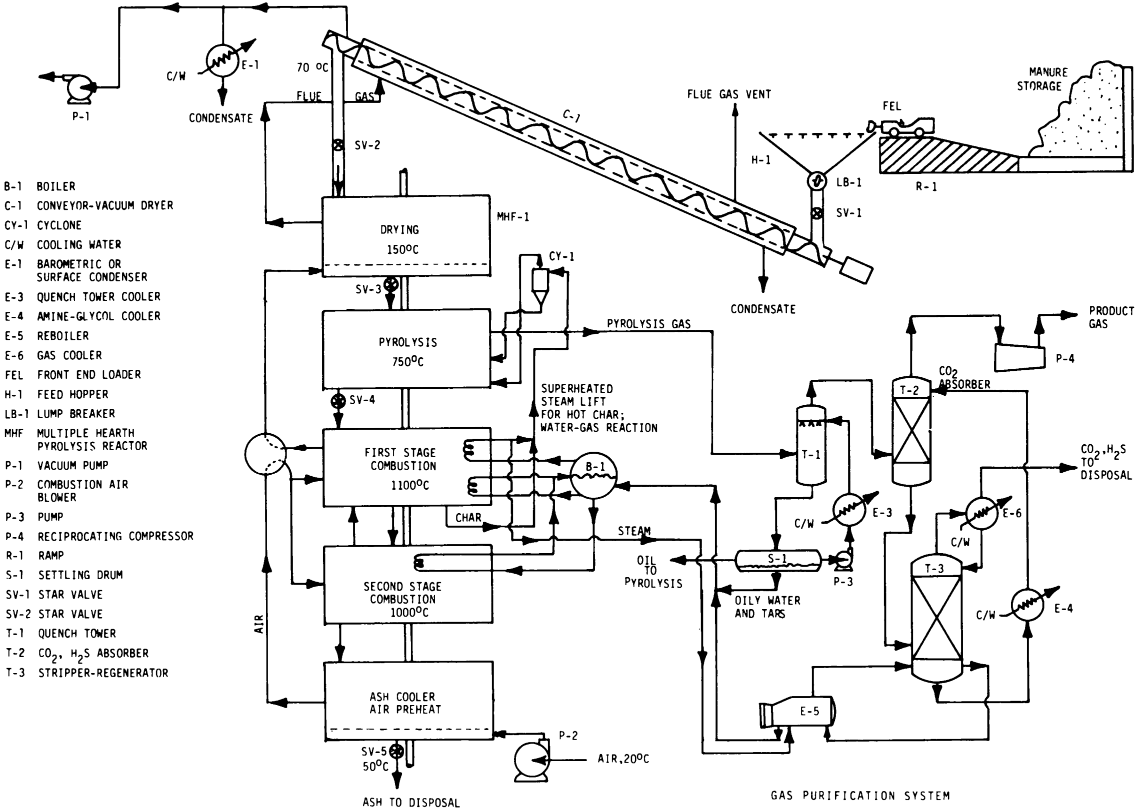

Figure 3. GERE projected commercial-scale biomass process

there appears to be no existing technology available that will allow an economical recovery. Nevertheless, the potential does exist for a correctly designed and engineered process which solves the problems of high moisture content, difficult feedstock handling, and generally low tonnage biomass availability at any one location. The GERE process shows promise in economically achieving this potential. The bench scale pilot plant study program indicates a technical feasibility for simple, low cost production of medium-Btu gas from biomass in modest sized plants.

Nomenclature

$(C_p)_G$ = specific heat of the gas, cal/^oC gm dry gas. A function of H.
$(C_p)_S$ = specific heat of the solids, cal/^oC gm dry solids. A function of M.
H = humidity of the outlet gas, gm H_2O/gm dry gas
H_{in} = humidity of the inlet gas, gm H_2O/gm dry gas
k = thermal conductivity, cal/cm min ^oC
M = moisture content of solids, gm H_2O/gm dry solids
M_{in} = moisture content of inlet solids, gm H_2O/gm dry solids
M_S = dry weight of fixed bed, gm dry solids
r = radius measured from the center of the rotating shaft, cm
T_G = temperature of the outlet gas, ^oK
$(T_G)_{in}$ = temperature of the inlet gas, ^oK
T_S = temperature of the solids, ^oK
u_r = local radial solids velocity across the baseplate, cm/min
U = overall heat transfer coefficient, cal/min cm^2 ^oC
V_G = superficial gas velocity, cm/min
W_G = gas mass flow, gm dry gas/min

α = thermal diffusivity, cm^2/min
λ_v = latent heat of evaporation, cal/gm water evaporated
τ = time, min
τ_R = solids residence time, min

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MARCH 3, 1978.

12

Activated Carbon and Fuel from Sewage Solids

JOHN E. SIGLER
2839 Tabago Place, Costa Mesa, CA 92626

Under a NASA development program for a source of carbon for insulation, Marshall Humphrey of Jet Propulsion Labs, (1) Pasadena, Ca., conceived a new process for treating wastewater. After bench testing, a trailer mounted unit was brought to the County Sanitation Districts of Orange County, California. After testing this 10,000 gallon per day unit plus some tests on reactors in Belle Meade, New Jersey, Springfield, Ohio and Louisville, Kentucky, the Sanitation Districts decided the system had enough promise to justify a pilot plant scale test.

Much of this justification was the elimination of a sludge disposal problem with which all present wastewater treatment systems are confronted. With the approval of the California State Water Resources Control Board, the Environmental Protection Agency, and the Sanitation Districts Board of Directors, a pilot plant was built and operated to obtain data for design of a full scale plant.

The one million gallon a day (MGD) size was considered an intermediate step between the .01 MGD trailer mounted unit and the full plant size of 175 MGD.

Process Description

Hydraulics.

The treatment system is a two-stage countercurrent adsorption process using activated carbon produced from the sewage solids.

The flow diagram in Fig. 3 illustrates the

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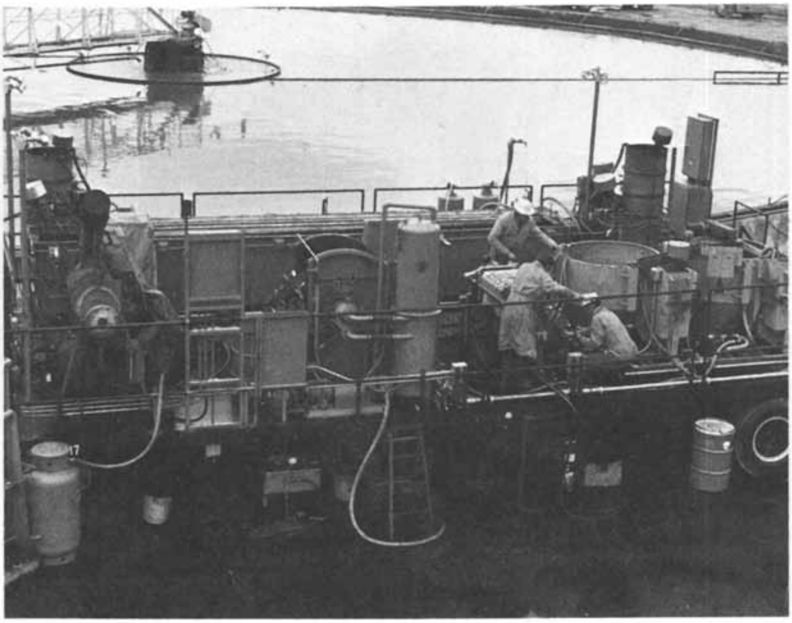

Figure 1. A 10,000 gallon/day demonstration plant—trailer mounted

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

Figure 2. The 1 MGD pilot plant

counter-current movement of the carbon with respect to the wastewater. Fresh carbon is added before the secondary clarifier where the action is essentially adsorption of dissolved organics on the carbon. The partially used carbon is removed from the secondary clarifier and mixed with the incoming wastewater. In the primary stage the action of the carbon is as a settling aid but some adsorption also occurs. Chemical additions of Ferric Cholorde and polymers were used to aid the sedimentation in the clarifiers.

A sand filter was installed on the effluent but the main thrust of the work was directed toward obtaining a satisfactory effluent from the secondary clarifier.

Mixing of the carbon with the wastewater before each clarifier was accomplished in two steps, a 2 minute fast mix and a 15 minute slow mix.

Solids Dewatering and Drying.

While the solids removed from the secondary stage were returned to the primary stage, solids from the primary stage were removed from the hydrualic system. These solids are of a mixture of activated carbon and sewage solids in about a 5 to 3 ratio.

The sludge solids removed from the clarifier were about 10% dry solids. Storing and decanting increased the solids content to about 15% before feeding to the filter press. The filter press regularly produced cakes which were greater than 50% solids content.

Further drying was accomplished in a flash dryer which can produce solids with only 2 or 3% moisture, but the product is dusty and fluffy. At 10 to 15% moisture content, the product is in a better condition for feeding to the reactor.

Reactor.

The reactor used was an indirectly fired, rotary kiln known as a calciner. The tube was 30 inches in diameter and had a hot zone of 30 feet. This is a standard unit as manufactured by Combustion Engineering except the tube was of $\frac{1}{4}$ inch Inconel 617 plate rolled and welded to shape.

Fifteen burners, divided into 3 zones for temperature control, were used to supply the heat to the reactor. The most common temperature used was 1600°F in all three zones but some runs were made with a temperature profile of 1100-1350-1500°F from feed to product end.

Feed was deposited well into the first zone by a screw feeder which also maintained a gas seal. Product discharged from the tube into an end bell and then through a rotary valve which maintained the seal on the discharge end.

Three things were to be accomplished in the reactor.

1. Complete the drying of the feed.
2. Pyrolyze the sewage solids.
3. Activate the carbon with superheated steam.

Superheated steam was fed into the last zone of the tube to accomplish the activation and reactivation of the carbon. It was generally accepted that the pyrolysis gas was evolved in the center section of the reactor while the drying was accomplished in the feed end zone.

Naturally, air was excluded from the reactor tube by the use of rotary seals connected to bellows on each end. All gas was taken from the reactor at the feed end and washed in a spray scrubber to remove any particles carried from the reactor.

It was originally intended that this gas be used to fire the flash dryer. However, a few days operation proved the flash dryer could operate using the hot stack gases from the reactor. The next logical place to use this gas was the reactor firebox but the pilot plant was not plumbed to allow this usage.

After cleaning, the gas was compressed, metered, analyzed, and flared.

Misc. Equipment.

An afterburner train consisting of an afterburner, pre-cooler, venturi scrubber, and fan were installed on the exhaust line from the flash dryer. However, the venturi scrubber performed well enough to meet the air quality standards so the afterburner and pre-cooler were not operated.

Discharge of dry carbon from the reactor is conveyed to a quench tank where it is cooled and slurried with water for transporting back to the carbon feed tanks.

Pyrolysis Reactor

Reactor Operating Condition.

While numerous profiles were used throughout the pilot plant operation, the most runs were made with the temperatures controlled around 1600°F in all three zones of the firebox. This allows the reactor to establish its own lengths necessary to accomplish the drying, pyrolysis and activation.

A rotational speed of 1.6 rpm gave a residence time in the reactor of about 40 minutes.

Steam was superheated externally to about 900°F and fed at rates of 20, 50, and 75 pounds per hour.

Description of Feed to Reactor.

The reactor feed was a dried cake which was about 5 parts of used reactor product and 3 parts of sewage solids. The feeds had moisture contents of 1 to 6%.

Analysis of Feeds on Dry Basis	
Ash	29-45%
Volatile Organics	12-21%
Fixed Carbon	37-58%

Feed rates were varied from 220 to 480 pounds of dry feed per hour.

Reactor Products.

The dry solids discharged from the reactor worked up to an ash content around 50% in the last runs. This ash content is necessarily a function of the ash losses in the system as a whole as well as the reactor operating conditions.

The activity of the carbon, as measured by the iodine number, finally stabalized in the range of 300 to 350 mg of iodine per gram of carbon.

A better measure of activity of the carbon for use on this process is an equilibrium curve where the removal of the actual contaminants is measured.

Figure 4 is a plot of several of these isotherms comparing reactor products with commercial carbons. The carbons produced in 1976 tended to match Nuchar while the latest carbons of Aug. 77 matched Calgon PC but were not as good as Calgon RB in adsorbing TOC from sewages.

After water washing, pyrolysis gas had the following analysis:

Constituent	% V/V
H_2	38-46

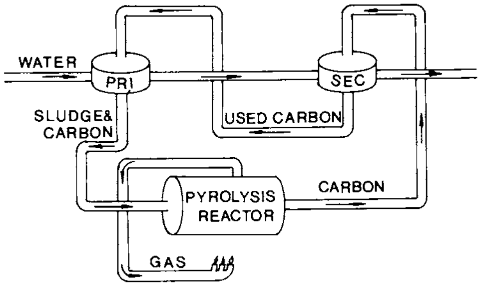

Figure 3. Flow schematic of activated carbon treatment system

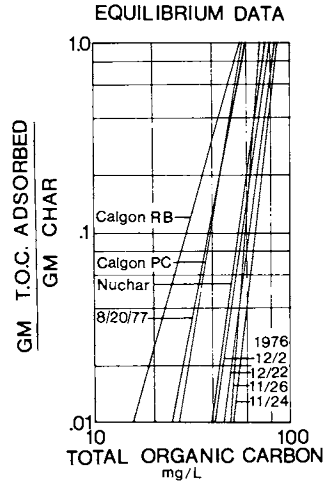

Figure 4. Equilibrium plots of total organic carbon adsorption of commercial carbons and reactor products

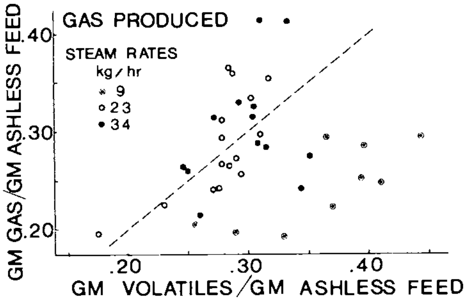

Figure 5. Gas production of the reactor compared with volatile matter fed at various rates of steam addition.

CH_4	10-22
CO	17-27
CO_2	7-9
N_2	4-8
C_2	4
O_2	$\frac{1}{4}$-1

This gas had a calculated net heating value of 300-400 BTU/ft^3.

A comparision of the volatile matter fed to the reactor vs. the pyrolysis gas produced shows that, except at the lowest steam rates, the volatiles in the feed are accounted for in the pyrolysis gas.

Material Balances.

Around the reactor, material balances were run ideally on an 8 hour basis when possible. The best checks on the input vs. the output were found when the ash balance was considered.

However, a check on the entire system showed the process has actually gained 1,000 pounds in fixed carbon in the system over a 30 day period of sustained operation. This is the important fact since even with the losses inherent to any pilot scale operation, the system produced enough activated carbon to sustain the system as a whole.

Energy Balance.

The pilot plant did not actually produce enough gas for the reactor to sustain itself. There were several reasons for this which could not be corrected in the time alloted.

The burners supplied with the reactor were sized so that the excess air used in the combustion could not be reduced below 120%. This meant excessive heat losses in the stack gases.

The insulation for the reactor firebox was designed in a period when energy was cheap. These losses will have to be substantially reduced in this period of energy conservation. For example, the reactor running at temperature but with no load required 1500 to 1700 cubic feet per hour of natural gas. When operating, the gas consumption was in the range of 1700 to 2000 cubic feet per hour.

Another area of energy saving can be found in the dewatering by the filter press. Some small scale testing showed that the use of higher pressures with a change in cloths could produce cakes in the range of 58 to 60% solids. In this range an increase from 50% solids to 60% solids means a reduction of 1/3 of a pound of water to be removed per pound of dry solids or a reduction in drying load of 33%.

While the system was actually self-sustaining in a material sense, there were still losses of fixed carbon. These losses of carbon represent energy which could be recovered in a water-gas type reaction.

When the above energy conservation measures are calculated, the reactor system then becomes self-sustaining on a energy basis.

Reactor Damage.

The reactor sustained damage in two distinct areas. The first developed in the bellows attached to the seals near each end of the rotating tube. Changing the materials of construction from 304 stainless steel, to 601 Inconel, to 316 stainless steel was not the answer. A coating of Dow Corning Silastic 834 was the only temporary measure that could prevent the pin holing and failure of these bellows.

The second failure was in the reactor tube when a weld in the center of the 3rd firing zone failed. There was evidence of sulfide attack in this weld material. In Inconel 617 the weld material is actually the material itself puddled with an inert gas arc. This means the weld is essentially a cast material whereas the plates are a rolled material. Areas of pits were found inside the tube, away from the welds, which showed evidence of sulfide attack of the metal.

The repair was accomplished by cutting out bad areas and replacing entire sections of the tube with new material. The smaller spots were ground and refilled with new weld material. After 6 weeks of operation the repaired areas showed no signs of attack.

Comparisions With Other Systems

Cost estimates were made, by our consulting engineers, for a 10 MGD sized plant for the Activated Carbon system compared with Activated Sludge and Trickling Filters. These are shown in Figure 8.

The construction costs match trickling filters and are about 12% less than activated sludge. Since a multiple hearth furnace is a known substitute for the calciner as the reactor, a cost estimate was made

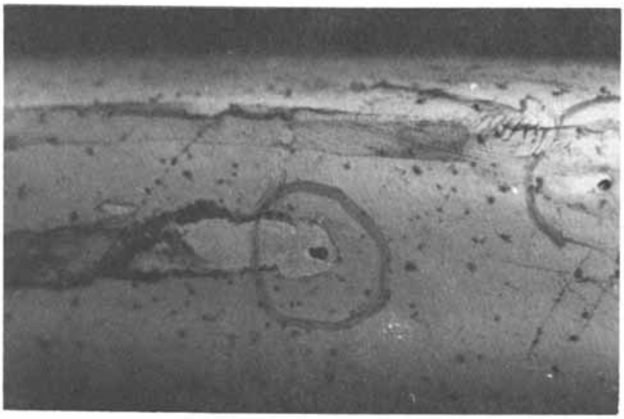

Figure 6. Pitting damage of Bellows Seal on reactor

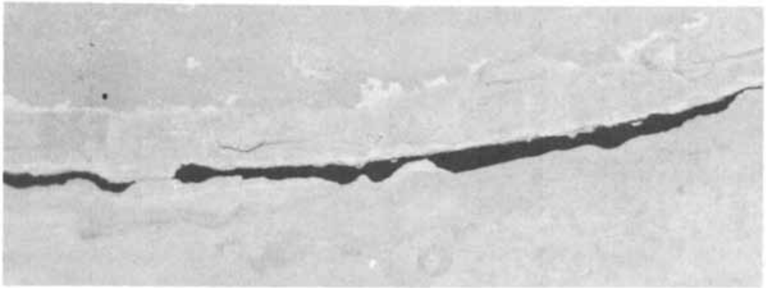

Figure 7. Failed weld of reactor tube

COMPARISONS

COSTS	ACTS	ACT.SL.	TR. FT.
Construction 10^6\$	14.4 (Kiln) 15.6 (Hearth)	16.3	14.4
O & M \$/MG	212	260	180
Energy \$/MG	37.30 (50% cake) 24.30 (60% cake)	61.26	25.44

Figure 8. Comparison of cost estimates at a 10-MGD plant size

on that system as well.

The O & M costs are between the activated sludge and the trickling filter costs. Energy costs are sensitive to the dryness of the cake in the activated carbon system but can be the lowest of the three. Both the activated sludge and the trickling filter systems received credit for gas that could be generated from digestion of sewage solids.

Different effluents were used in the systems evaluated for comparison. These are tabulated below.

	ACTS	ACTS Sludge	Trickling Filter
Effluent			
SS mg/l	25	25	40
BOD mg/l	50	25	50
Grease mg/l	8	7	10

It should be noted that the BOD of the carbon system does not match the effluent from activated sludge at the present state of the art. More work will have to be done on reactor operating conditions to produce a carbon which will remove more of the BOD from wastewater.

A complete report of all operational data has been published. (2)

Literature Cited

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2. "Activated Carbon Treatment System for Municipal Treatment Systems." County Saintation Districts of Orange County, California, November, 1977.

March 3, 1978.

13

Power from Wastes via Steam Gasification

J. A. COFFMAN and R. H. HOOVERMAN

Wright–Malta Corporation, Ballston Spa, NY 12020

The Wright-Malta system had its genesis in a wide-ranging analytical and conceptual study into ways and means of generating power from "difficult" fuels like solid waste and sewage sludge. The system was born complete, rather than growing from research findings or existing product lines. It has changed in detail with refinement in conceptual design, and will undoubtedly change again as the pilot plant is designed, built, and operated. However, the basic rotary kiln gasifier-gas turbine-recuperator-feedback system and cycle have weathered many reviews and appear to be sound and viable.

Projected Power Plants

An artist's conception of a single unit plant is shown in Figure 1. For perspective, it might be noted that the pressurized (300 psi) rotary kiln is projected to be 5 ft. in diameter and 150 ft. in length. The plant would consume 200 tons/day of solid waste and 300 tons/day of liquid (97% water) sewage sludge and produce 10 megawatts of electric power.

Plant operation can perhaps best be described by following material through it. Solid waste is delivered by compactor trucks, is coarsely shredded but not classified, passes through lock hoppers to a metering auger where it mixes with sewage sludge and alkaline catalyst. In the kiln, the material rotates and tumbles, moving forward at about 3 ft/minute. During its slow steady travel to the hot (1100°F) end of the kiln, the organic material, mostly cellulosic, partially dries and pyrolyzes into gases, volatile liquids, and char. The liquids and tars then steam-reform into additional gas; the char undergoes catalyzed steam gasification to yield still more gas. The

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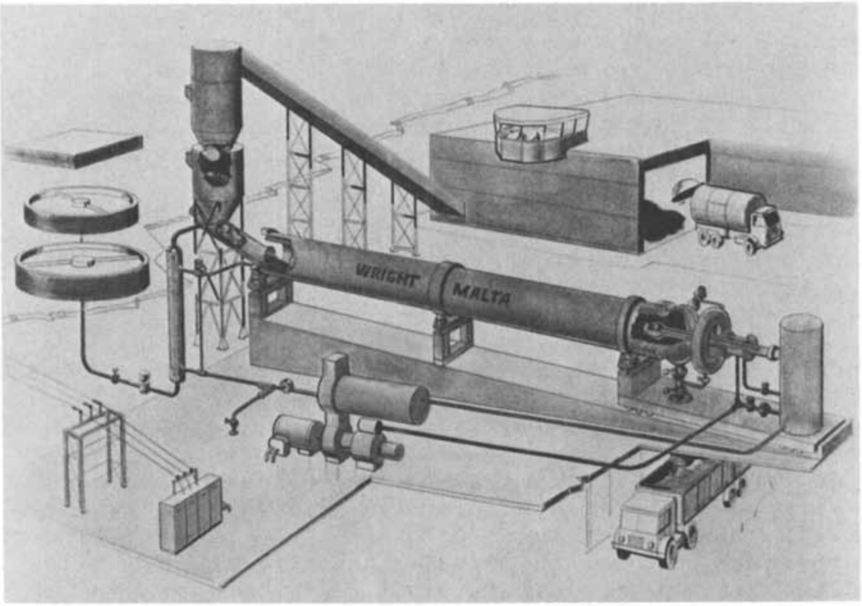

Figure 1. Single-unit power plant

residue, mostly cans, broken glass, and other inorganic debris, is discharged through a rotating triad of lock valves; the steam-laden fuel gas (125 Btu/scf) exits through a filter and a small diameter rotary seal assembly. The whole process is called steam gasification and differs from ordinary gasification in the absence of oxidation.

Most of the hot fuel gas flows to the combustion chamber of the gas turbine, burns, drives the turbine-generator, and gives up its residual heat in a heat recovery boiler (recuperator). Steam so raised and super-heated is super-heated further in a gas-fired unit at the hot end of the kiln, and then, flowing through passages in the kiln wall, gives most of its heat to the gasification process. The condensate exits the kiln at the cool end, gives its remaining heat to the incoming sewage sludge, and then returns to the recuperator.

Recovery of iron and steel from the residue is made easy by the absence of bulky organic material. Cans are prime scrap: clean, unoxidized, detinned. The remainder of the residue is usable as dense, sterile fill for construction purposes.

A W-M plant of this size would serve the homes, commercial establishments, and industries of a community of about 40,000 people, consuming their solid, liquid, and hazardous wastes, and providing one-quarter of their electricity. Units as small as one-tenth this size would still be practical if attached to plants having operating and maintenance crews with appropriate skills. Larger cities would use groupings of kilns and turbines operating in parallel. Still larger cities would have several plants strategically located to keep waste haulage to a minimum. New York City, for example, might have thirty 1000 ton/day 50 MWe plants tucked into appropriate spots in all five boroughs. Such decentralized generation of power and disposal of waste is more economical, efficient, and environmentally attractive than huge regional systems.

Power Cycle

In Figure 2 is shown the W-M power cycle for a 10 MWe plant fueled by 200 ton/day of solid waste and 300 ton/day of sewage sludge. Two major flow streams are evident: the primary power stream consisting of solid waste and sludge, hot fuel gas, very hot combustion products, and somewhat cooler, but still hot, exhaust gases; and the feedback loop of superheated steam and water which returns recovered exhaust heat to the

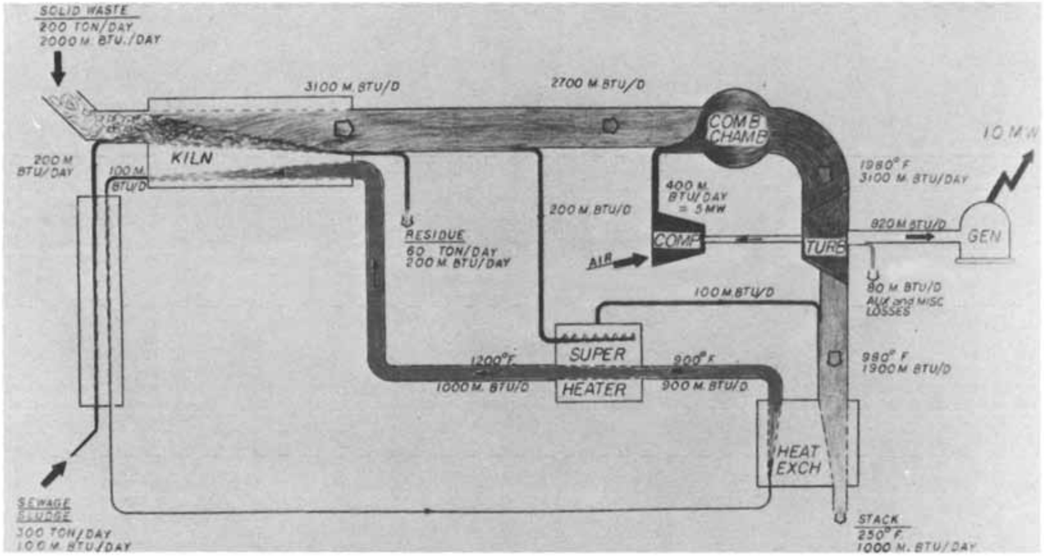

Figure 2. W–M power cycle

gasifier. The width of the channels denotes the magnitude of energy flow through each part of the system.

The diagram is largely self-explanatory, particularly when considered in conjunction with the plant description in the previous section, but these comments might be made. The abundance of steam in the hot fuel gas exiting the kiln means that a less than normal amount of dilution air is required in the combustor. This makes possible a smaller compressor and larger generator than normal, e.g., a 7.5 MW turbine driving a 5 MW compressor and a 10 MWe generator. The turbine is really a partial steam turbine since it derives a significant portion of its power from the steam raised in the kiln. Thermodynamically, the cycle is similar to that in the combined gas turbine/steam turbine system and the efficiencies are similar, about 40%. For comparison, it might be noted that steam plants fueled by petroleum have efficiencies of 32-35%.

The cycle is attractive as drawn but still has room for improvement. One change for the better might be to place the sludge pre-heater in the lower end of the turbine exhaust, thereby eliminating a step in the transfer of heat and pulling the exhaust temperature down by, say, another 50°F. It is probable, too, that further development work will lead to reduction in the energy content of the kiln residue, from 200 MBtu/day to, say, 100 MBtu/day. The cycle efficiency calculable from the diagram is conservative because it is based on present experimental data and performance characteristics of present turbines and does not take into account improvements that are likely to result as the system matures.

Experimentation-Minikiln

W-M research on steam gasification of organic solid materials has been performed in two types of equipment: a batch-type miniature kiln (minikiln) and a continuous feed biomass gasifier (biogasser).

The minikiln research apparatus is shown in Figure 3, installed in the test bay of the gasification lab. It is, in effect, a rotating autoclave, intended to simulate, by tumbling action, gradual increase in temperature, and change in gas composition, the slow traverse of a charge of material through a long, pressurized rotary kiln. There is provision for steam injection during the course of a run and controlled release of gas. There also is provision (not shown) for condensation of water and organic liquids from the effluent gas, and analysis of the fixed gas by chromatograph. For perspective, it might be noted that the inner barrel of the minikiln is about 3½ ft. in length by 1 ft. in diameter.

In a typical run, 10 lbs of coarsely shredded solid waste, 2 gallons of liquid sludge (in milk cartons), and ½ lb of sodium carbonate are placed in the minikiln. The door is bolted on, the thermocouple leads attached, and rotation and heating started. Steam is injected as needed to hold the desired water-to-waste ratio. Waste decomposition starts at about 350°F and reaches its peak at about 500°F (mostly CO_2, CO, and H_2O). Methane starts evolving at about 500°F, hydrogen a little higher; by 900°F, the flow of CO has stopped. From 1000 to 1100°F, the gas is mostly hydrogen and CO_2 in a ratio of about nine to four.

$$4CH_{.5}(\text{char})+8H_2O \rightarrow 4CO_2+9H_2$$

During the period from about 400 to 900°F, liquids and tars are also being evolved and are caught in the condenser. It is estimated that perhaps a half of the organic waste is volatilized as condensable liquids. And so the gas analysis does not give a true picture of the composition that would be realized in a continuous process, where gases from steam-reforming of the liquid volatiles would combine with those from the initial pyrolysis and the char gasification.

An approximation was, however, obtained by running the minikiln as a bomb with no release of gas, starting with a sufficiently small charge of wood, water, and catalyst so that pressure limits were not exceeded. In a run to peak pressure and temperature of 400 psi and 1100°F, the gas had this composition:

CH_4	18%
C_2H_6 etc.	8%
H_2	28%
CO	1%
CO_2	45%

Note that the CO had almost entirely water gas-shifted to CO_2, but that the methane and rich ethane were unaffected. The liquids and tars, which must have been present at intermediate temperatures, had steam-reformed cleanly, judging by the fact that there was no soot in the residue and its weight was slightly less than normal.

On the graph in Figure 4 are shown char weights,

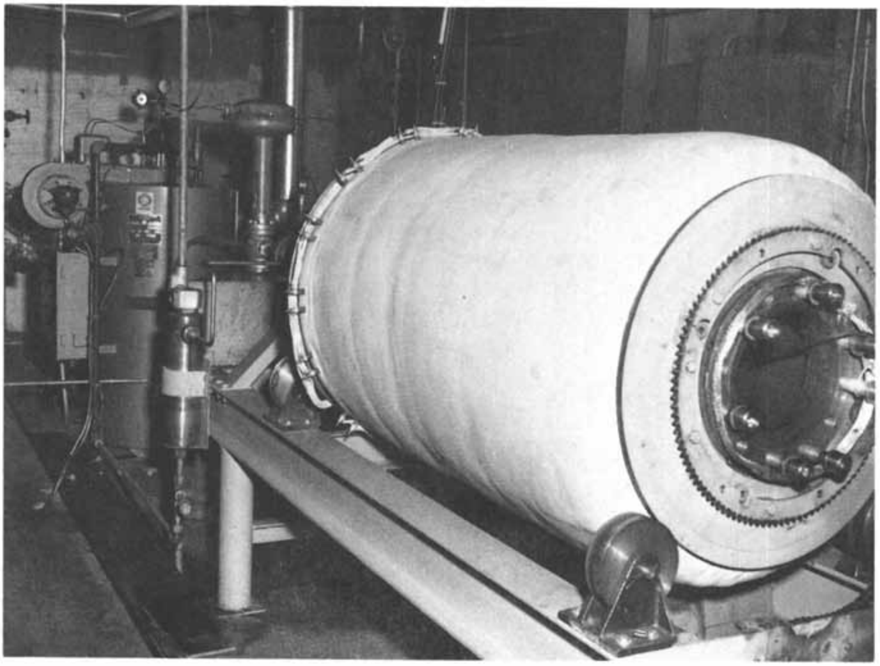

Figure 3. Minikiln research apparatus

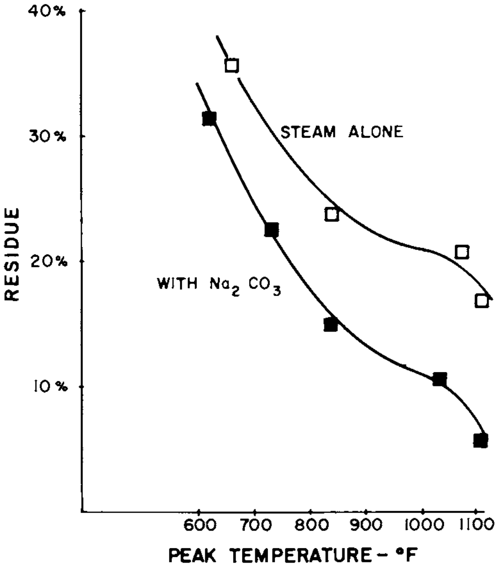

Figure 4. Residual char as function of peak temperature, paper at 200 psi

as percentages of dry paper starting weights, after reaction with water/steam to the temperatures shown. Note that the weight, which is dropping steeply with paper decomposition, starts to level off, then drops again with the onset of char gasification. Note also that the catalyst exerts an influence from the lowest temperature. It is probable that the catalyst operates by several mechanisms: hydrolysis at the lowest temperatures, then decarboxylation and dehydroxylation, then reaction with benzylic acidity, and finally, reaction with free radicals in the char and promotion of ring opening by steam.

Figure 5 shows residual char weight of wood chips after heating to 1100°F at the pressures shown. Again the effect of the catalyst is marked and apparent from the lowest steam pressure, and the curve still seems to have a substantial slope at 400 psi. The pressure and temperature curves appear to be independent, and this gives the designer some degree of freedom to pick the combination of pressure, temperature, feed rate, and residue which is most economical.

Experimental work on coal showed that both lignite and bituminous are substantially more refractory than is solid waste. There was some reaction of coal with steam, but under conditions where waste was essentially consumed, the residues from both types of coal were only about one-third less than if they had been pyrolyzed. More strenuous conditions may suffice, and if they do so, coal can be considered as a supplemental fuel in a W-M waste-to-power system.

Several experiments on polychlorinated biphenyls (PCB's) were made to determine the efficacy of W-M reaction conditions for destroying PCB's in municipal and industrial wastes. Bomb runs were employed to prevent the escape of any material; typical charges were equal parts by weight of PCB's, water, paper, and sodium carbonate in amounts sufficient to generate about 400 psi at 1100°F. The data are shown in Figure 6. It is evident that a time-temperature of about an hour at 1100°F is necessary and sufficient to complete the PCB decomposition. By inference, since PCB's are among the most stable organic materials known, any hazardous organic waste would be consumed in a W-M plant at little cost and no detriment to the environment.

One other finding might be mentioned before leaving the minikiln work. The apparent activation energy for char-steam gasification at 200-400 psi in the temperature range of 1050-1100°F was calculated to be only one-half (30 kcal/mole) that at one atmosphere

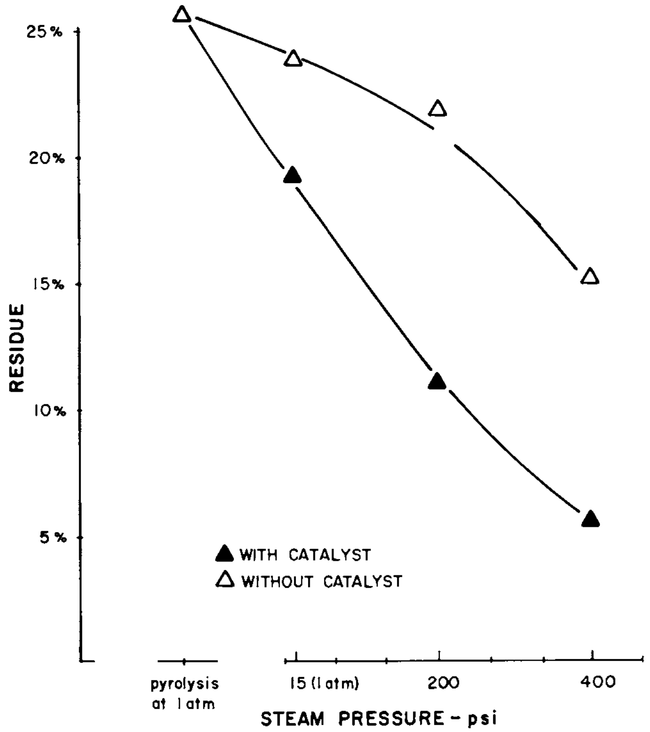

Figure 5. Residual char of wood as function of pressure: peak temperature 1100°F

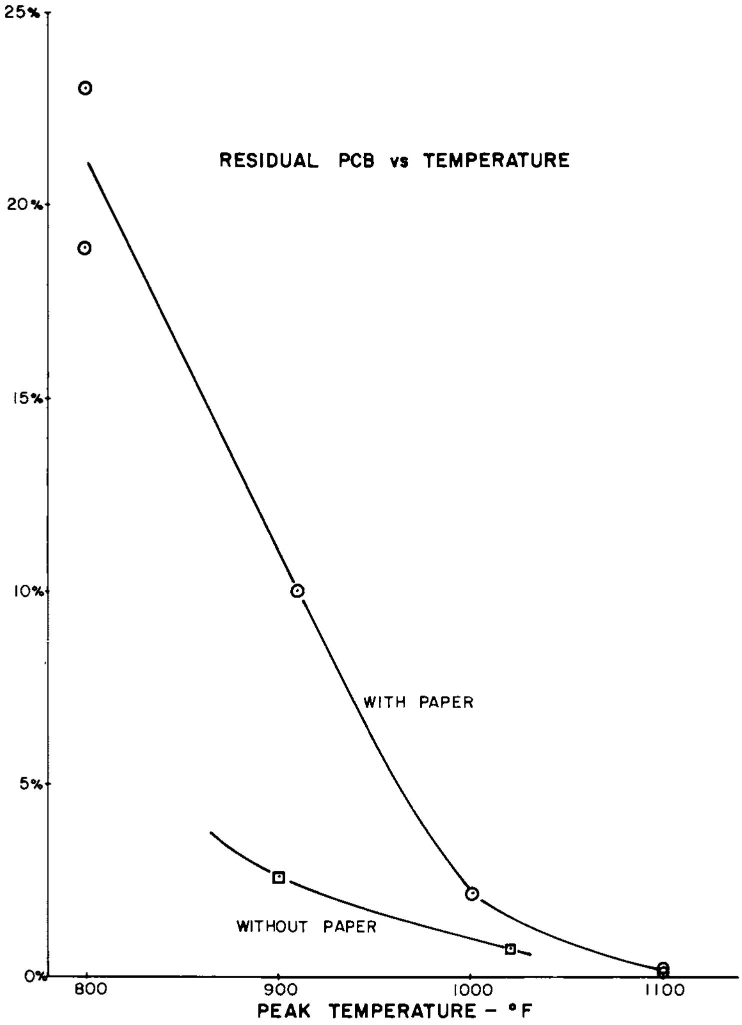

Figure 6. Residual PCB vs. temperature

(60 kcal/mole). It is probable that the lowered activation energy is due to the presence of a condensed aqueous film on the char particles, which promotes reaction between water and char. It is probable, too, that the hygroscopic nature of the catalyst helps to create the condensed phase and sustain it to higher temperatures.

Experimentation-Biogasser

Research on gasification of biomass has production of fuel gas as its ultimate purpose, and material transport within the gasifier is by auger rather than kiln action. However, the chemistry is essentially the same, and the experimental findings are directly applicable to the waste-to-power program.

The biogasser is shown in the final stages of its construction in Figure 7. In the upper foreground is the fuel reservoir, the insulated reactor with protruding thermocouple connections slants back through the center of the picture, and in the background are the residue trap and gas release valves. The condenser and gas/liquid separator are along the lower right-hand side of the reactor. For perspective, it might be noted that the biomass slurry reservoir is about 6 in. ID and 6 ft. in height, the slanted reactor is 1.75 in. ID and 10 ft. in length, and the residue trap is about 3 in. ID and 3 ft. in height. The equipment is stainless steel throughout.

Its operation can be described by moving through the sequence of a typical run. The cap and piston are lifted from the fuel reservoir, the reservoir is filled with sawdust or other shredded biomass, water, and catalyst, and the piston and cap replaced. Then the heaters are turned on, the hydraulic pump driving the piston down is started as is the auger drive motor, and the slurry moves slowly down out of the reservoir and up the slope of the reactor. Temperatures rise until the desired profile is reached, typically 100°F at the feed end and 1200°F at the trap and reducing valve; the pressure rises to the desired figure (up to 3200 psi) by evolution of gas and formation of steam, and then that pressure is maintained by controlled release of gas to the condenser. As the gas is released, the condensate is monitored for liquids and tars, and the non-condensing gas is analyzed by chromatograph. At the end of the run, the reactor and trap are examined for residue. In satisfactory steady state operation, the gasification is reasonably complete, the condensate contains little organic liquid or tar,

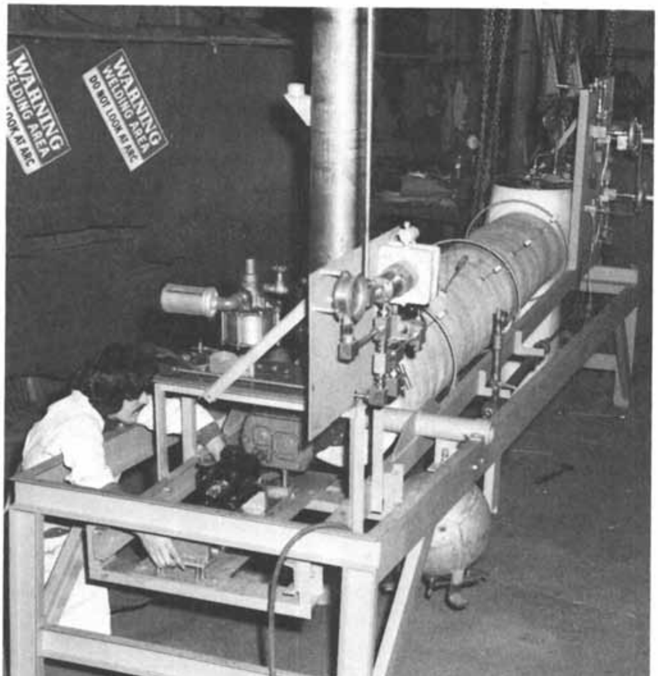

Figure 7. Biogasser in its final stages of construction

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

and the residue is essentially inorganic.

In the graph of Figure 8 are shown data on residues as a function of pressure from the biogasser (triangles) and the minikiln (circles). The minikiln data show how the char residue decreases rapidly with pressure in the range of 0-400 psi and how the effect of pressure is greater in the presence of added Na_2CO_3 catalyst. The residues from biogasser runs at one atmosphere and 300 psi lie slightly below the corresponding minikiln curves because the biogasser was operated at a slightly higher temperature; otherwise the biogasser data seem consistent with the minikiln data in this region and show a catalyst effect of the same magnitude.

By 600 psi, however, the effect of the catalyst seems to have disappeared; the residue from the catalyzed biogasser run is as large as that from the uncatalyzed run. And at 1000 psi, the char residue is clearly heavier than those at lower pressures. The exact shape of the curve is not well-defined by these experiments because of difficulties in maintaining the same peak temperatures in runs at different pressures. Nevertheless, it is evident that the trends in the lower-pressure minikiln data may not be extrapolated to higher pressures.

Qualitatively, there is also a change in the nature of the residue at higher pressures. In both the minikiln and biogasser runs at pressures up to 300-400 psi, the char residue is loose and granular. In the biogasser runs at 600 and 1000 psi, much of the residue is stuck together in hard lumps. This type of caking and increase in residue become progressively more apparent as the pressure increases, at least to 3200 psi.

The observations suggest that there are two pressure influences working in opposite directions. In the lower range, increase in pressure promotes reaction by increasing the concentration of steam and formation of aqueous films. But above 300-400 psi, this effect is swamped out by an increasing tendency toward formation of coke-like material, presumably from chemical condensation of volatilized liquid intermediates. The latter effect is consistent with LeChatelier's principle.

It is concluded that for maximum conversion of solids to fuel gas, there is an optimum in reactor pressure in the neighborhood of 300-400 psi. This finding is significant to further development of the W-M process, in that it is now established that there is no advantage in going to higher reactor pressure.

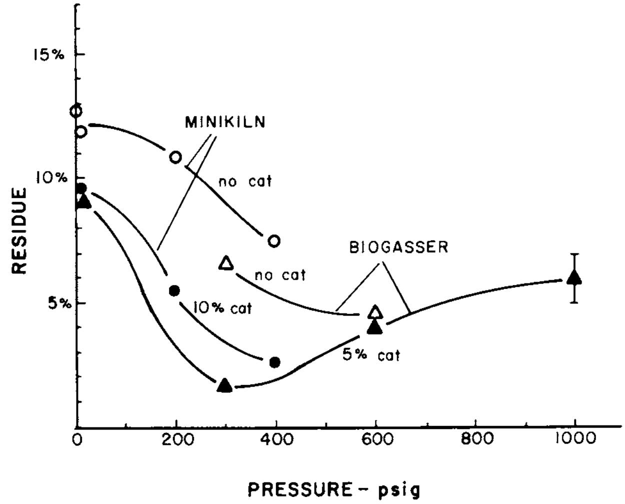

Figure 8. Residual wood char vs. pressure, minikiln and biogasser

It is fortunate that the optimum pressure coincides with that deemed most desirable for kiln operation for other reasons: gas turbine operation, long gas dwell time in the kiln, low particulate, heat transfer, etc.

Gas composition data from the biogasser support earlier work in the minikiln, the proportions of hydrogen, methane, carbon dioxide, etc. being similar to those in minikiln bomb runs. Under conditions of essentially complete conversion of solids to fuel gas (1150°F, 300 psi), the fixed gases had an integrated hydrogen-to-oxygen ratio of two to one, the same as in cellulose and water. At less than complete conversion, the ratio fell below this figure, because the oxygen tends to come off first, as CO_2 and CO, and the hydrogen tends to stay behind in the char and liquids. Use of this gas measuring and integrating technique enabled the gathering of much more data on the degree of conversion versus temperature and pressure than would have been possible by measure of residue weights. The graph in Figure 9 shows the hydrogen to oxygen ratio (times $\frac{1}{2}$) as a function of reactor temperatures. It is to be noted that the temperatures are maximum wall temperatures as measured by thermocouples embedded in the reactor wall. It is estimated that the corresponding gas temperatures are about 200°F lower. The bend in the curve, therefore, comes at about 1150°F, corresponding to complete gasification as judged by residue and condensate.

Calculations have been made from heats of formation to determine equilibrium gas compositions as a function of water/wood ratio, temperature, and pressure. These calculated equilibria have been found to correspond reasonably well to experimental compositions above 1150°F; below that temperature, the experimental compositions deviate in favor of the first formed products. This finding has theoretical significance but little practical utility in the kiln power system where parameters like water-to-waste ratio are fixed by cycle considerations. However, in the biomass gasification system, the water/wood ratio can be varied to maximize the utility of the product gas toward its intended use: medium Btu fuel gas, synthesis of methanol or ammonia, or methanation to pipeline quality gas.

Pilot Plant Gasifier

The transition from bench scale research to pilot plant will be accomplished in stages, the first being gasifier design. A schematic is shown in Figure 10.

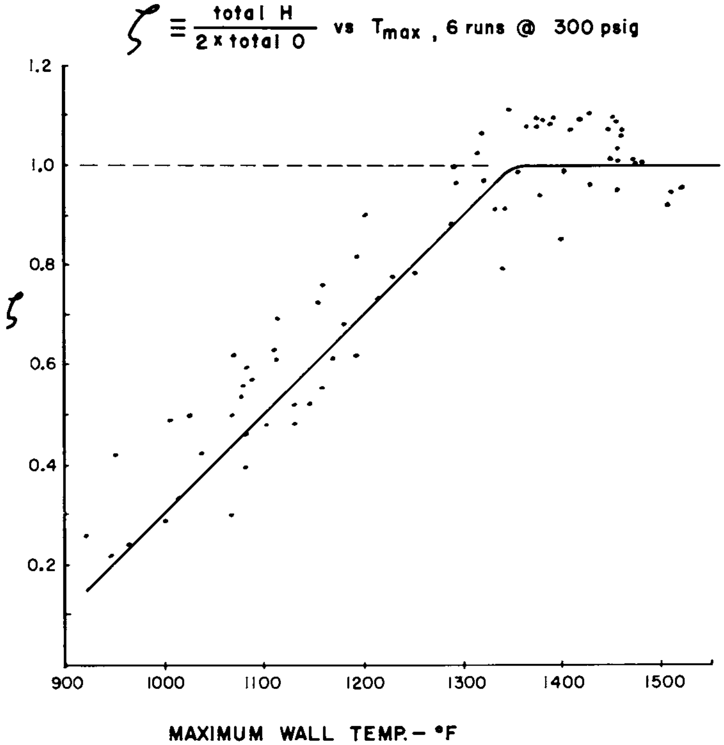

Figure 9. The hydrogen to oxygen ratio as a function of reactor temperatures

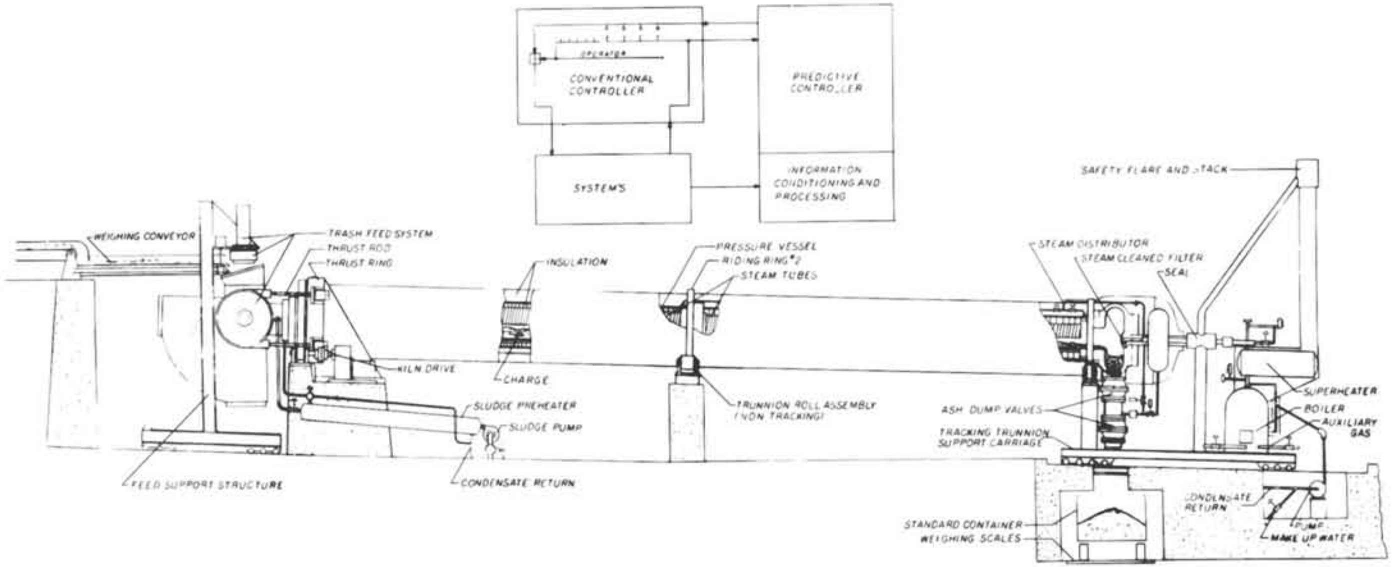

Figure 10. Schematic of pilot plant gasifier

The pilot plant gasifier is projected to be 60 ft. in length and $2\frac{1}{2}$ ft. in diameter at the cool end, and thus has about one-tenth the volume of the production unit described earlier. Its nominal throughput is 20 tons/day of solid waste and 30 tons/day of liquid sewage sludge, and its product gas, consumed in a gas turbine, would generate 1 MWe.

There are significant differences between this design and that described earlier, these reflecting the continuing evolution of the W-M system.

Feed. Gravity-feed lock hoppers have been discarded in favor of a positive piston-feed system. Coarsely shredded but unclassified solid waste is delivered from the left by weighing conveyor and deposited into the open throat of a cylindrical cavity. The material is tamped lightly into place and the cylinder sealed with the tamping piston. Then the inner structure rotates ninety degrees, and the waste is pushed from the cavity into the kiln by a second piston. At the completion of this piston stroke, the inner structure back-rotates to the vertical position, the tamping piston lifts, the inner piston drops, and the cycle is repeated. Liquid sludge is pumped continuously through its pre-heater into the throat of the kiln, mixing there with the solid waste. Homogenization will occur in the first few feet of the kiln.

The feed system is held in place against kiln pressure by a thrust rod, thrust ring assembly working against the front flange of the kiln, and requires no external bracing. The feed support structure is mounted on rollers so that the feed system can be unbolted and moved away for maintenance of the front rotating seals, which are inside the thrust ring and independent of it.

Kiln Support. The fixed point for the kiln is established through the thrust assembly at the front end. This means that the front end trunnion roll assembly and the middle one are non-flanged and non-tracking.

The entire non-rotating assembly at the hot end of the kiln is mounted on a support carriage on rollers. Its position relative to the kiln is determined by having this end of the kiln supported by a tracking trunnion. Movement of the support carriage with thermal expansion and contraction of the kiln means no allowance for axial thermal motion is required in the rotating seal assembly.

Stepped Kiln. The outer barrel of the kiln has a uniform, slightly greater than 2½ ft. diameter throughout its length. The inner shell, which contains the waste and on which the steam heating tubes are wrapped, is 2½ ft. in diameter for the first 30 ft., then decreases to 2 ft. for the next 15 ft., and decreases further to 1½ ft. for the final 15 ft. The space between the steam coil and the outer barrel is pressure-equalized and filled with thermal insulation. This reduces the temperature of the outer barrel at the hot end to about two-thirds of what it would be if the barrel were immediately adjacent to the steam tubes. This means that the barrel can more readily carry its pressure load and can be less massive than if the inner shell were not stepped down

Another benefit of the stepped design is the slowing of the forward movement of the material as the diameter decreases. Much of the material bulk will have been lost by the time it reaches the mid-point, and the kiln will not dam up at the step. The resulting increase in dwell time will aid in completing steam gasification of the char.

Steam Heating. The steam heating system will operate at 600 psi, approximately twice kiln gas pressure. This will ensure an adequate temperature differential between the condensing zone in the steam coils and the evaporating zone within the kiln. The boiler and superheater will be fired by a portion of the gas from the kiln, including the low pressure gas from the dump valves; the remainder will be flared. Gas will leave and steam will enter the rotating structure through a small diameter stuffing box seal, of the type constructed for the minikiln and proved satisfactory there. The condensate will exit the kiln through the center of the face seal at the front end and, by virtue of its higher pressure, provide assurance against gas leakage.

Gas Filter. Because of the high humidity in the kiln, the relatively low gas velocity, and the gentle tumbling action, little entrainment of particulate in the gas is anticipated. Nevertheless, a gas filter is provided in the exit line. It has a convex face which is continuously wiped from behind as it rotates by a jet of steam from a stationary finger, shaped to the contour of the filter. Dirt thus dislodged falls into the discharge cavity.

Residue Discharge. The dump valves are full-bore ball valves and the channels above and between the valves have the same diameter as the channels within the valves; there are no shoulders on which material will lodge. The valves swing with the kiln. At the bottom of the first swing, the top valve opens and the accumulated material drops through the valves into the cavity between them. Then the valve closes, and as it swings, the cavity is slowly drained of its fuel gas. At the bottom of the second swing, the bottom valve opens and drops the residue into the receiving container. Then the valve closes and the cycle is repeated.

Controls. The control system will be designed for manual or automatic operation For automatic operation, the predictive controller will be programmed to maintain the apparatus within pre-set limits. The operator will have the ability at any time to override the predictive controller or to re-set its limits. Because of the slow traverse of material through the kiln and hence the sluggishness of response to changes in quality and quantity of feed material, the control system will have anticipatory character, sensing the rate of change as well as the actual values of control parameters.

When operated with a gas turbine in later phases of development, the gasifier pressure will be set about 50 psi higher than that required at the turbine combustion chamber inlet. The kiln thus becomes a small surge tank, and its pressure fluctuations will militate against changes in fuel value of the waste being felt as surges on the power line. (That, and a smart operator.)

Economics

Economic analysis work to date assumes utility ownership and operation, with distribution of the generated power through the utility grid. (A W-M plant is to be thought of as a power plant fueled by waste rather than a waste disposal plant with electricity as a by-product.) The municipality would have responsibility for collection and delivery of solid waste to the plant, delivery of liquid sewage sludge, and removal and disposal of plant residues. Putting the interface between utility and municipality at the unloading dock would seem to be the most reasonable alternative because it preserves the traditional role of each. Other arrangements are possible, of course:

municipal operation and distribution of power, municipal operation with sale of power to the utility, third party operation, etc. They would entail different assumptions regarding tax and interest rates, and would yield different results.

The capital cost of the basic 10 MWe, 200 ton/day solid waste, 300 ton/day sewage sludge plant is estimated to be $9 million or $900/KW. The annual direct operating cost is estimated at $850,000. To this must be added depreciation, interest on depreciable investment, local taxes, and income taxes. If one assumes tipping charges of $8/ton for solid waste and $10/ton for liquid sewage sludge, and values of 2, 3, and 4¢/KWhr for the electricity generated, the 20-year average return on investment for the three electricity values is 8, 15, and 22% respectively.

At the 2¢ level, about one-half of the revenue is derived from the electricity and one-half from the tipping charges. At the 4¢ level, the ratio of electric revenue to tipping revenue is two-to-one. This means that, at 4¢, the tipping charges could be eliminated and the plant would still earn 8% after taxes. (For perspective, it might be noted that the $8 and $10 charges are comparable to present average costs across the country for solid waste and sewage sludge disposal. Four cents per kilowatt hour is projected by up-state New York utilities as the generating cost in new fossil fueled plants, and this figure also appears to be about average nation-wide.)

The economics of W-M plants in metropolitan areas will be even more attractive. The scaling factor is not large but is significant; the profit margin increases by about 45% in going from 200 to 1000 ton/day plants. In New York City, for example, where disposal costs for solid waste and sewage sludge are presently about $12 and $16/ton, and busbar electricity is worth at least 4¢/KWhr, 1000 ton/day, 50 MWe plants would earn the utility a handsome 43% annual return on investment.

Any economic analysis at this stage of development must, of course, be only an approximation, but there can be no doubt of the economic worth of the system. Capital investment will almost certainly be no more per kilowatt than coal-fired steam plants with adequate environmental protection and less than nuclear. Labor costs in the waste plants will be somewhat higher but more than offset by the negative cost for fuel. The result for communities: lower electric rates, lower waste disposal costs, or both.

Environmental Characteristics

The positive environmental characteristics of the W-M system are evident: its ability to convert solid, liquid, and hazardous wastes into power. Its negative aspects appear to be minimal.

A W-M power plant will use no cooling water, nor will there be any aqueous discharge from gas purification.

The stack gases are likely to be cleaner than those of conventional gas turbine or steam plants operating on natural gas or oil. The kiln gas is rich in hydrogen and steam and will burn about 600°F cooler than usual, thereby minimizing formation of NO_x. Complete oxidation of CO to CO_2 will also be aided by the high steam content. Chlorine and most of the sulfur will be caught by the alkaline catalyst. Particulate will be within acceptable limits. (It has to be to avoid turbine erosion.)

With regard to hazardous metals, kiln conditions are so mild, compared to incineration or partial oxidation gasification, that their volatilization is unlikely to be a problem. Most hazardous metals will wind up in the residue as partially hydrated oxides and carbonates. The fate of mercury is uncertain and will be investigated.

Acknowledgements

The W-M gasification/power program has been supported by these agencies: Empire State Electric Energy Research Corp (NYS utilities) for construction of the minikiln; US Environmental Protection Agency for experimentation therein on solid waste and sewage sludge; New York State ERDA for coal experiments; US ERDA-Conservation Research & Technology, now DOE-Power Systems Div. for analysis of the power cycle; and DOE-Solar for construction of and experimentation with the biogasser.

MARCH 15, 1978.

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Pyrolysis of Scrap Tires Using the Tosco II Process—a Progress Report

B. L. SCHULMAN and P. A. WHITE[1]

Tosco Corporation, 18200 West Highway 72, Golden, CO 80401

During the course of developing the TOSCO II process for pyrolyzing oil shale, Tosco examined the application of this process to other organic materials such as coal, tar sands, municipal waste and scrap tires. Initially, the processing of scrap tires became of interest because of the high oil yield obtained from the decomposition of the rubber — about 150 gallons per ton, compared to 30-35 gallons per ton for oil shale. At that time, carbon black had approximately the same value as oil and, although it was known that the carbon black was recovered in good quality, its contribution to the potential economics was modest. In 1974, both carbon black and oil prices rose sharply, and later operations focused more closely on the carbon black product. After a series of pilot plant investigations, The Goodyear Tire & Rubber Company and Tosco entered into a program to define the commercial potential of the tire program. A major part of this effort was to construct and operate a 15 ton a day pilot plant specifically designed for pyrolyzing scrap tires and develop the appropriate techniques to recover the valuable liquid and carbon black materials in a suitable commercial form. Generally, the goals of the program were to define the methods for improving the process efficiency in three areas:

1. Optimize process conditions for maximum plant capacity and process yields.
2. Demonstrate best techniques for recovery of the carbon black with an acceptable commercial quality.
3. Demonstrate new mechanical hardware which could improve product recovery or lower total plant investment.

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[1] Current address: Tosco Corporation, 10100 Santa Monica Blvd., Los Angeles, CA 90067

0-8412-0434-9/78/47-076-274$05.00/0

Pyrolyzing Scrap Rubber Showed Surprising Yield — Polymer Cracks Easily With No Coke Formation

The pyrolysis of rubber tires was found to be surprisingly different from processing other organic materials in that the materials making up the tires decomposed in a "cleaner" manner than other organic materials. Scrap tires consist of many elements which can roughly be categorized into four elements:

1. Synthetic rubber (butadiene-styrene copolymer)
2. Extender oils
3. Organic cord and inorganic belting
4. Carbon black and fillers

Analysis of the pyrolysis products which form at 900 to 1000°F showed that the organic polymers simply decomposed into liquid and gas with no coke deposition on the residue. The carbon black, fillers and inorganic belting remain behind as a solid residue. The form of the residue is essentially the same as the original materials used in the tire. For example, fiberglass cord may unravel into single strands; steel (as belting or bead wire) will show up as single or tangled strands; and the carbon black, together with the fillers, will appear as individual particles or loose agglomerates less than one micron in diameter.

The major technical advantage of this controlled pyrolysis is that the carbon black is recovered in a "virgin form" — its reinforcing properties are the same as those of the mixture of the various grades of the carbon black which is used in the initial formulation of the tire. The inorganic additives are indicated by the ash content — the typical tire char will show an ash content in the range of 13-16%, as illustrated in Table I. The ash consists mostly of zinc, silica and titanium, with varying amounts of other elements as shown in Table II. Standard rubber formulation tests, together with the use in making tires has shown that the carbon black can be re-used in portions of the tire according to the char's grade.

The Key To A Successful Process Is To Maximize Capacity And Maintain Carbon Black Quality

The general TOSCO II process description has been well documented.(1) A simplified flow diagram of the process is illustrated in Figure 1. The shredded scrap tires are dried and fed to the rotary pyrolysis drum. Hot ceramic balls from a ball heater flow to the drum, also, where the mixing action raises the reaction temperature to 900-1000°F. The organic materials decompose

Table I. Ultimate Analysis of a Typical Tire Char

	Non-Belted	Steel-Belted	Fiberglass-Belted
Moisture, %	---	---	---
Carbon, %	81.15	81.49	82.66
Hydrogen, %	0.74	0.75	0.85
Nitrogen, %	0.17	0.13	0.27
Chlorine, %	0.25	0.24	0.18
Sulfur, %	2.34	2.46	2.33
Ash, %	15.36	14.95	13.71
	100.00	100.01	100.00

Table II. Chemical Analysis of a Typical Tire Char Ash (Average of Several Samples)

Element (wt %)	Non-Belted (Ash 15.4 wt %)	Steel-Belted (Ash 15.0 wt %)	Fiberglass-Belted (Ash 13.7 wt %)
Antimony	<0.03	<0.03	<0.03
Aluminum	2.40	1.50	2.40
Chromium	0.15	0.13	0.16
Iron	2.00	1.10	2.00
Lead	0.16	0.18	0.25
Barium	0.27	0.25	0.60
Magnesium	2.30	1.50	2.50
Phosphorus	1.60	0.80	1.40
Potassium	0.38	0.32	0.36
SiO_2	19.62	19.09	21.31
ZnO	45.45	52.33	45.64
TiO_2	11.99	14.27	10.28
CaO	6.04	5.06	6.08
Sulfur	2.67	2.77	1.27

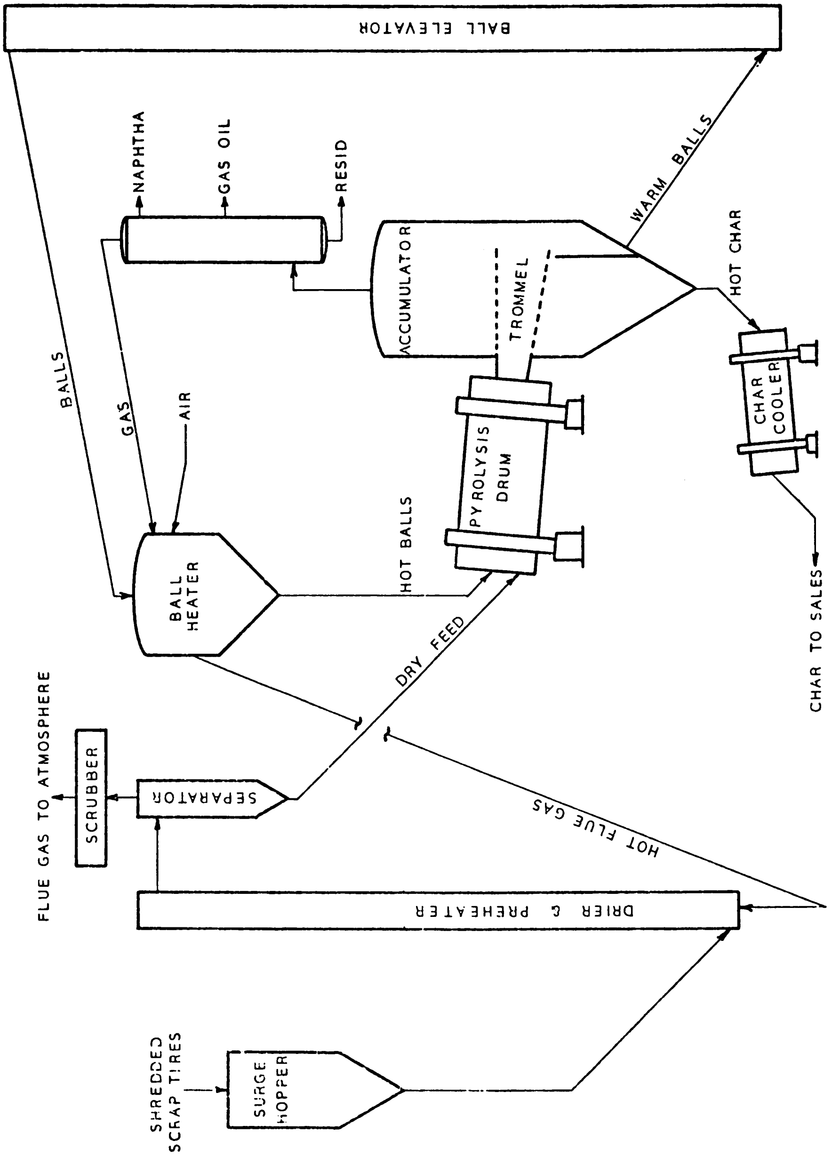

Figure 1. The Tosco II process

into oil and gas vapor and pass out the top of the accumulator vessel into a fractionator. The oil is condensed into various fractions and separated from the gas. The gas, after sulfur removal, can be used as ball heater fuel. The mixture of balls and solid residue is separated by allowing the fine residue to fall through a fine rotating screen, called a trommel. The warm balls are collected separately and returned to the heater. The hot residue is cooled and treated to remove extraneous materials such as steel, fiberglass, stones and other material from the carbon black. For ease of handling, the carbon black is pelletized for shipment to the consumer.

The key to a successful process is the efficienct handling of the feed material into the pyrolysis system and maintaining good recovery of the carbon black from the process. For example, due to its fineness, carbon black may be entrained into the oil fraction, which represents an economic loss. Similarly, it is important that carbon black be effectively separated from the steel and fiberglass contaminants without a loss on the less valuable materials. Such efficient handling, together with the prevention of carbon black being emitted to the atmosphere, is a key part of the process development.

The recovery of the liquid product is also an important factor since the oil should have many valuable uses. It is very fluid liquid containing about 1% sulfur and can be used directly as an industrial heating oil. Typical properties are shown in Tables III and IV.

The gas produced is highly olefinic, containing substantial amounts of ethylene, propylene and butylene — as illustrated in Table V. Depending on the economic situation, the process can be run at high temperature to maximize gas and minimize liquids. If the temperature is kept at the low range, gas yield is minimized and liquid yields increase. At the same time, the lower temperature provides a large increase in capacity so that careful economic judgments must be superimposed on the processing results to define the optimum conditions. In addition to the capacity, a key item is to keep carbon black yield high and maintain its good quality. Our initial results show this can be done. The impact of the carbon black on product realization is significant, as shown in Table VI.

Table III. Inspection Data for Tire Oil and Condensate
(Average of four samples of oil and four samples of condensate)*

	Tire Oil		Typical No.6
	Whole Oil	Condensate	Fuel Oil[1]
API Gravity, @ 60 F*	14.8	38.9	12.4
Specific Gravity, @ 60 F	0.9674	0.8310	0.9833
TCC Flash Point, @ 760 mm*	100 F	---	208 F
Pour Point*	20 F	---	60 F
Ash, wt %*	0.10	---	0.06
Viscosity, @ 100 F, cSt*	8.54	---	118.7
Viscosity, @ 210 F, cSt*	1.96	---	---
Heating Value, Gross Btu/lb	18,030	---	18,283
Reid Vapor Pressure, psig	0.2	0.2	---
Doctor Test	---	Positive	---
Mercaptans, ppm	0.8	3	---
Research Octane Number	---	95.0	---
Conradson Carbon, wt %	3.41	---	10.35
BaP Content, ppm*	6-20	---	---
Saturates, vol %	1	3	---
Olefins, vol %	4	5	---
Aromatics, vol %	95	92	---

[1] Schmidt, Paul F., "Fuel Oil Manual," Industrial Press, Inc., 1969.

Table IV. Inspection Data for Tire Oil and Condensate
(Average elemental analysis of filtered tire oil and condensate)

	Tire Oil		Typical No. 6
	Whole Oil	Condensate	Fuel Oil[1]
Carbon, wt %	87.11	86.85	86.4
Hydrogen, wt %	9.74	10.67	12.9
Nitrogen, wt %	0.59	0.25	---
Oxygen, wt %	1.07	0.76	---
Sulfur, wt %	1.04	0.76	1.20
Arsenic, ppm	1.1	---	---
Copper, ppm	0.2	---	---
Iron, ppm	133	---	---
Nickel, ppm	0.3	---	---
Sodium, ppm	27.5	---	---
Vanadium, ppm	0.8	---	---

[1] Schmidt, Paul F., "Fuel Oil Manual," Industrial Press, Inc., 1969.

Table V

TYPICAL PILOT PLANT PRODUCT GAS YIELD

Gas Yield (lb/ton)	
H_2	1.49
CO	12.25
CO_2	41.24
H_2S	1.82
C_1	24.39
C_2	18.50
C_2-	22.76
C_3	9.24
C_3-	24.27
i-C_4	2.19
N-C_4	2.28
C_4-	61.59
	222.02

Table VI. Gross Realization for Tire Pyrolysis
(Basis: 1 Ton of Tire Feet)

Item	Production	Product Value*	Most probable Revenue
Oil	3 to 4 bbls	$14.00/bbl	$49.00
Carbon Black	575 to 700 lbs	$ 0.095/lb	$60.00
Steel	85 to 100 lbs	$40.00/T	$ 2.00
Fiberglass	70 to 80 lbs	-0-	-0-
		Gross Realization - U.S.	$111

*These are TOSCO's estimates of the market value for these products. The values shown have been discounted to reflect potential marketing penalties.

Demonstration Program is Continuing

The 15 ton a day pilot plant has been operating since early 1977 in various stages. A series of tests has been made in key areas to define the key operating parameters and critical elements in several areas which can affect the process economics:

1. Tire shredding, including separation of steel from rubber feed
2. Pyrolysis operations
3. Oil recovery
4. Carbon black recovery and handling

This series of tests has identified major leads to improving the various areas to maintain product quality, increase capacity and maintain high carbon black recovery of good quality. Several simplifications of the mechanical equipment for solids and liquids handling have been made to the unit. Operations are continuing in 1978 to define optimum process conditions, demonstrate long term equipment operability and make large tonnage samples of carbon black for commercial evaluation.

Literature Cited

1. Haberman, Charles E., "Pyrolysis of Scrap Tires Using the TOSCO II Process", prepared for The Rubber Division, American Chemical Society, Chicago Symposium, May 3-6, 1977.

MARCH 3, 1978.

Disposal of Sewage Sludge and Municipal Refuse by the Occidental Flash Pyrolysis Process

EUGENE J. CONE[1]

Occidental Research Corporation, 1855 Carrion Road, La Verne, CA 91750

For the last several years, Occidental Research Corporation has been interested in the production of synthetic oil and/or gas by pyrolysis of "waste" organic materials. Within the last two years, this interest has extended to disposal of sewage sludge and sewage sludge/municipal solid waste mixtures by pyrolysis. The emphasis of the current work has been on defining and optimizing product yields (i.e., controlling the mix of liquid, gas, or char products) and on controlling their composition. System variables being investigated include temperature, pressure, and residence time.

Objectives

Specific objectives of the research program are as follows:

1. Demonstrate in small pilot scale equipment that pyrolysis of sewage sludge can be accomplished by the Occidental Research Corporation Flash Pyrolysis™ process as practiced with municipal solid waste at El Cajon, San Diego, with only minor modification to existing equipment.
2. Define any interaction or "nonlinear" effects caused by copyrolysis of sewage sludge/municipal solid waste as compared to separate pyrolysis of the materials.
3. Define variables controlling product mix (liquid, gas, char).
4. Optimize product quality.

Experimental Apparatus

The pyrolysis system used in the experiments reported here was based on transport of sewage sludge or municipal solid waste through an electrically heated one inch nominal diameter pipe. Typical solids feed rates were 2 kg/hr, and typical carrier gas rates were 6.6 standard cubic meters per hour.

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[1] Current Address: Envirotech, One Davis Drive, Belmont, CA 94002

0-8412-0434-9/78/47-076-287$05.00/0

The system components consisted of a modified screw feeder, a 7 m long U-shaped section of electrically heated pipe, two cyclones, a glass fiber packed filter, a liquid product condensing section, and a gas flow sampling and metering section. Schematics of the system are shown in Figures 1 and 2. Products leaving the system were passed through a thermal oxidizer to eliminate any possible air pollution.

Two different liquid recovery systems have been used in our pyrolysis experiments. The first (Figure 1) consisted of two externally cooled "knock out pots", and the second (Figure 2) used direct contact spray quenching with an immiscible hydrocarbon liquid, followed by phase separation and product recovery. This second method of product recovery models the technology now in use at Occidental's municipal solid waste demonstration plant at El Cajon, in San Diego County, California; however, all data actually discussed here were taken with the first system.

Discussion of Experimental Results

The two feed materials used in these experiments were finely shredded municipal refuse as obtained from the "front end" of the Occidental Resource Recovery system, (1,2) and sewage sludge as obtained from the San Elijo, CA sewage treatment plant. The municipal refuse was usually reground in a hammer mill before pyrolysis so as to increase ease of handling in the small scale equipment used. The sewage sludge was air dried and then ground in the same mill. Typical particle size distribution for the pyrolysis feed materials are as shown in Table I.

TABLE I
PYROLYSIS FEED-PARTICLE SIZE DISTRIBUTION

Screen Opening, μm	Municipal Solid Waste - Percent Through Screen	Sewage Sludge - Percent Through Screen
841	96.8	100
297	90.4	77.0
149	82.4	54.4
105	77.2	50.6
75	66.4	36.2
45	44.0	26.8
37	36.6	20.2

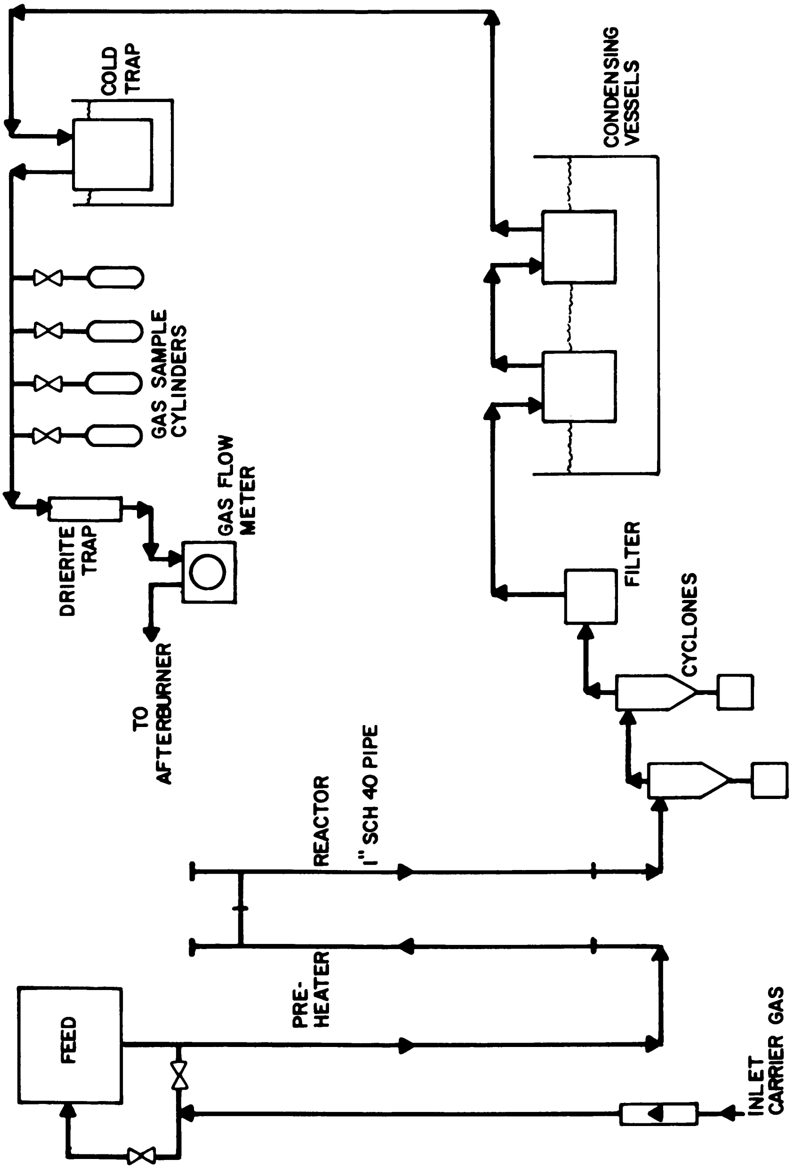

Figure 1. Pyrolysis pilot plant initial design

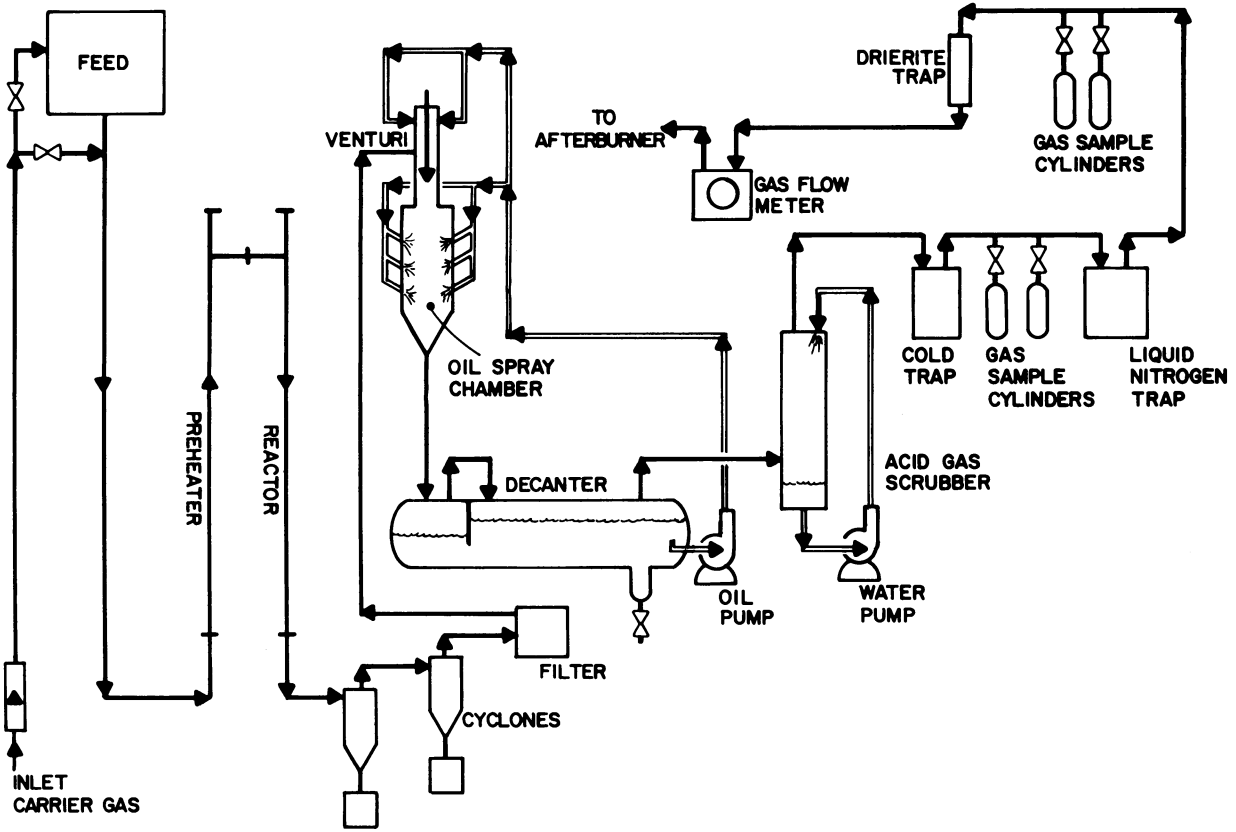

Figure 2. Pyrolysis pilot plant with improved product collection system

Clearly the sewage sludge particles used in these experiments tend to be larger than the corresponding municipal solid waste particles. Thus, the median sewage sludge particle is about 100 µm whereas the median municipal solid waste particle is about 50 µm. This should tend to make the sewage sludge used here more difficult to pyrolyze, but the exact magnitude of the effect has not been determined.

Typical feed compositions are as shown in Table II.

TABLE II
PYROLYSIS FEED COMPOSITION

Weight Percent	Municipal Solid Waste	Sewage Sludge
C	41.9	29.9
H	6.3	5.0
N	0.5	2.5
S	0.3	1.0
Cl	0.4	0.1
Ash	12.4	44.0
H_2O	4.0	6.0
O (by difference)	38.2	18.4

Note that the significant difference in ash content between municipal solid waste and sewage sludge makes it necessary to compare product yields in terms of moisture and ash free (MAF) materials.

Figure 3 shows a plot of oil yield as a function of temperature for pyrolysis of sewage sludge and municipal solid waste. Data is, of course, on a MAF basis for both feed and products. Note that the oil yield for sewage sludge is about 10% lower than that for municipal solid waste at 900°F, but essentially the same over the rest of the temperature range investigated. Figures 4 and 5 show similar plots for gas and char yields, respectively. Gas yield for sewage sludge is about the same at 900°F and about 10% lower over the rest of the temperature range. Char yield for sewage sludge is about 10% higher over most of the pyrolysis temperature range, but only about 4% higher at 1600°F.

Composition of pyrolytic oil is shown in Table III. Again, data is on an MAF basis. Note that the sewage sludge oil formed

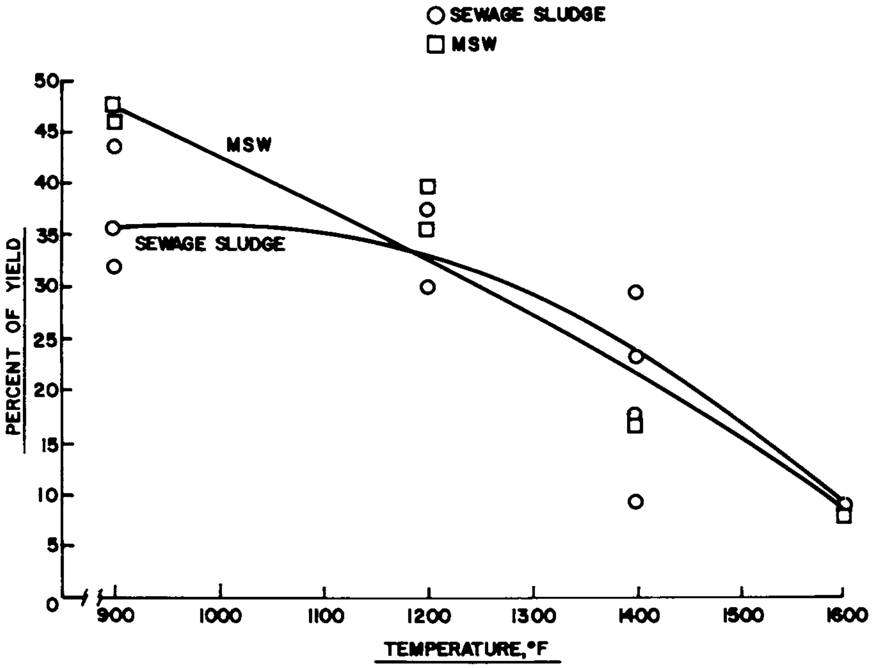

Figure 3. Yield of oil vs. temperature for pyrolysis of sewage sludge and municipal solid waste (MSW) (data based on MAF oil and feed)

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

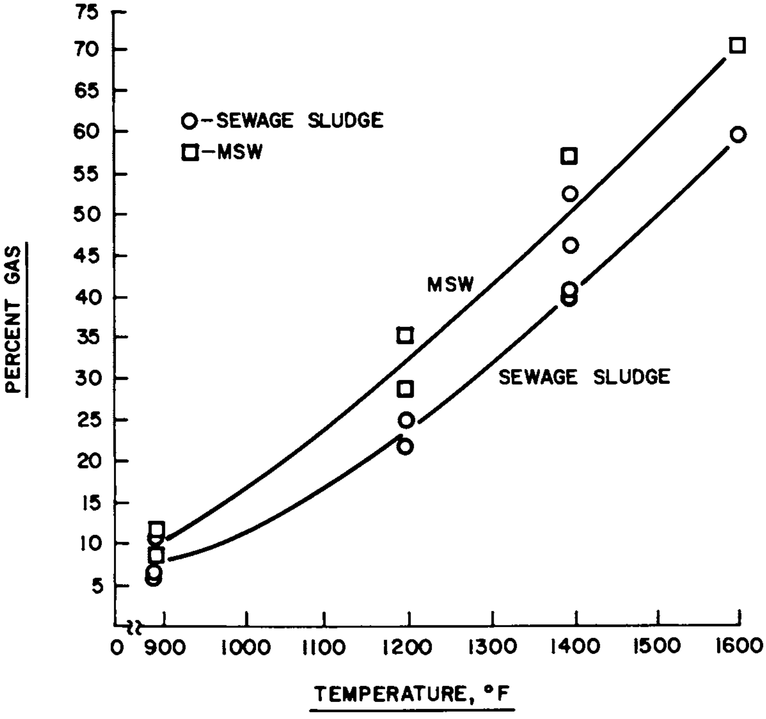

Figure 4. Yield of gas vs. temperature for pyrolysis of sewage sludge and municipal solid waste (MSW) (data based on MAF feed)

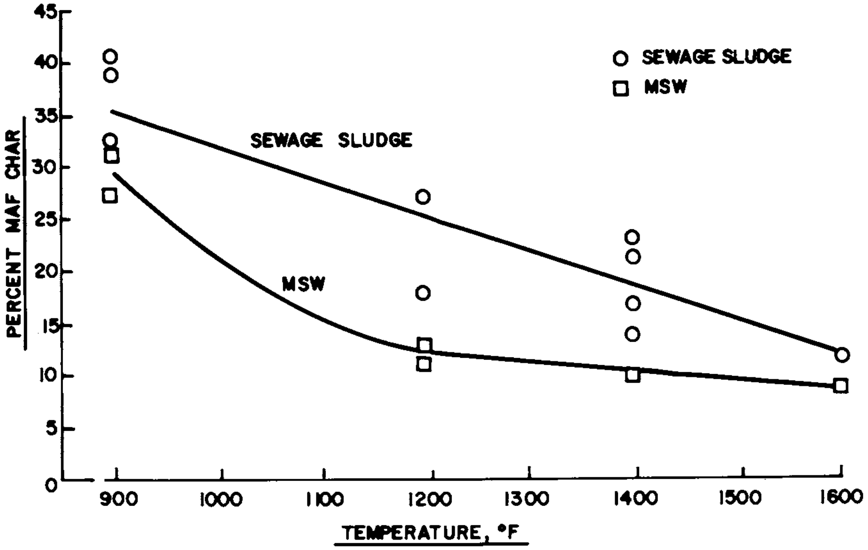

Figure 5. Yield of MAF char vs. temperature for pyrolysis of sewage sludge and municipal solid waste (MSW) (data based on MAF feed and char)

TABLE III
PYROLYTIC OIL PROPERTIES (MAF OIL)

Pyrolysis Temperature, °F	Oil From Municipal Solid Waste							Oil From Sewage Sludge						
	C	H	N	S	Cl	HHV (BTU/lb)	Run No.	C	H	N	S	Cl	HHV (BTU/lb)	Run No.
900	54.3	6.8	0.6	0.1	0.2	9286	5	62.6	9.9	5.7	0.7	0.2	13372	24
	52.4	6.6	0.7	0.1	0.2	8684	6	69.2	9.5	6.4	0.6	0.1	14501	29
								66.4	8.4	6.6	0.5	0.2	13114	35
1200	60.1	7.0	1.3	0.2	1.0	10781	7	65.7	8.6	9.4	1.1	0.1	13211	36
	60.7	7.6	0.8	0.1	0.4	11243	19	69.9	9.5	7.1	0.8	0.2	14736	28
1400	64.2	6.5	2.0	0.2	1.3	11394	17	66.3	8.4	11.9	1.4	0.3	13206	26
								69.7	7.9	12.5	1.5	0.1	13635	31
								70.5	7.6	11.9	1.3	0.1	13588	32
1600	78.0	10.3	3.1	0.3	1.0	17080	18	65.0	7.8	12.6	1.9	0.3	12579	27

TABLE IV
CONCENTRATION OF MAJOR COMPONENTS IN PRODUCT GAS FROM PYROLYSIS OF SEWAGE SLUDGE

Pyrolysis Temperature, °F	900	900	900	1200	1200	1400	1400	1400	1400	1600
Volume %										
H_2	3.2	7.3	0	7.7	9.6	12.9	13.0	12.3	24.0	18.0
CO	16.1	20.7	14.6	24.2	31.3	24.3	31.4	32.4	28.0	23.9
CO_2	68.1	54.9	79.3	28.6	33.9	14.7	20.8	19.5	15.6	12.8
CH_4	2.3	3.5	0	8.3	2.0	11.5	10.7	10.7	10.0	13.7
C_2H_4	1.0	2.0	0.8	8.4	7.7	15.7	9.5	9.9	11.8	18.9
Gas/oil residence time (sec.)	2	2	0.3	0.3	0.3	0.3	0.5	0.5	2	0.3
Run number	24	29	35	28	36	26	31	32	34	27

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

TABLE V
CONCENTRATION OF MAJOR COMPONENTS IN PRODUCT GAS FROM PYROLYSIS OF MUNICIPAL SOLID WASTE

Pyrolysis Temperature, °F	900	900	1200	1200	1400	1600
Volume %						
H_2	4.7	6.7	13.0	9.5	15.4	19.6
CO	33.6	34.5	49.3	50.7	49.7	48.4
CO_2	55.2	52.6	15.3	18.8	9.4	7.5
CH_4	0.3	2.7	8.5	7.5	10.4	12.0
C_2H_4	5.0	1.1	4.8	4.0	7.3	7.9
Gas/oil residence time (sec.)	0.5	0.5	0.5	0.3	0.3	0.3
Run number	5	6	7	19	17	18

in the 900-1200°F range is generally higher in carbon and hydrogen than the corresponding municipal solid waste oil, and thus is apparently a high quality oil. The nitrogen content might be considered a problem but blending with municipal refuse oil and fuel oil can produce a composite fuel with nitrogen content below 1%. Table III also shows the higher heating value of pyrolytic oil from municipal solid waste and sewage sludge calculated from the Dulong-Petit equation. As would be expected from the relatively large percent of carbon and hydrogen in the sewage sludge oil, its higher heating value compares favorably with that of municipal solid waste oil for oils formed at or below 1400°F.

Table IV shows the concentration of the major components in the product gas from pyrolysis of sewage sludge, and Table V, similar data for municipal solid waste. Figures 6-10 present this data graphically.

Examination of these tables and figures shows that, as expected, gas composition varies appreciably with pyrolysis temperature. Variation in, and absolute value of hydrogen and methane concentrations are seen to be about the same for sewage sludge and municipal solid waste over the entire temperature range investigated. Carbon monoxide concentration is seen to be about 20% lower for sewage sludge, and carbon dioxide concentration is about 15% lower in the 900-1200°F range, and several percent lower in the 1400-1600°F range. Ethylene concentration is considerably higher for sewage sludge gas, particularly at higher pyrolysis temperatures. Table VI compares the calculated average gross heating value for product gas from the two feed materials. It should be noted that these values appear higher than might be expected because of the presence of several percent C_5+ in the gas stream.

TABLE VI
PYROLYSIS PRODUCT GAS GROSS HEATING VALUE

Temperature, °F	Municipal Solid Waste (BTU/SCF)	Sewage Sludge (BTU/SCF)
900	260	410
1200	680	960
1400	670	880
1600	600	960

Tables VII and VIII show product recoveries for sewage sludge, and municipal solid waste, respectively. It is particularly interesting to note that overall, carbon, hydrogen and ash material balances generally close to within ±10%.

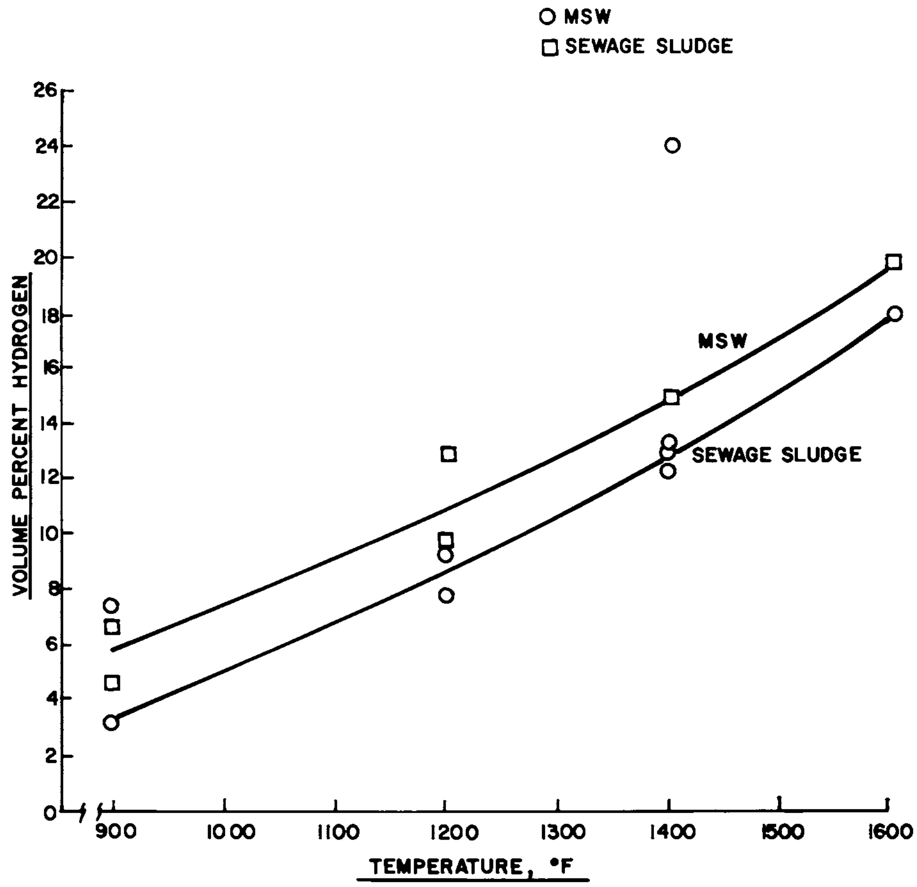

Figure 6. Volume percent of H_2 in pyrolysis product gas for pyrolysis of sewage sludge and municipal solid waste (MSW)

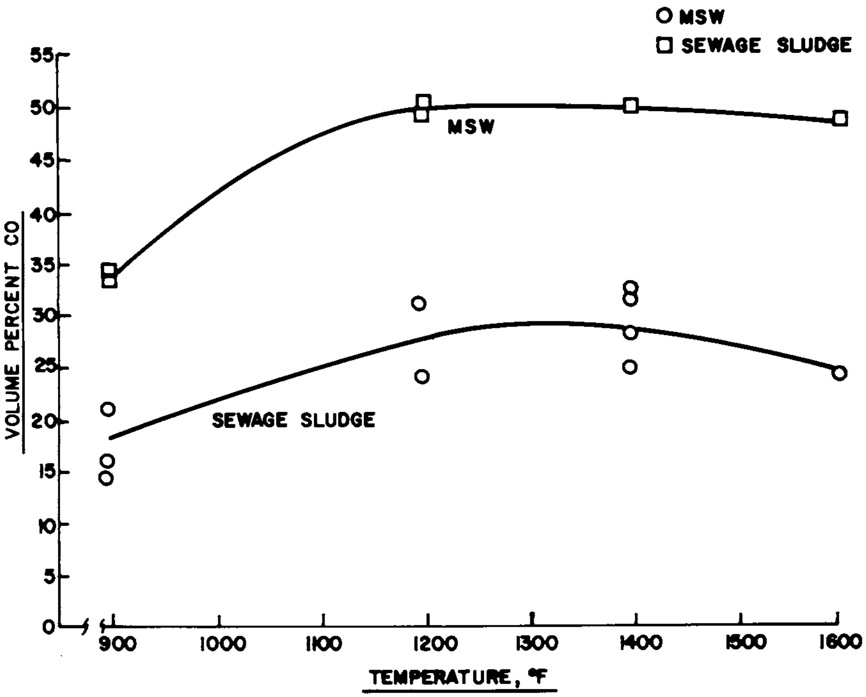

Figure 7. Volume percent of CO in pyrolysis product gas for pyrolysis of sewage sludge and municipal solid waste (MSW)

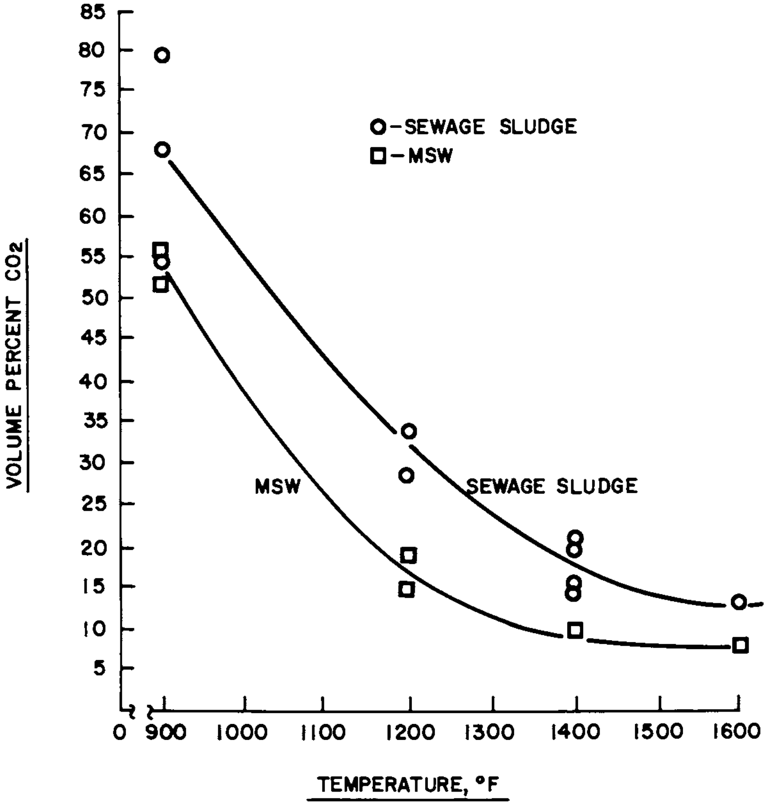

Figure 8. Volume percent of CO_2 in pyrolysis product gas for pyrolysis of sewage sludge and municipal solid waste (MSW)

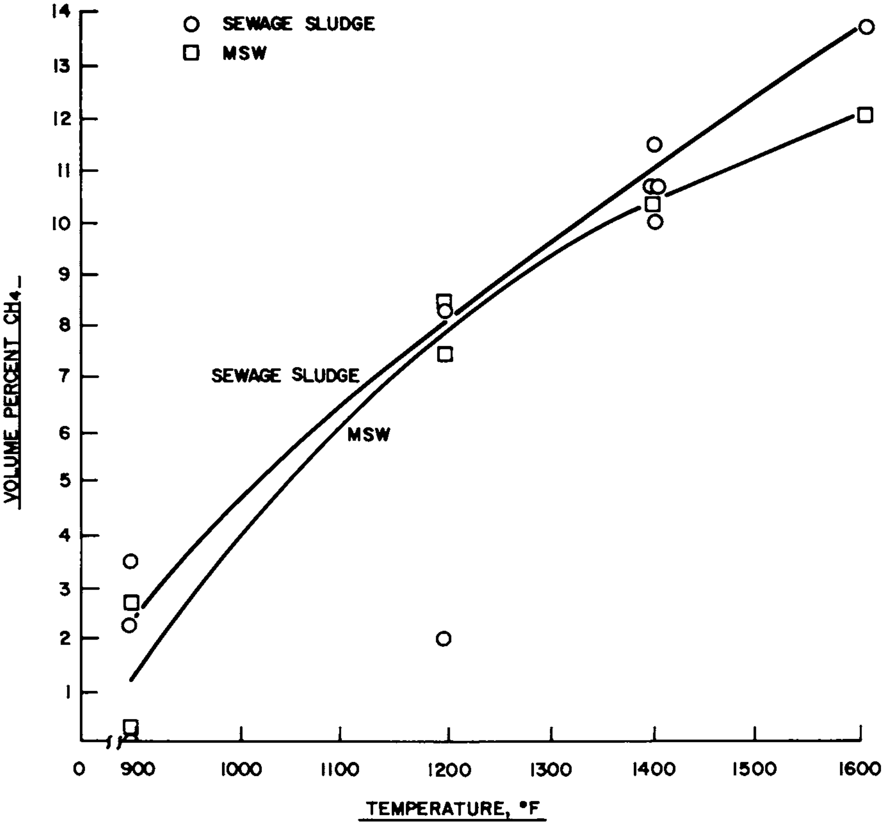

Figure 9. Volume percent of CH_4 in pyrolysis product gas for pyrolysis of sewage sludge and municipal solid waste (MSW)

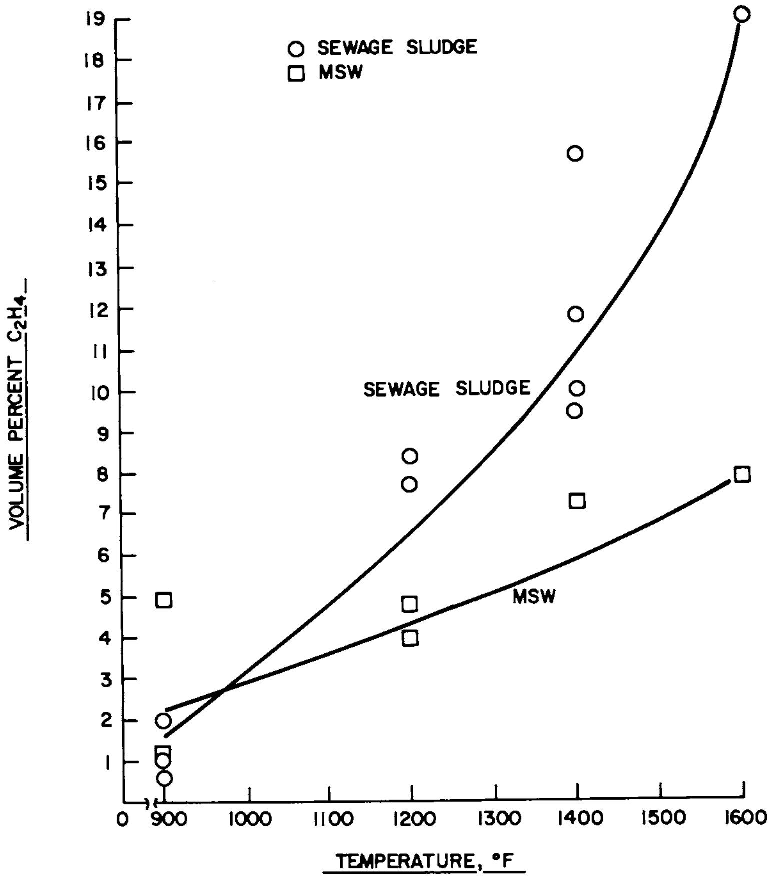

Figure 10. Volume percent of C_2H_4 in pyrolysis product gas for pyrolysis of sewage sludge and municipal solid waste (MSW)

TABLE VII
PYROLYTIC PRODUCT RECOVERY FOR SEWAGE SLUDGE PRYOLYSIS

Pyrolysis temperature, °F	900	900	900	1200	1200	1400	1400	1400	1400	1600
Weight % MAF oil based on MAF feed	36.3	43.4	32.3	37.5	29.8	17.2	23.0	29.2	8.9	8.6
Weight % gas based on MAF feed	6.9	10.7	6.4	21.7	24.8	40.8	40.4	46.2	53.5	59.3
Weight % MAF char based on MAF feed	40.8	32.6	39.1	17.8	27.2	14.1	23.1	29.9	16.6	11.6
Water of pyrolysis--%	5.3	7.0	16.0	3.2	11.6	7.9	12.6	3.5	11.8	8.4
Weight % dry oil based on as used feed	18.5	20.2	12.8	19.3	13.0	8.9	9.3	11.2	3.7	4.5
Weight % gas based on as used feed	3.5	5.0	2.6	11.1	10.3	31.1	16.2	17.6	21.9	30.7
Weight % dry char based on as used feed	60.7	62.1	65.9	51.4	59.1	49.2	60.8	59.5	55.7	47.4
Weight % water based on as used feed	8.5	8.3	11.4	6.2	9.5	10.0	10.3	9.3	10.0	9.9
Run number	24	29	35	28	36	26	31	32	34	27
Nominal gas/oil residence time (sec.)	2	2	0.3	0.3	0.3	0.3	0.5	0.5	2	0.3
% Overall recovery	91.2	95.7	92.6	88.0	91.9	89.1	96.5	97.6	91.3	92.5
% C recovery	89.1	104.9	86.4	87.4	98.6	94.2	100.1	104.0	92.1	90.3
% H recovery	86.6	97.1	95.7	83.7	99.6	93.0	109.7	98.3	108.6	99.9
% Ash recovery	92.4	97.0	91.1	96.0	90.0	98.6	94.2	95.7	91.8	97.4

TABLE VIII
PYROLYTIC PRODUCT RECOVERY FOR MUNICIPAL SOLID WASTE PYROLYSIS

Pyrolysis temperature, °F	900	900	1200	1200	1400	1600
Weight % MAF oil based on MAF feed	46.1	47.6	35.8	39.9	17.0	8.2
Weight % gas based on MAF feed	11.4	8.8	35.4	28.9	57.0	70.7
Weight % MAF char based on MAF feed	27.5	31.5	13.2	11.1	10.4	8.4
Water of pyrolysis--MAF feed	14.4	14.5	14.1	15.9	15.1	12.8
Weight % dry oil based on as used feed	39.5	29.9	29.9	33.4	14.4	7.0
Weight % gas based on as used feed	9.6	7.3	29.6	24.1	47.3	59.0
Weight % dry char based on as used feed	33.1	35.1	20.5	19.9	18.2	16.1
Weight % water based on as used feed	16.1	16.1	15.8	19.0	19.1	16.8
Run number	5	6	7	19	17	18
Nominal gas/oil residence time (sec.)	0.5	0.5	0.5	0.3	0.3	0.3
% Overall recovery	98.0	98.3	95.9	96.3	98.9	98.9
% C recovery	97.6	96.5	98.4	99.3	99.0	101.3
% H Recovery	96.8	98.9	96.3	105.4	103.7	112.7
% Ash Recovery	105.8	72.0	102.0	99.9	93.6	88.4

The conclusion from all the above is that copyrolysis of sewage sludge and municipal solid waste seems quite possible. The sludge pryolyzes quite readily during Flash Pyrolysis^{TM} to give apparently reasonable products. No special problems in the pyrolysis of sewage sludge are apparent at this time.

Abstract

Air-dried and ground municipal sewage sludge, and municipal solid waste have been pyrolyzed in a 2 kg/hr pilot plant. An electrically heated vertical flow transport reactor with short residence time was used. Temperatures in the range 755-1144 K (900-1600°F) were investigated. Oil, gas, and char yields and composition varied appreciably with pyrolysis temperature. The results presented here suggest that the Occidental Flash Pyrolysis^{TM} process can be applied successfully to the co-disposal of municipal sewage sludge with municipal refuse.

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Variable Velocity Fluidized Bed for Pyrolysis of Biomass

STEVEN R. BECK, WILLIAM J. HUFFMAN,[1] UZI MANN, and HO-DER WONG
Department of Chemical Engineering, Texas Tech University, Lubbock, TX 79409

Current interest in converting solid hydrocarbon resources to liquid and gaseous products has spawned the development of many new technologies and the resurrection of some old technologies. One such technology is pyrolysis in a fluidized bed to produce gas that can be used as fuel or as a feedstock for petrochemical products. Processes utilizing fluidized bed technology are in varying stages of development for coal gasification, oil shale gasification and biomass gasification (1, 2, 3, 4, 5). Most of these processes employ steam and air or oxygen as the fluidizing medium. In the vicinity of the distributor, part of the solid is combusted to provide the heat to pyrolyze the remaining solid. The pyrolysis zone in the reactor is generally above and separate from the combustion zone. At the top of the reactor is an expanded disengaging section to prevent excessive solids entrainment in the produced gas.

A fluidized bed designed for partial oxidation of solid hydrocarbons presents some interesting design problems. The different reactions which occur throughout the bed can lead to severe longitudinal temperature profiles unless the bed is very well mixed. In addition, the pyrolysis of solids results in an increase in the superficial mass flow rate of gas as the reaction proceeds. The combination of these effects leads to a fluidized bed in which the gas velocity varies throughout the bed even in the constant diameter section of the column. Add to this the reduction in gas velocity due to an increase in diameter in the disengaging section and the resulting problems in scale-up of the process are enormous. Figure 1 illustrates how the gas velocity can vary in a fluidized bed used for pyrolysis.

Synthesis Gas From Manure Process

At Texas Tech University, the Synthesis Gas From Manure (SGFM) Process has been under development for seven years (6, 7,

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[1] Current Address: Battelle Memorial Institute, Columbus, OH

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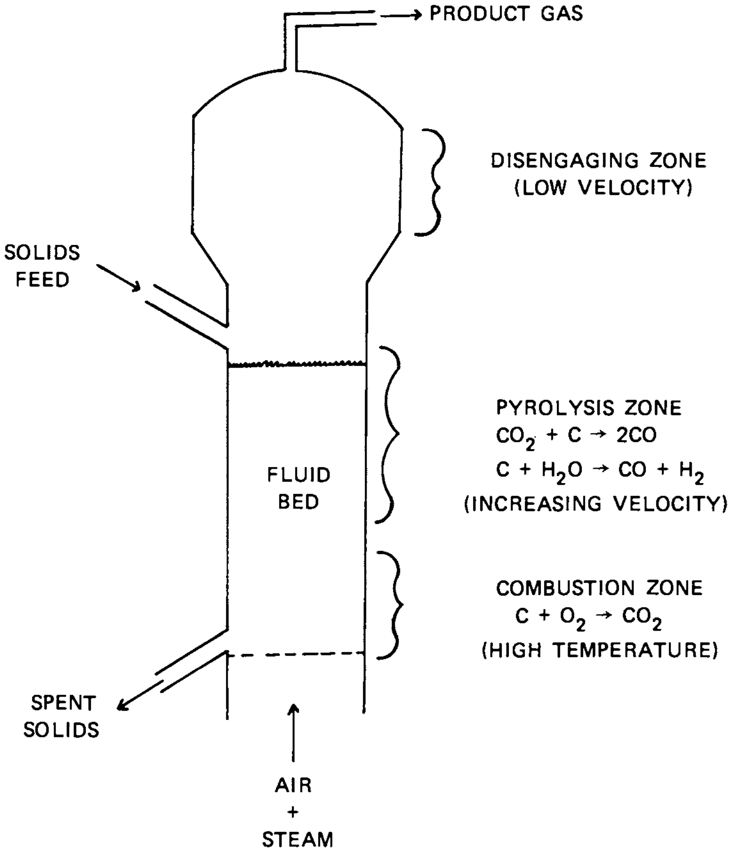

Figure 1. Fluidized bed pyrolysis reactor

8). The original objective of this project was to develop a process by which cattle feedlot manure could be converted to a synthesis gas suitable for ammonia production. This would provide a means by which the nitrogen content of the manure could be transported for fertilizer applications at locations far distant from the feedlot. The project is now investigating thermochemical gasification of other biomass feedstocks to produce a medium-BTU (1.12×10^7 to 1.86×10^7 J/m^3) gas suitable for direct combustion or further synthesis.

The SGFM process is based on a counter-current fluidized bed for pyrolysis of the biomass. The solid feed is introduced at the top of the reactor and is fluidized by air and steam injected at the bottom of the reactor as shown in Figure 2. This process is currently in the pilot plant stage of development. The reactor shown in Figure 2 has a capacity of ~500 Kg/day manure feed and was scaled-up from a 2.5 cm x 45 cm laboratory reactor. This pilot plant has been operational for two years during which a design data base was developed for thermochemical gasification of manure. A complete description of the pilot plant has been previously reported (9).

Internal Gas Samples

As part of the data base development for the SGFM pilot plant, a series of internal reactor samples were taken to determine gas concentration profiles within the reactor. These samples were obtained by inserting sample probes at 30.5 cm intervals along the length of the reactor. The sample probe was 0.64 cm (0.5 cm ID) tubing with twenty-five ports along one side of the tubing. The probe section with ports traversed the entire diameter of the reactor. Hence, each sample is thought to represent an average, radial concentration. During sampling, the probe was connected to a condenser to remove steam and liquids. The dried sample was then routed through a rotameter which was followed by a gas sample bottle.

The first step in obtaining a sample was to purge the entire sample system and sample bottle with helium. The sample velocity was also set using helium. Then, the actual sample was obtained by opening a ball valve. The gas was allowed to purge through the sample bottle for three to six residence times. As the gas sample purged the sampling system, the gas velocity would tend to decrease due to plugging in the probe and lines. The gas velocity was adjusted to the desired rate by use of a needle valve just prior to the sample bottle. Two gas samples were obtained at each location to check the reproducibility of the sampling and gas analysis.

The sample velocity relative to the isokinetic sample velocity and residence time in the sampling system were areas of concern in regard to the accuracy of the results. The residence time of the gas sample in the hot section (>400°C) of the sample probe

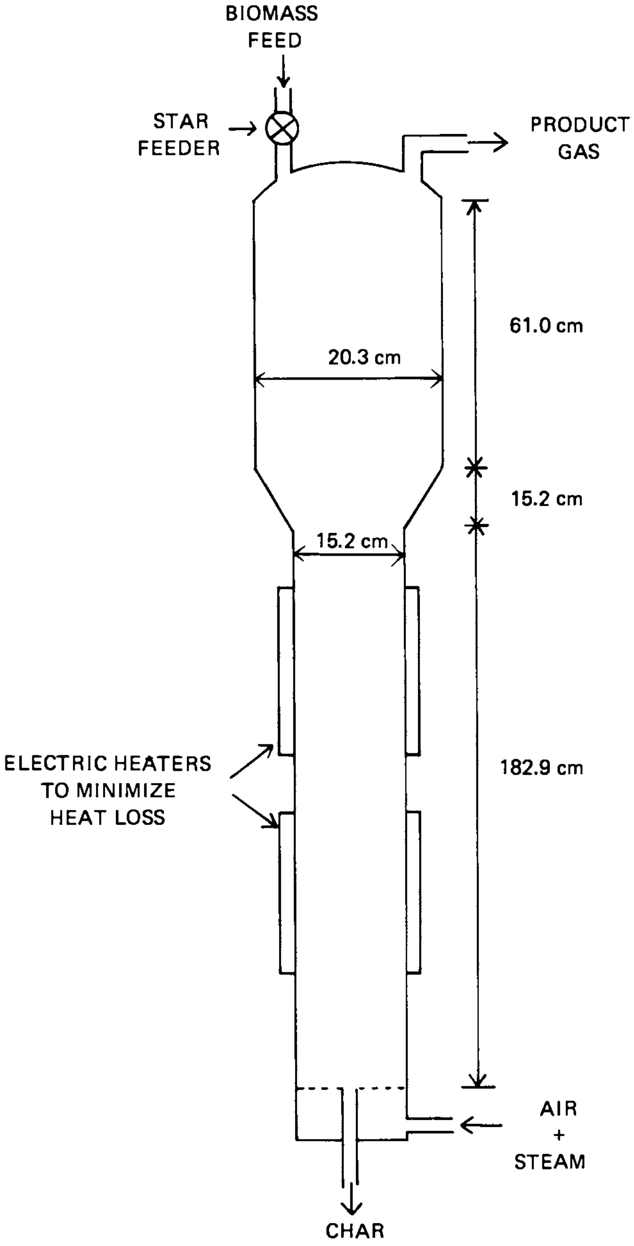

Figure 2. SGFM pilot plant reactor

was less than 0.2 seconds which minimized reactions in the probe. To determine the effect of sample velocity relative to isokinetic velocity, samples were taken at different rates. Table I shows some typical results obtained in these tests. The sample velocity appeared to have negligible effects on the gas compositions.

The complete results of the internal sample data will be reported at a later date. For this paper, the nitrogen concentration throughout the reactor is reported. Figure 3 shows the nitrogen concentration as a function of height in the pilot plant reactor for five runs. Since nitrogen does not react, the decrease in nitrogen concentration can be related to an increase in total gas mass flow rate due to pyrolysis gas formation. In addition to the varying gas flow rate, the SGFM reactor shows a wide variation in temperature along the reactor as shown in Figure 4.

Part of this temperature profile is due to the fact that the lower heater on the reactor (see Fig. 2) was inoperable. Consequently, very high heat loss was experienced in the lower part of the reactor while the heat loss in the center section of the reactor was minimized by the upper electric heater. Even if both heaters were operable, a significant longitudinal temperature gradient would have been present. The change in mass flow rate of gas and the temperature profile results in the gas velocity increasing 2-4 times in the reactor as shown in Figure 5.

The widely varying gas velocities in the SGFM reactor present problems in trying to scale-up the reactor. Fluidized beds are extremely difficult to scale-up and still be able to predict the bed performance even when the velocity is constant. In the case of the SGFM reactor, no methods to scale-up a variable velocity are known to be completely reliable. For this reason, a program was initiated to study a variable velocity fluidized bed.

Variable Velocity Fluidized Bed

The objective of this project was to build and operate an ambient temperature, non-reacting, gas-solid fluidized bed to investigate the characteristics of a variable velocity fluidized bed reactor such as the SGFM pilot plant reactor. To obtain the desired variation in velocity, the bed could have been constructed with a changing cross-sectional area and fluidized by a constant volumetric gas flow rate or a constant diameter vessel could have been operated with a subliming solid or a porous solid containing a volatile solvent. The former method was chosen due to its ease of operation and the lack of mass transfer effects between phases which would have complicated the analysis of the results. This approach should be considered as a first attempt to gain an understanding of the phenomena involved and must be followed by studies in a constant diameter bed using a volatile component.

The equipment used for this study was a two-dimensional, Plexiglas^R bed as shown in Figure 6. The inlet of the bed is 7.5 cm x 45.7 cm and decreases to 7.5 cm x 15.2 cm at the high velocity

TABLE I. EFFECT OF SAMPLING VELOCITY ON INTERNAL GAS CONCENTRATIONS, RUN 19

Height Above Distributor, cm	Sample Rate, ℓ/min	N_2	CO	H_2	CH_4	CO_2	C_2H_4	C_2H_2	O_2
45.7	0.7-1.2	38.4	7.8	14.3	6.3	28.4	3.6	0.6	0.6
	1.0-1.1*	33.6	7.3	20.8	6.4	27.2	3.4	0.7	0.5
	~ 0.6	37.8	8.1	16.6	5.1	27.6	3.6	0.5	0.6
152.4	0.7-1.2	24.3	12.6	17.6	8.2	26.5	4.8	0.4	0.5
	0.5-0.8	29.4	13.4	17.8	8.4	25.2	4.8	0.5	0.4
	0.24-0.6	28.2	14.7	17.9	8.7	24.6	4.9	0.5	0.4
	~ 0.25	29.3	13.6	18.2	8.3	25.0	4.6	0.5	0.5

The isokinetic sample rate is estimated to be 0.6 ℓ/min.

*A reactor upset occurred while this sample was being obtained.

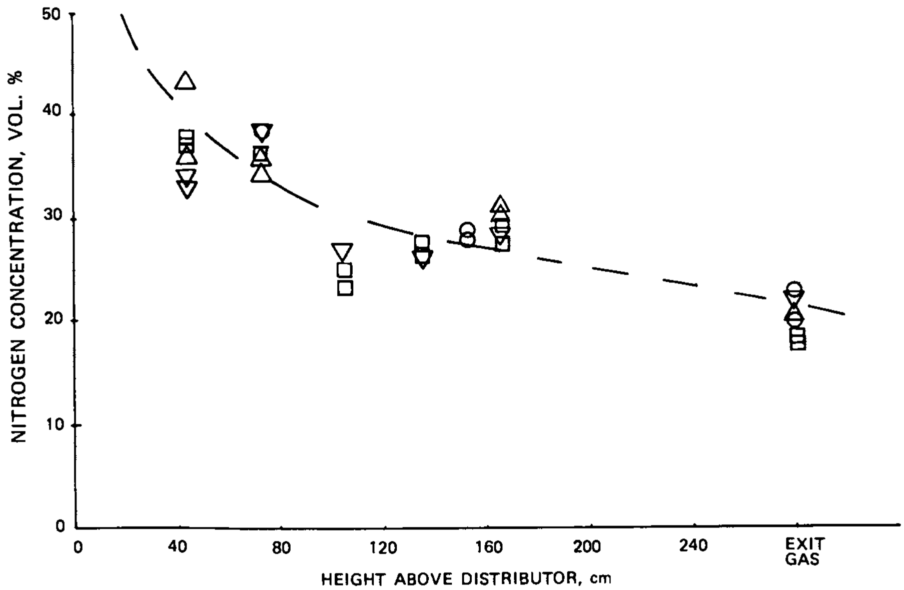

Figure 3. Nitrogen concentration in the SGFM reactor: ○, Run 19; △, Run 20; □, Run 21; ▽, Run 23

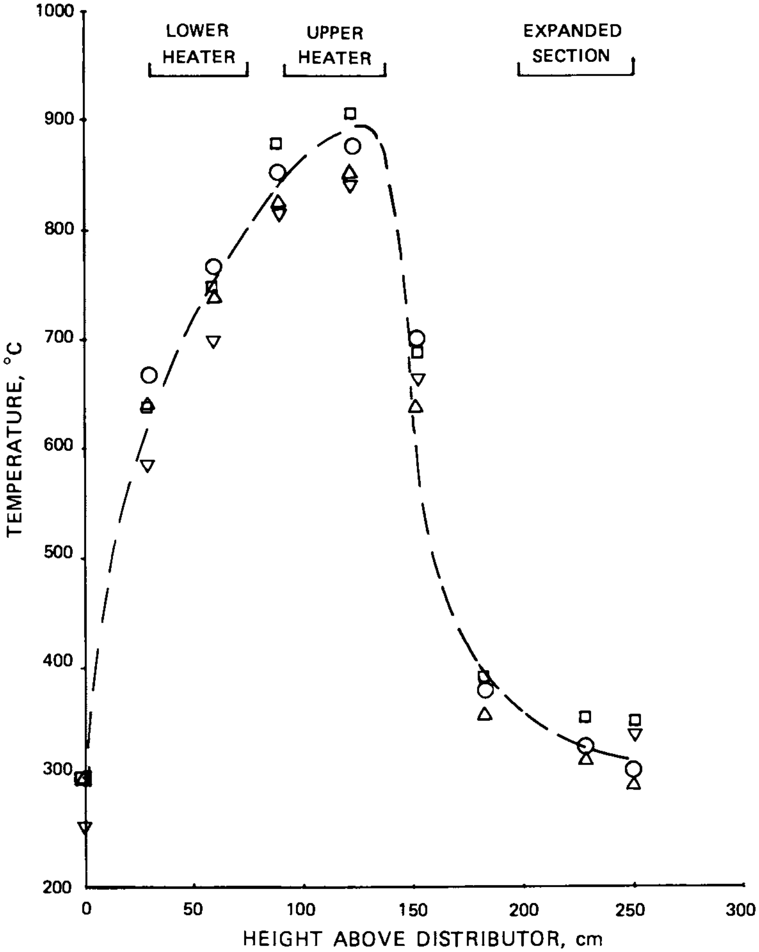

Figure 4. Temperature profile in the SGFM reactor: ○, Run 20; △, Run 21; □, Run 22; ▽, Run 23

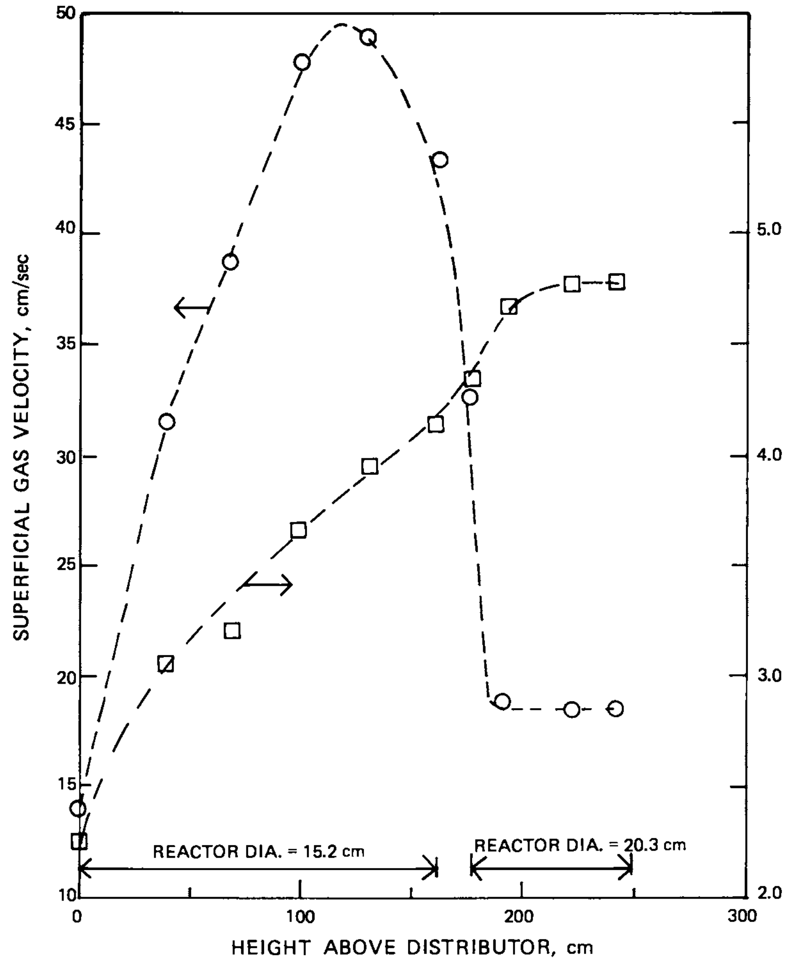

Figure 5. Typical gas velocity profile in the SGFM reactor, Run 22: ○, gas velocity; □, mass flow rate

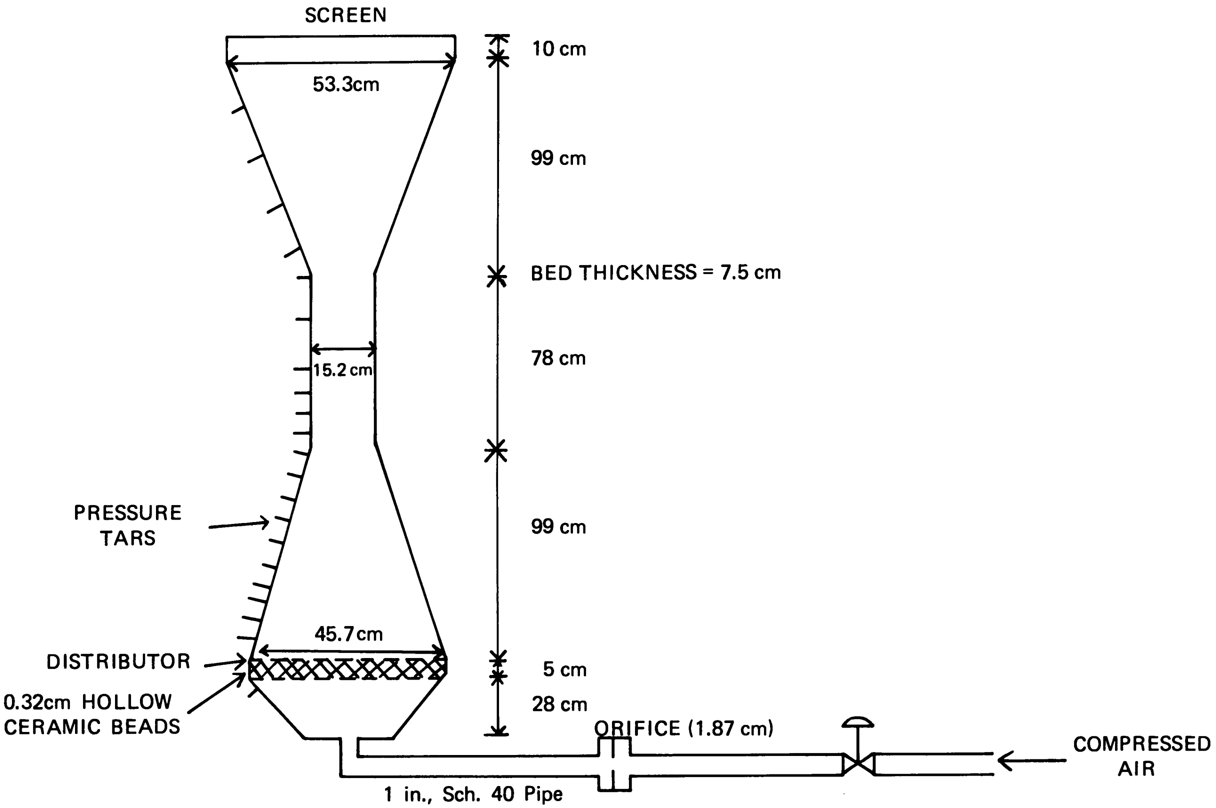

Figure 6. Cold model, variable velocity, fluidized bed

section. The upper portion of the bed expands to 7.5 cm x 53.3 cm at the exit. Pressure taps are placed at approximately 5 cm intervals along the bed. Pressure drop is measured by both vertical U-tube and inclined manometers. The distributor plate is Plexiglas^R with 1/32" holes on a 1/2 inch triangular spacing. Fluidizing air is supplied by a Hybon Model HB-50 compressor with a maximum flow of 1700 ℓ/min at 1 atm and 15°C which corresponds to a superficial gas velocity in the bed of 70.9 cm/sec at the entrance depending on pressure. The air flow rate is measured by pressure drop across a 1.87 cm orifice in a 1 in., Sch. 40 pipe which was calibrated with a turbine meter.

The longitudinal velocity profile created by the varying cross-sectional area is very similar to that measured in the SGFM pilot plant reactor. This mode of operation may have certain advantages in pyrolysis reactors. This gas velocity will be relatively low in the combustion zone of the reactor, high in the pyrolysis zone and low in the disengaging section. The high velocity in the pyrolysis zone will lead to low residence times for the hydrocarbon gases produced and minimize reforming or cracking of these gases. This should result in a gas product with a higher heating value.

Another advantage of this system could be three distinct fluidization regimes in the reactor. The low gas velocity zones at the top and bottom would result in dense phase fluidization and the high velocity pyrolysis zone would be a dilute phase fluidized bed. This concept, shown in Figure 7, has been qualitatively investigated in the cold model fluidized bed.

Results

Initial fluidization tests were conducted with 1.2 cm styrofoam cubes in the bed. This was chosen as the solid phase because it was readily fluidized and the large particles were easy to observe. The preliminary results are qualitative, but do indicate that the concept of a three-stage fluidized bed is feasible.

To conduct each test, a batch of solids was loaded in the bed to a predetermined static bed height. The air flow was then gradually increased until the bed was fully fluidized and particles began entering the high velocity section of the equipment. This flow was maintained until the lower and upper sections of the bed reached a steady-state solids content determined by visual observation. Figure 8 shows the steady-state solids distribution in the system that developed from a 53.3 cm static bed height.

Figure 8 is an excellent example of the staged, fluidized bed concept. The lower section of the bed behaves as a dense phase, gas-solid fluidized bed. As can be seen in the picture, the high velocity, center portion of the bed contains very few solid particles. The upper portion of the bed is the most interesting. The solids content is very similar to a dense phase

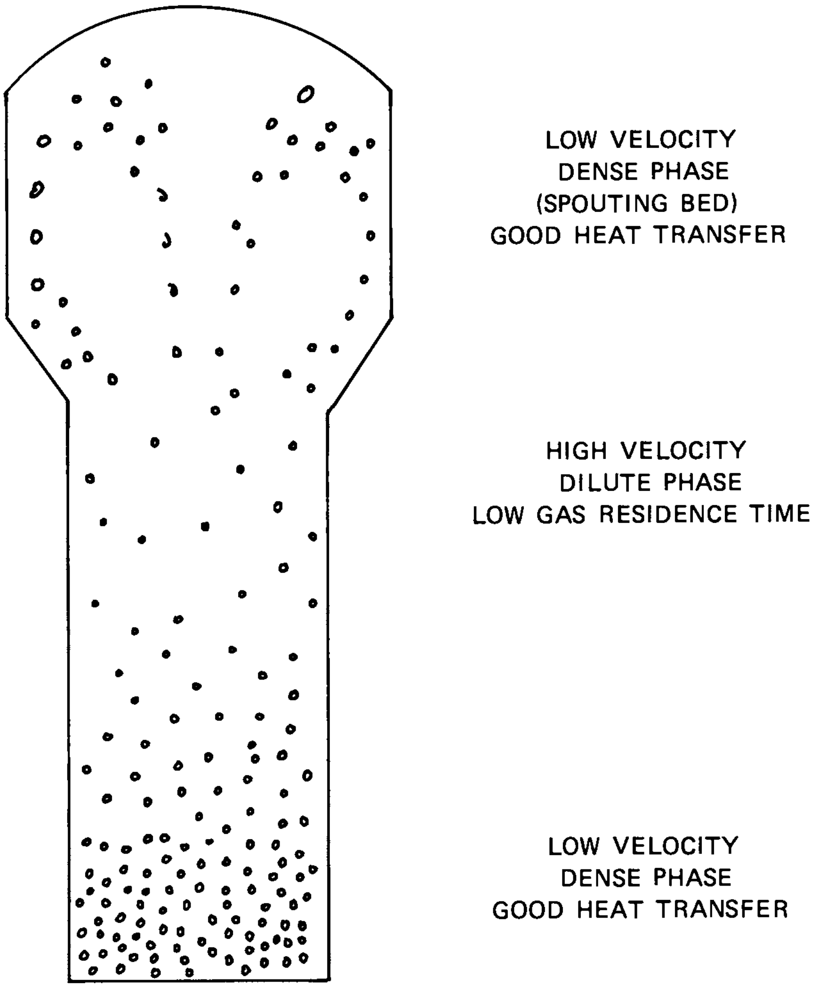

Figure 7. Three-stage fluidized bed concept

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

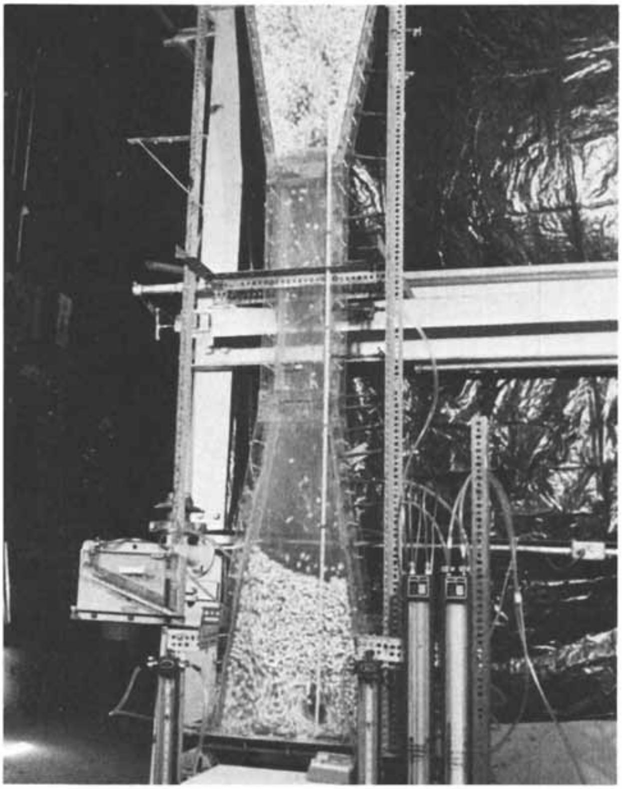

Figure 8. Steady-state solids distribution for styrofoam showing three distinct fluidization regimes

fluidized bed, but there is no distributor plate helping to support the bed. In fact, this upper portion behaves more like a spouting bed with solids moving downward near the wall and being entrained in the center of the bed. This behavior is like a spouted bed except that no draft tube, downcomer or distributor plate is present. This proves that a dense phase, fluidized bed can be supported by a high velocity air stream.

The geometrical configuration of the expanded top section of the apparatus could have contributed to this spouting bed effect. However in a more conventional type design, the transition section between the lower section and upper, disengaging section of a fluidized bed will have sloped walls. This would mean that a spouted bed could be established even in a fluidized bed of standard design.

Following the styrofoam experiments, manure was placed in the bed. The particles were much smaller (<2 mm) which created many problems. The first to appear was trying to fluidize a broad particle size range. The result of this was that the small particles were elutriated and the large particles did not fluidize. This is a potential problem which must be considered when designing fluidized bed pyrolysis reactors. To alleviate this problem, the manure was screened to narrow particle size fractions and additional tests were conducted on a particle size of -0.50 mm + 0.25 mm. The manure exhibited the same staged, fluidized bed behavior as the styrofoam with an additional phenomenon observed.

The spouted bed formed in the upper section of the reactor, but some of the manure recirculating downward on the walls continued to fall down the walls of the high velocity section and returned to the lower section of the bed. This can be seen in Figure 9.

It is felt that the manure was able to fall down the walls due to the low gas velocity in the boundary layer. This was not observed with the styrofoam which is probably because the particles were too large and were exposed to higher velocities on the outer edge of the boundary layer. These manure experiments demonstrated that there was solid circulation from the bottom bed to the top and back to the bottom. This is very critical to the SGFM reactor because solids are fed to the top of the reactor and must be able to reach the lower section of the reactor in order to react. The cold bed studies indicate that this is possible.

Conclusions & Discussion

The results reported in this paper have been qualitative in nature. When the project was initiated, there was no way to reliably predict the bed behavior and the first step in the study was to determine if the concept of a variable velocity, staged, fluidized bed was feasible. Consequently, the experiments were designed to obtain a maximum of qualitative information in order

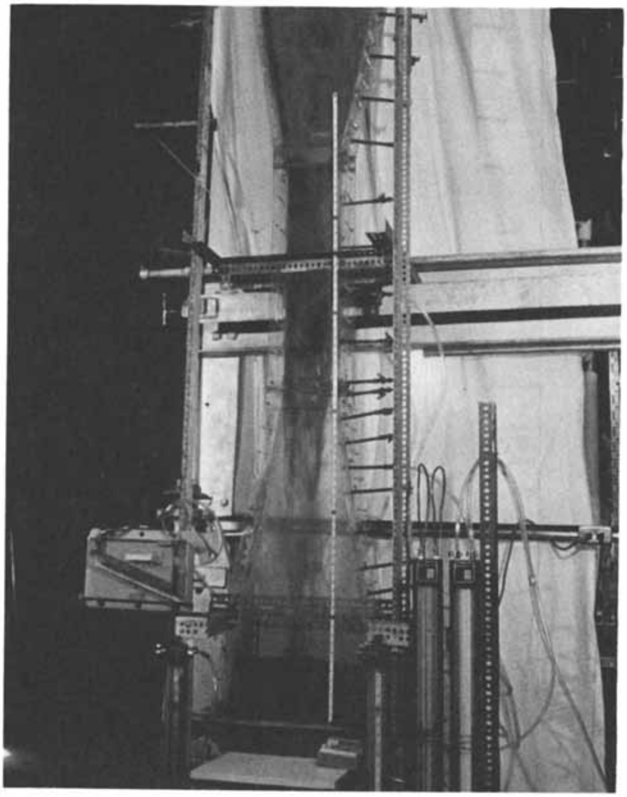

Figure 9. Steady-state solids distribution for manure showing manure from upper bed "raining" down into lower bed

to plan the second phase of the project which is quantitative in nature. The conclusions of the first phase are as follows:

1. The three-stage (dense phase, dilute phase, dense phase) fluidized bed concept is feasible.
2. Solid particles can fall through the high velocity portion of the SGFM pilot plant reactor and thus reach the lower section of the fluidized bed.

The study of variable velocity, fluidized beds is continuing at Texas Tech University. The goal of this study is to obtain sufficient information to develop the correlations required to scale-up pyrolysis reactors with some degree of reliability. This will reduce the time required to develop a process from the laboratory scale to a commercial plant.

Acknowledgements

The authors wish to thank the U. S. Department of Energy for supporting this work under Contract No. EY-76-S-04-3779.

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EPA's R & D Program in Pyrolytic Conversion of Wastes to Fuel Products

GEORGE L. HUFFMAN and WALTER W. LIBERICK, JR.

U. S. Environmental Protection Agency, Industrial Environmental Research Laboratory,
Cincinnati, OH 45268

Only in recent years have we become properly concerned about the final disposition or utilization of the large volumes of wastes generated by our everyday activities. Many environmental insults have resulted from improper disposal; for example, the obvious insults from open burning, leaching from dumps, and improper discharges to surface and groundwaters. These insults led to passage of the 1965 Solid Waste Disposal Act and the subsequent Amendments that mandate the Federal Government (EPA) to support research, development, and demonstrations and applications of new and improved technologies for processing and recovering materials and energy from solid wastes. This mandate, coupled with recent concerns about our dwindling energy supply, has focused our attention on the utilization of these wastes as sources of energy. Although recovering energy-from-waste has been practiced in Europe for many years, this technology has not been pursued in the United States until recently.

Many options to utilize these wastes as an energy resource have been investigated. Options have varied from combustion in waterwall incinerators, to co-firing of prepared wastes with coal in utility boilers, to the thermochemical conversion of wastes into usable fuel products.

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This paper focuses on current efforts being supported by EPA to develop new concepts of thermochemical (or pyrolytic) conversion of wastes to energy. More specifically, this paper discusses four EPA projects that are investigating basic reaction kinetics and ways to enhance the yield and quality of pyrolysis end-products; these are:

- Subpilot-scale pyrolysis of mixed wastes to fuel gas, fuel oil and char in a 500 lb/hour fluidized bed reactor (by the Energy Resources Company, ERCO);
- Bench-scale steam gasification of agricultural and feedlot wastes to energy products (by Princeton University);
- Laboratory tests on a process that converts a troublesome waste, discarded tires, into fuels and other products by utilization of a molten salt catalytic technique (by the University of Tennessee); and,
- Integrated bench-scale development of a non-catalytic process that transforms pyrolysis off-gases into polymer gasoline (research being conducted via Interagency Agreement with China Lake, California, Naval Weapons Center).

Subpilot-Scale Conversion of Mixed Wastes to Fuel (ERCO)

The objectives of this project are to investigate the kinetics of the pyrolysis of solid wastes and to develop a data base that will allow design engineers to select pyrolysis system operating conditions for commercial systems that will provide a predetermined mix of fuel product yields. To accomplish these objectives, this project is divided into the following tasks:

1. Plan and Experimental Study.

   The initial task included a literature study to determine the physical and chemical properties of the wastes to be utilized and to determine the reactor variables which would have a significant influence on product quality and yields. From this study, the project team determined the experiments to be run and the data necessary to model the results. This task included reactor and support equipment design.

2. Equipment Procurement, Assembly and Check-Out.

   This task included the fabrication of the test apparatus (a fluidizedbed reactor) and its check-out. During this task, several modifications were made to increase system operating reliability.

3. Subpilot Plant Testing.

   This task included acquisition of data on several waste components, some of which were taken during tests run in different operating modes. These data were compiled for use in modeling the test results of this project.

4. Development of the Prediction Model.

   Based on the data collected in Task 3, a mathematical model was developed to predict the yields to be expected from a known waste composition under various pyrolysis operating conditions. The theoretical model was then verified with the experimental data.

5. Chemical Conversion.

   In addition to the variable pyrolysis operating modes investigated, an effort was undertaken to look at partial oxidation and steam gasification effects.

6. Product Characterization.

   Detailed product characterization tests were conducted on samples of the oil and char products.

7. Trace Metals Investigation.

   A material balance for trace metals was completed for several of the runs on different waste components.

An extensive literature search was conducted to: (1) physically and chemically characterize solid waste; (2) determine the parameters which would affect pyrolysis yields; and (3) determine the theoretical kinetics to be used in modeling the reactions.

The literature search provided the data contained in Table I. These average chemical compositions of municipal solid waste were selected as a base-line for this study (1). Actual analyses were of course used in the material balances throughout the work.

TABLE I

AVERAGE CHEMICAL COMPOSITION

				COMPOSITION OF AVERAGE REFUSE IN Lb/100 Lb Dry Solids(2)							
CATEGORY	WT % "AS FIRED"	WT % "AS DISCARDED"	% MOISTURE "AS DISCARDED"	Lb MOISTURE	Lb ASH	Lb C	Lb H_2	Lb O_2	Lb S	Lb N_2	Wt(dry) (lb)
Metal	8.7	8.2	2.0	0.22	10.13	0.50	0.067	0.481	0.0011	0.0056	11.19
Paper	44.2	35.6	8.0	3.97	2.74	20.70	2.781	19.193	0.0547	0.1368	45.59
Plastics	1.2	1.1	2.0	0.03	0.17	0.90	0.125	0.285	0.0045	0.0150	1.50
Leather and Rubber	1.7	1.5	2.0	0.04	0.24	1.23	0.170	0.390	0.0062	0.0205	2.05
Textiles	2.3	1.9	10.0	0.27	0.08	1.10	0.152	0.995	0.0048	0.0523	2.38
Wood	2.5	2.5	15.0	0.52	0.09	1.43	0.178	1.260	0.0033	0.0089	2.96
Food Waste	16.6	23.7	70.0	23.10	2.17	4.13	0.574	2.730	0.0248	0.2772	9.90
Yard Waste	12.6	15.5	50.0	10.79	0.54	5.31	0.701	3.890	0.0378	0.3129	10.79
Glass	8.5	8.3	2.0	0.23	11.21	0.06	0.008	0.041	-	0.0034	11.32
Miscellaneous	1.7	1.7	2.0	0.05	1.62	0.30	0.046	0.278	-	0.0696	2.32
Average	100.0	100.0		39.72	28.99	35.66	4.802	29.543	0.1372	0.9022	100.00

To determine the parameters that would affect product yields, several efforts reported in the literature were reviewed (1). Table II is a summary of this work. It was determined that reaction temperature is the dominant parameter in most systems. Also, ultimate yields seem to vary from one type of reactor to another (fixed bed, entrained flow, and fluidized bed). Another conclusion was that secondary reactions (reactions occurring in the gas phase following initial pyrolysis) have a significant impact on final yields. It is suspected that the secondary reactions take the form of a further breakdown of the tars into char and gases. After analysis of the results of previous work, it was determined that the fluidized bed reactor would be ideal for studying chemical kinetics because its excellent heat and mass transfer characteristics result in good temperature uniformity. The conclusion that secondary reactions were important in ultimate yields forced the investigators at ERCO (and their co-workers at MIT) to include these reactions in their modeling effort.

Based on the literature search, it was determined that the reaction kinetics should be controlled by the composition of the solid waste and its temperature/time history while undergoing pyrolysis. The wastes investigated in the ERCO project are shown in Tables III and IV. The independent variables for testing are shown in Table V. The resulting test plans for pyrolysis are shown in Tables VI and VII. The test plan for partial oxidation, steam gasification and catalytic effects runs is shown in Table VIII. Table IX summarizes tests conducted during the initial phase of this project (1).

The experiments were conducted in two reactors (a batch type and a fluidized bed). The batch reactor (see Figure 1) was utilized early in the program to: (1) perform preliminary runs on different wastes to identify potential problems before assembling the fluid bed; (2) develop sampling and analysis techniques; (3) compare data with values reported in the literature; and (4) measure varying product yields (1).

After gaining experience with the batch unit, a fluidized bed reactor was constructed (see Figure 2). The reactor was designed to accommodate 200 Kg/hr of feed waste materials with the capability to measure those parameters necessary to determine reaction kinetics. The system consisted of: (1) a heat source; (2) an inert gas supply to maintain an oxygen-free environment; (3) inert fluidizing media; (4) a gas distribution and bed support system; (5) a continuous feeder; (6) the product collection system; (7) a gas cleanup and after-burning facility; (8) sampling and analysis equipment; and (9) safety and control equipment. Figure 3 shows a schematic of the fluid media section of the reactor.

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

TABLE II

SUMMARY OF PREVIOUS WORK

REACTOR TYPE	INVESTIGATOR	MATERIAL	TEMPERATURE RANGE
Fixed Bed			
Retort	Kaiser and Friedman (2)	Newspaper	Ambient to 1500°F
Retort	Hoffman and Fitz (3)	MSW	900°F, 1200°F, 1500°F, 1700°F
Retort	Cities Service Oil Co. (4)	MSW	Ambient to 1500°F
Retort	Georgia Inst. of Tech. (5,6)	Bark, Sawdust	800°F to 1300°F
Retort	U. S. Bureau of Mines (7)	MSW	900°F to 1700°F
Retort	Folks, et al. (8)	Sewage Sludge	1100°F to 1840°F
Retort	Firestone (9)	Tires	900°F to 1700°F
Entrained Flow			
Recirculation of Hot Char	Garrett Research Corp. (10)	Rice Hulls, Manure, Bark, Straw	800°F to 1000°F
Freefall	Univ. of Calif. (Berkeley)(11)	Wood, MSW	900°F to 1800°F
Fluidized Bed			
Combustion Gas	Itoh, et al. (12)	MSW	450°C to 850°C
Hot Nitrogen Gas	Imperial College (13)	Sawdust, Cellulose, Sucrose	400°C
Combustion Gas	West Virginia Univ. (14)	MSW, Sawdust, Sludge and Others	1280°F to 1520°F

TABLE III

SEGREGATED WASTES

WASTE COMPONENT	MATERIAL	SOURCE
Wood	Pine Sawdust	Local Dealer
Wood	Pine Woodchips (1/2-1 in. long)	Local Dealer
Agricultural Wastes	Ground Corncobs (20-40 mesh)	Local Dealer
Agricultural Wastes	Manure	Local Dealer
Paper	Punches (1/8-1/4 in.) from Notebook Manufacture	Courtesy of National Blank Book Co., Springfield, Mass.
Textiles	Shredded Cotton Cuttings	Local Dealer
Plastics	Polyethylene Beads (1/8 x 3/8 in.)	Local Dealer
Tires	Metal and Fabric Free Shredded Tires--2 Sizes (-10 mesh and -4+10 mesh)	Solid Conversion Systems, Inc. Burlington, N.J.
Waste Oil	Automobile Crankcase Oil	Local Dealer

TABLE IV

WASTE MIXTURES

SYNTHETIC MIXTURE	APPROXIMATE COMPOSITION
Paper + Sawdust	50/50
Paper + Plastics	50/50
Corncobs + Manure	35/65
ACTUAL WASTES	SOURCES
Municipal Solid Waste	National Center for Resource Recovery, Washington, D. C., glass and metal free, pelletized
Sewage Sludge	Sewage treatment plant, City of Marlborough, Mass.

TABLE V

INDEPENDENT VARIABLES

VARIABLE	NUMBER OF VALUES
BATCH REACTOR	
Feed Material Composition	18
Reactor Temperature	3
Heating Rate	3
Particle Size	2
TOTAL POSSIBLE COMBINATIONS	324
FLUIDIZED BED REACTOR	
Feed Material Composition	18
Reactor Temperature	3
Liquid Recycle	2
Fluidization Velocities	2
Particle Size	2
Bed Height	2
TOTAL POSSIBLE COMBINATIONS	864
TOTAL COMBINATIONS FOR TWO REACTORS	1,188

TABLE VI

TEST PLAN

PARAMETERS AND TESTS ON MATERIALS REQUIRING COMPLETE COVERAGE

Fluidization Velocity		Bed Height		Feed Rate		Temperatures (°C)																				
						Sawdust			Paper			Mixture 1: Paper and Sawdust			Mun. Solid Waste 1			Corncobs			Sludge 1			Pyrolytic Oil		
Low	High	Low	High	Low	High																					
●		●		●		450	540	740	{520 520	705	872}	{555	725	785 880}	550	685	—	—	728	840	450	535 600	675	570	620	—
●		●			●	465	—	766	{510 495	825 750	760 860}	427 410	740	795	560	685 610	720	555	722	850	482	520	655	—	660	—
●			●	●		{633 463	636 640 570	733 671 700}	485	560	770	400 410	420 605	760 795	{550	760	801 800}	{657 643	753	939}	—	—	—	—	—	—
●			●		●	{375 405	550 430	670 560 900}	485	575	750	—	600	—	{610	710 750}	800 810	{430	560 680	660 923}	—	600	755	—	—	—
	●	●		●		{500	580	650 710}	535	753	853	555	770	830	530	725	—	590	670	831	490	625	—	—	700	—
	●	●			●	—	—	—	{535 540	610 770	700 750}	525	780	—	600	692	705	530	710	—	515	655	—	—	660	—
	●		●	●		{480 680	500 700	815 555 900}	485	675	780	—	610	—	{490 550	645	800}	{540 693	789 770	672 851}	—	620	—	—	—	—
	●		●		●	—	—	—	470	550	750	—	600	—	{510 730	760 787	825 870}	732	760	815	—	—	—	—	—	—

TABLE VII

TEST PLAN

SCREENING RUNS USING ADDITIONAL MATERIALS

MATERIAL	FLUIDIZATION VELOCITY	TEMPERATURES (°C)					
Manure	Low	475		618		809	
	High	723		804		921	
Tires, Small	Low	500		454		___	
	High	454		630	810	930	920
Tires, Large	Low	546		614		___	
	High	517		595	727	932	
Wood Chips (7% Moisture)	Fixed	550 935	610	675	825	866	
Wood Chips (17% Moisture)	Fixed	571	740	847	856	916	
Wood Chips (24% Moisture)	Fixed	641		781		863	
Textiles	Fixed	___		___		___	
Plastics	Fixed	640	690	706	770	800	877
Mixt. #2 (Paper + Textiles)	Fixed	___		___		___	
Mixt. #3 (Paper + Plastics)	Fixed	620	650	715	760	765	
Mixt. #4 (Corncobs + Manure)	Fixed	550	675	757	875	974	
Mixt. #5 (Paper + Plastics + Sawdust)	Fixed	___		___		___	
Mun. Solid Waste #2	Fixed	___		___		___	
Sludge #2	Fixed	___		___		___	

TABLE VIII

TEST PLAN

PARTIAL OXIDATION, STEAM GASIFICATION AND
CATALYTIC EFFECT RUNS

PARTIAL OXIDATION

Run Conditions

Material	Oxygen/Feed Ratio Low	Oxygen/Feed Ratio High	Oxygen Content Low	Oxygen Content High	Feed Rate Low	Feed Rate High	Temperatures, °C □		△		▽		○	
Sawdust	■ △		■ △		■	△	668	665	—	—	—	—	—	—
	▽ ○			▽ ○	▽	○	730		—		—		—	
		□ △	□ △		□	△	—	—	—	—	640	880	—	—
		▼ ○		▼ ○	▼	○			—		790		—	
Paper	■ ▲		■ ▲		■	▲	648	—	650	—	—	—	670	—
	▽ ●			▽ ●	▽	●	—		827		—		—	
		■ △	■ △		■	△	423	—	—	—	476	745	473	720
		▼ ●		▼ ●	▼	●	—		—		705	833	705	

STEAM GASIFICATION AT 35-50% STEAM CONTENT

Material	Oxygen Content 0%	5%	21%	Feed Rate Low	Feed Rate High	Temperatures, °C □	▽	⌂	⍒	♀	Y
Sawdust	□ ▼	♠ ⍒	♀ ▽	■ ■ ♀	▼ ⍒ Y	556 667 804	538 643 —	— 600 790	— 585 745	— — —	— — —
Paper	■ ▼	♠ ⍒	♀ Y	■ ♠ □	▼ ⍒ Y	490 650 760	515 680 795	515 710 810	— 510 700	— — —	— — —

TABLE IX

SUMMARY OF TESTS

Conditions	Number of Runs	Material and Energy Balances	Oil Analysis	Char Analysis	Trace Metal Samples[(a)]
Pyrolysis	262	262	39	35	14
Waste Oil	24	24	11	0	9
Steam Gasi-ication	20	20	3	4	0
Partial Oxidation	23	23	8	2	3
Combustion	16	16	2	2	0
	345	345	63	43	26

(a) Ten metals each

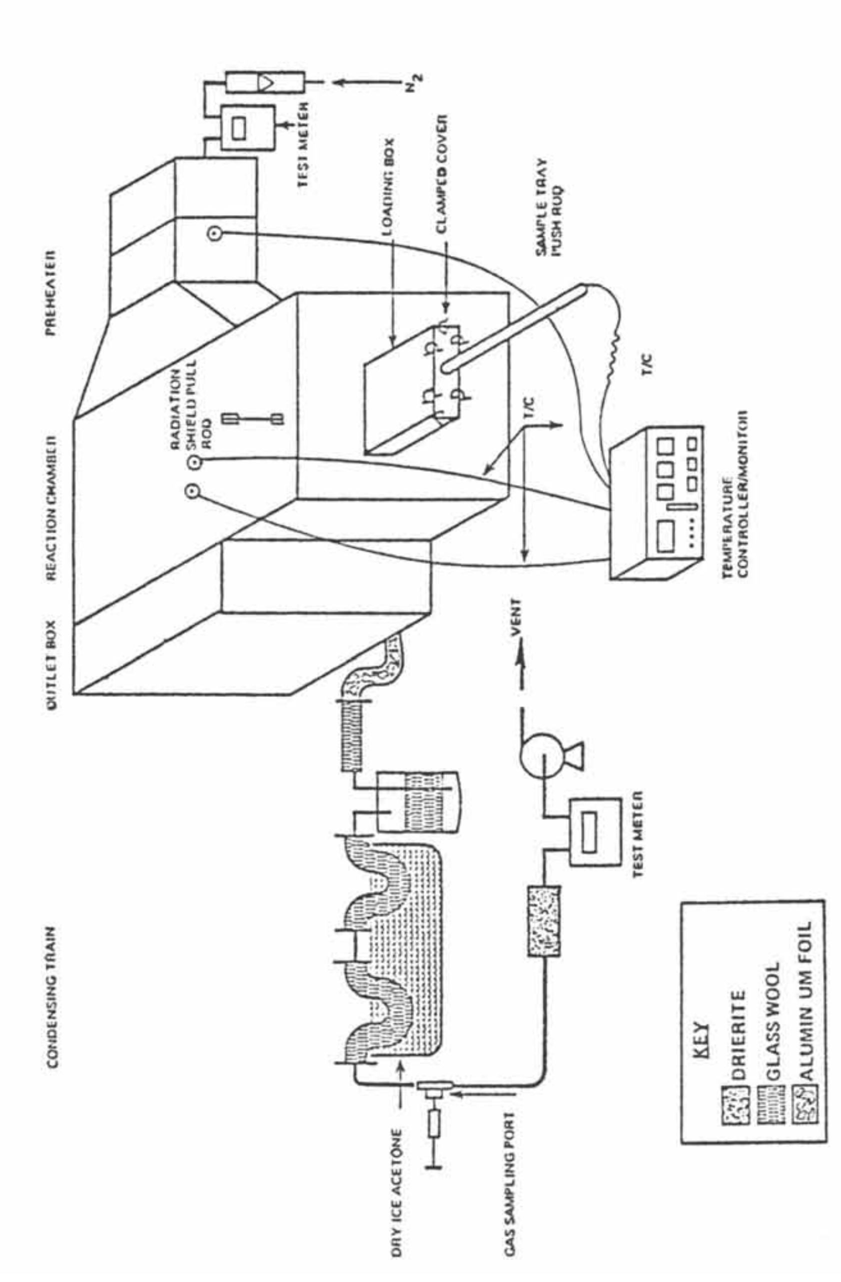

Figure 1. Batch reactor—ERCO project

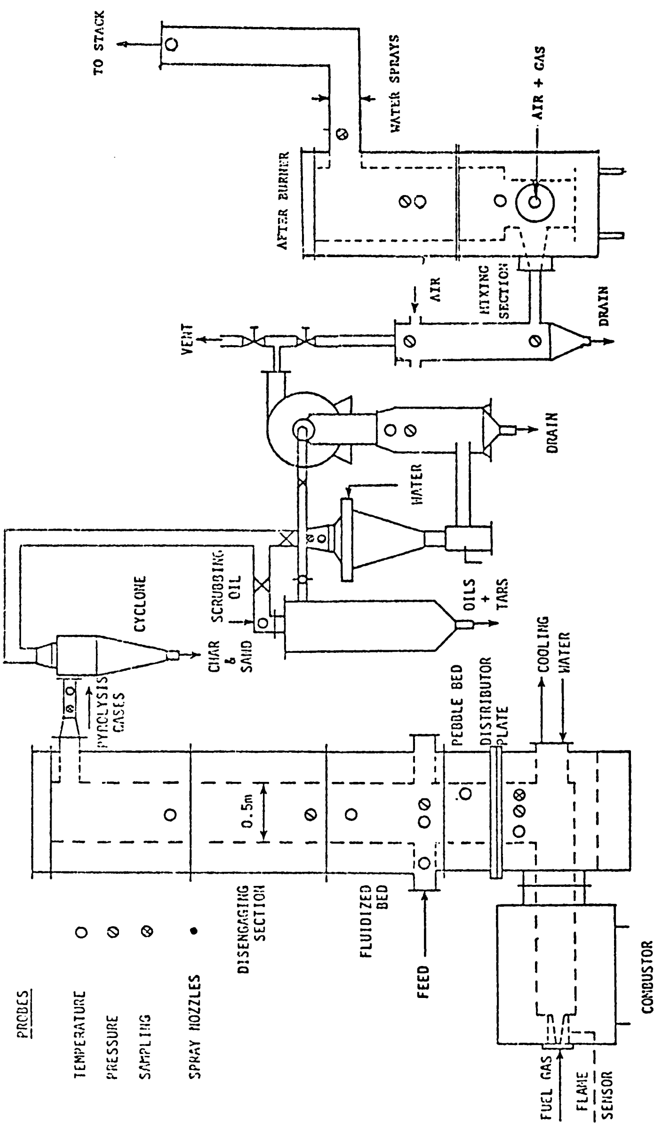

Figure 2. Energy resources pilot plant for fluidized bed pyrolysis of solid waste

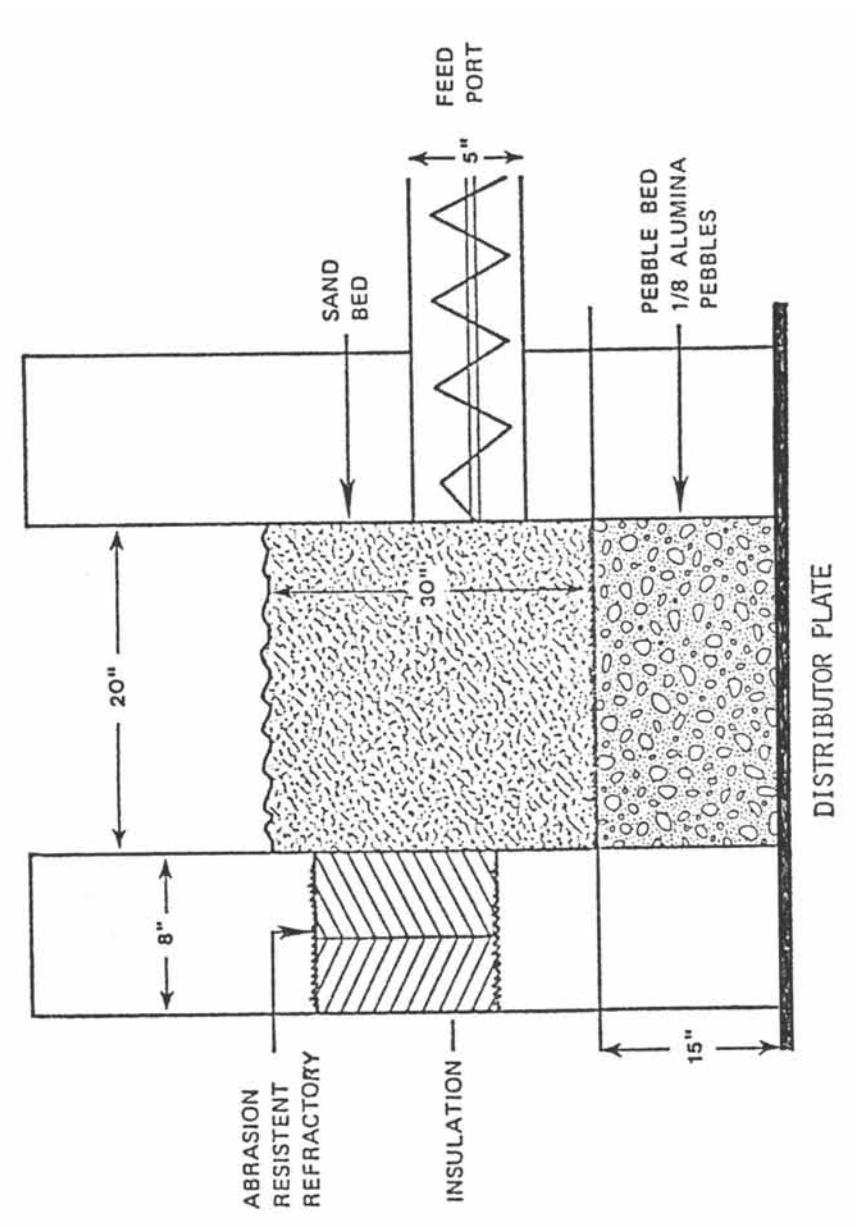

Figure 3. Fluidized bed—ERCO project

The operating procedure for each test is relatively simple. The fluidization velocity is controlled by regulating the flow of natural gas and air to the gas burner. Waste material is fed into the bed with a screw feeder. The sampling apparatus is operated during steady-state conditions for 2 or 3 minutes. At the end of each run, excess air is introduced to the reactor to burn out the char accumulated in the bed.

The data resulting from a run are fed into a computer program for appropriate conversions and manipulation resulting in a typical printout and is eventually reported in tabulations similar to that illustrated in Table X (1).

The major conclusion to be derived from this project concerns the influence of secondary reactions on product yields in the pyrolysis mode. As an example, with sawdust, the secondary reactions begin to dominate above 400°C. At temperatures above 400°C, because of the secondary reactions, the oil and tar produced in the primary decomposition is further broken down into additional gas and char. This conclusion applies to all the feedstocks tested in some 345 data runs and shows the predominant effects of reaction temperature on product yields.

Concurrent with the ERCO project, EPA is also supporting a small effort to evaluate the feasibility of multi-waste gasification. This evaluation is being done by Monsanto Research Corporation. One of the systems being analyzed is the proposed ERCO system. Based on very preliminary information, the economics of the ERCO-like system appear to be attractive. An analysis was done for three different sizes of systems. Tables XI and XII illustrate the projected economics (15).

The Pyrolysis of Organic Wastes in Inert and Reactive Steam Atmospheres (Princeton University)

Over the past year, the Environmental Protection Agency has also been supporting an investigation at Princeton University to determine optimal gasification conditions for producing usable fuel products by the pyrolytic conversion of solid waste in a reactive steam atmosphere. This project is complementary to our Energy Resources Corporation (ERCO) project. Princeton, too, is investigating the chemical phenomenon taking place in a steam environment, but their bench-scale research will look at the influence of elevated pressures on reactor yields as well.

The research at Princeton has shown that the steam gasification of organic wastes is a two-step process. At relatively low temperatures (300° to 400°C), organic wastes lose most of their volatile matter content by pyrolysis reactions. Because of the low temperatures at which pyrolysis occurs, the rates of these

TABLE X

PYROLYSIS OF SAWDUST

FILE NO.	TEMP. °C	FEED RATE Kg/hr	FLUIDIZING GASES			BED HT. (m)	MATERIAL YIELD, %				ENERGY YIELD, %		
			FLUID. VEL. (m/s)	O_2%	H_2O%		GAS	OIL	CHAR	WATER	GAS	OIL	CHAR
68	463	41	0.24	0.1	19.6	0.6	17	22	27	35	8	41	43
63	570	48	0.27	0.2	19.5	0.6	23	18	21	37	19	35	34
66	700	45	0.31	0.2	19.5	0.6	44	9	15	32	47	17	24
75	730	47	0.38	0.3	19.4	0.6	61	6	9	25	91	10	14
71	550	133	0.30	0.5	19.1	0.6	18	36	16	29	16	68	25
69	500	112	0.29	0.1	19.5	0.6	15	36	13	36	6	68	20
64	560	69	0.28	0.1	19.6	0.6	20	20	20	39	18	38	33
67	670	97	0.30	0.2	19.5	0.6	35	19	13	34	33	36	20
72	670	137	0.36	0.1	19.5	0.6	43	20	12	25	63	38	19
83	710	143	0.37	0.1	19.5	0.6	46	16	10	28	64	30	16
87	900	139	0.41	0.1	19.5	0.6	67	5	9	20	81	8	13
78	480	50	0.51	0.1	19.5	0.6	20	35	15	29	11	67	24
77	500	47	0.52	0.1	19.5	0.6	20	25	16	39	10	47	25
65	555	48	0.49	0.7	19.0	0.6	22	20	18	39	19	38	28
73	680	48	0.63	0.3	19.4	0.6	60	7	8	25	44	12	12
84	700	44	0.64	0.2	19.5	0.6	59	8	9	24	77	15	14

TABLE XI. CAPITAL AND OPERATING COST REQUIREMENTS, ERCO PROCESS (15)

System	Processing capacity, tons/day		
	750	1500	2250
ERCO			
Total capital investment, $	9.9M̄	17.2M̄	24.1M̄
Capital cost per yearly ton, $	44	38	36
Unit operating cost, $/ton[b]	16	14[a]	14[a]

[a] Values are identical due to the highly modularized approach to process scale-up.

[b] Includes no byproduct credits.

M̄ = millions of dollars.

TABLE XII. FERROUS METALS, ALUMINUM, FUEL GAS, & DROP CHARGE REVENUES IN $/TON, ERCO PROCESS[a] (15)

Gas price, $/M Btu	Drop charge, $/ton			
	0	5	10	15
		ERCO		
1.50	12.0	17.0	22.0	27.0
1.75	13.1	18.1	23.1	28.1
2.00[b]	14.3	19.3	25.0	30.0

[a] Revenues from ferrous metals and aluminum assumed at $5.01/ton.

[b] Price currently charged by some utilities to industrial natural gas consumers.

reactions are not affected by the presence of steam as a gasification medium (16). At higher temperatures (500° to 700°C), the volatile pyrolysis products react with the steam producing a medium Btu synthesis gas composed primarily of CO, H_2, CO_2, CH_4, C_2H_4, and higher hydrocarbons. During the past year, research at Princeton has focused on experimental studies and mathematical modeling of the kinetic rates of the primary pyrolysis reactions and the secondary gas phase reactions.

The Dupont 990 Thermal Analysis System has been used to measure the primary pyrolysis rates of reaction of cellulose, newsprint from *The New York Times* and *The Wall Street Journal*, hardwood and soft wood sawdust, and cow manure at nominal heating rates of 5°, 10°, 20°, 50°, and 100°C/minute. Figure 4 gives a weight loss vs. temperature curve for cow manure for these five heating rates. Curves such as these can be simulated using the rate equation:

$$\frac{dV}{dt} = k(V^* - V)^n \qquad (1)$$

where k = A exp(-E/RT). In Eq. (1), the volatile weight fraction of the organic material is V*, the rate of weight loss (on a mass fraction basis) is dV/dt, K is the temperature dependent rate constant, T is the temperature, R the universal gas constant, A the frequency factor, E the activation energy, and n the reaction order. Figure 5 portrays a typical fit of the mathematical model given by Eq. (1) with the experimental results for a newsprint pyrolysis experiment (17).

Gas phase reactions of the volatile pyrolysis products with steam are being studied in a tubular quartz plug flow reactor designed and fabricated at Princeton. A schematic of the experimental layout is given in Figure 6. A small (0.1g) sample of organic material is rapidly heated (100 to 200°C/min.) and pyrolyzed in the pyrolysis reactor. Evolved volatiles are carried by the steam flow into the preheated, isothermal gas phase reactor, where they react with the steam to produce synthesis gas. Rates of production of 13 gaseous species are measured at 15 different times during the experiment using the 6 port and 34 port valves combined with a Hewlett Packard 5834a Gas Chromatograph.

Results of these experiments will be used to develop mathematical models describing the effects of feedstock composition, heating rate, final temperature, and residence time on the gaseous product composition. Although Princeton's modeling effort is not yet complete, data obtained from their reactor has already demonstrated the critical effect of temperature on the gas phase chemistry. The results of three experiments with cellulose using identical operating conditions *except* a variation in the temperature of the gas phase reactor are summarized as follows (17):

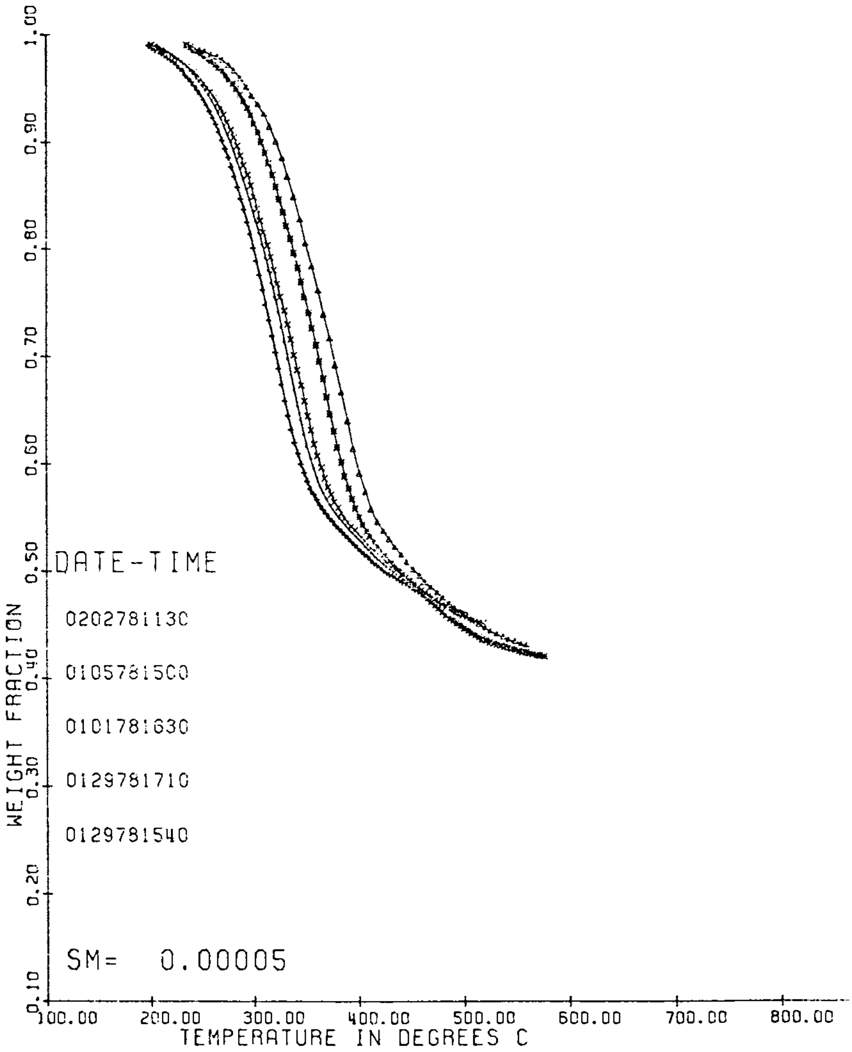

*Figure 4. Cow manure in nitrogen—weight loss vs. temperature: +, 5.360; ●, 10.960; ×, 22.930; *, 49.360; TRI, 138.400.*

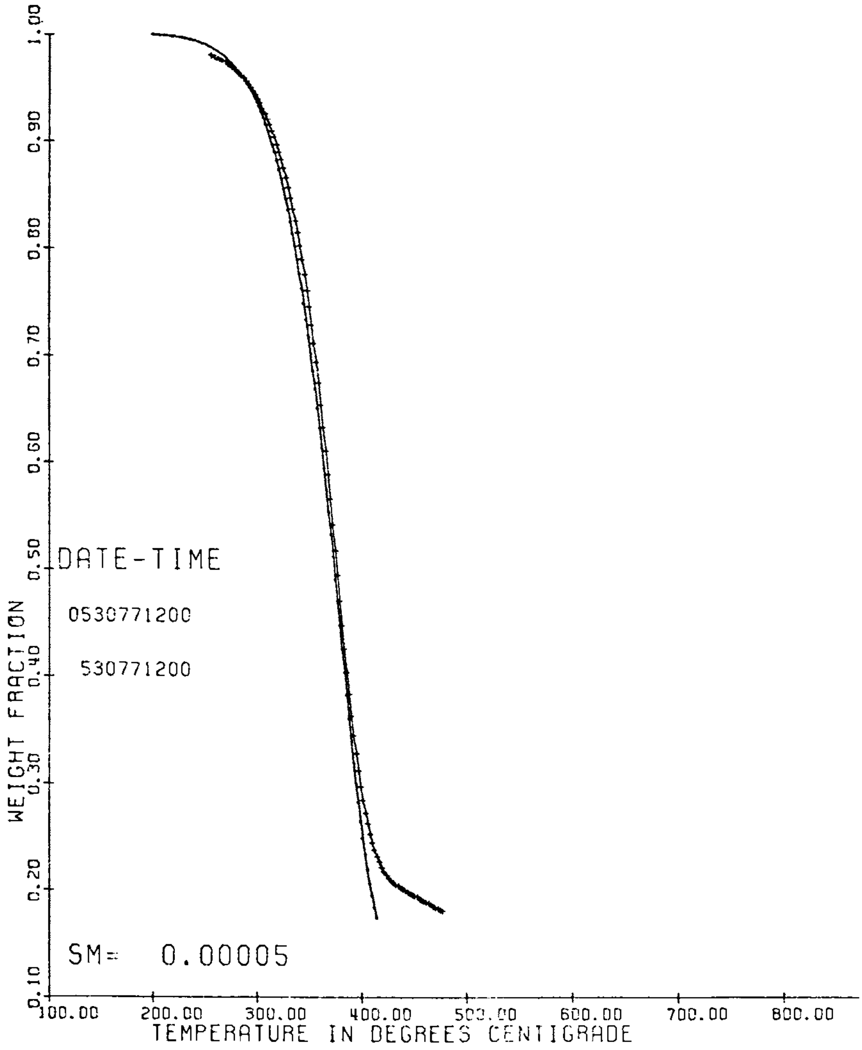

Figure 5. New York Times in nitrogen—weight loss vs. temperature (theoretical vs. experimental): +, 22.100; ●, 22.100.

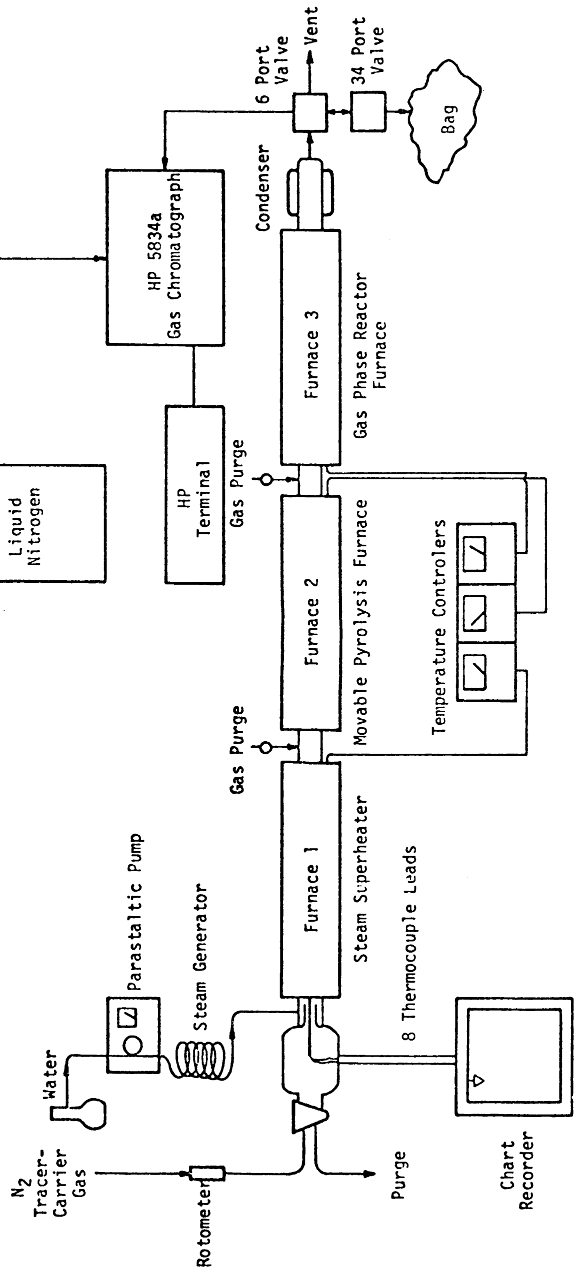

Figure 6. Schematic of the Princeton University tubular quartz reactor experiment

Gas Phase Reactor Temperature (°C)	Volume of Combustible Gas Produced (cc/g)	Calorific Value of Gas (10^6 Btu/Metric Ton)
500	94	1.0
600	318	6.0
700	824	15.0

In the experiment with the gas phase furnace at 700°C, less than the 2% of the sample weight was converted to condensible products (tars) and the gas had a heating value of about 500 Btu/scf. These results indicate that 90% of the weight of organic wastes can be converted into a medium-Btu gas with a heating value essentially equal to that of the feedstock. The remaining 10% of the waste weight is a char in the form of activated carbon.

Princeton is using their experimental results and mathematical models to design a chemical reactor for the gasification of organic wastes. Their design activities will be published at a future meeting of the International Solar Energy Society. Economic projections of Princeton's gasification reactor system were presented at the International Symposium on Alternative Energy Sources held in Barcelona, Spain (18).

Chemical Reclamation of Scrap Rubber (University of Tennessee)

For the past year, the Environmental Protection Agency has been supporting a project at the University of Tennessee to investigate and verify the technical feasibility of a concept to recover fuel products and carbon black from scrap rubber.

The proposed concept, shown schematically in Figure 7, utilizes molten salt as the heat sink to drive the pyrolysis reaction. Some major projected advantages of this concept are: (1) its ability to accept whole tires rather than shredded tires; and (2) the physical integrity of the recovered carbon black should make it desirable for re-use in the rubber industry.

The investigators at the University are now beginning to conduct experimental runs with the bench-scale unit (see Figure 8) (19). The apparatus consists of molten salt reservoirs, a flow-through reaction chamber, necessary piping and flow control, and necessary sampling ports with required recording equipment. The molten salt (zinc chloride) is utilized at 400°C. The capacity of the reservoir allows molten salt flow durations in the range of 10 to 30 minutes. This is well within the reaction times estimated to be necessary to complete the pyrolysis reaction in the batch-fed reactor. Start-up tests on the system are being conducted at this time, consequently, no data are available as yet.

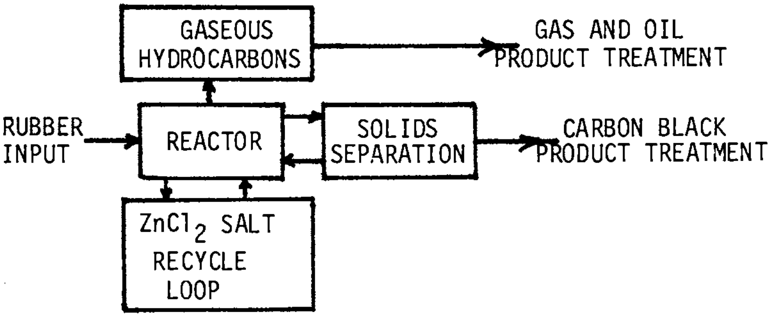

Figure 7. The University of Tennessee project—laboratory tests on a tire pyrolysis process that uses a molten salt catalytic technique

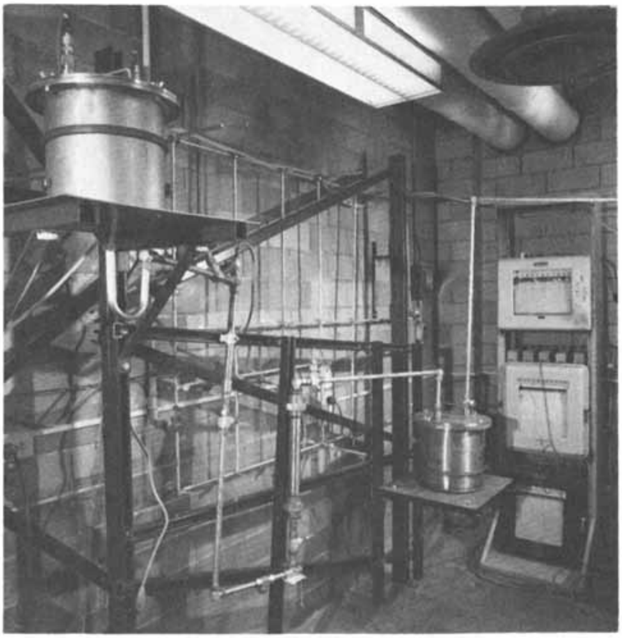

Figure 8. Bench-scale tire pyrolyzer at the University of Tennessee

The data to be obtained in these tests can hopefully be utilized in making decisions concerning the unit processes necessary for any larger-scale system (e.g., feedstock preparation, materials handling, materials separation, etc.). It is felt that the most favorable setting for such a plant would be next to existing ethylene plants. This would allow the sale of a tire-derived gas and liquid feedstock to the ethylene plant and this supplemental feed would have no significant impact on the plant capacity. Assuming 175 million tires are scrapped in the United States annually and that 85% of these could reach Molten Salt Process Plants, it is projected that the gross annual revenues for the products could range from $224.2 to $581.6 million (gases $2.3 to $37.6 million, liquids $57.9 to $380.0 million, and solids $164.0 million). No operating and capital cost estimates are projected at this time so that no estimate of net profit is available. Additional work on this project will include a total economic analysis (19).

Conversion of Solid Waste to Polymer Gasoline (The Navy at China Lake)

Over the past three years, EPA has been supporting a project conducted by the Naval Weapons Center, China Lake, California, to investigate and develop a system capable of converting solid wastes (organic materials) into synthetic petroleum products (primarily gasoline and oil). This effort has resulted in a bench-scale system capable of technically accomplishing the stated conversion. The system utilizes pyrolysis to gasify the solid waste to a gas that is rich in reactive, unsaturated hydrocarbons. These gases are compressed and purified. The purified hydrocarbons are then thermally polymerized to form the larger molecules of gasoline and oil. The crude polymer product has been shown to be 90% gasoline with an unleaded ASTM motor octane rating of 90. After simple distillation, this product can be stored, handled and utilized like any petroleum-derived gasoline. The process is projected to produce 0.21 liter of synthetic petroleum product per Kg of solid waste (minimum of 70% organic content), which translates to about 50 gallons of gasoline per ton of solid waste (20).

Experiments to date have been conducted in various pyrolysis reactor configurations. Figure 9 depicts one of the most recent configurations utilized to gasify the solid waste feedstocks. An initial analysis of several feedstocks was made to determine initial composition (see Table XIII). For ease of handling and feeding, ECO II "fuel" has been utilized as the principal feedstock. Utilization of ECO II fuel in pyrolysis tests resulted in data of the type illustrated in Table XIV. Many runs of this type have been done and have resulted in the optimization of a pyrolysis system that produces the maximum quantities of gasoline

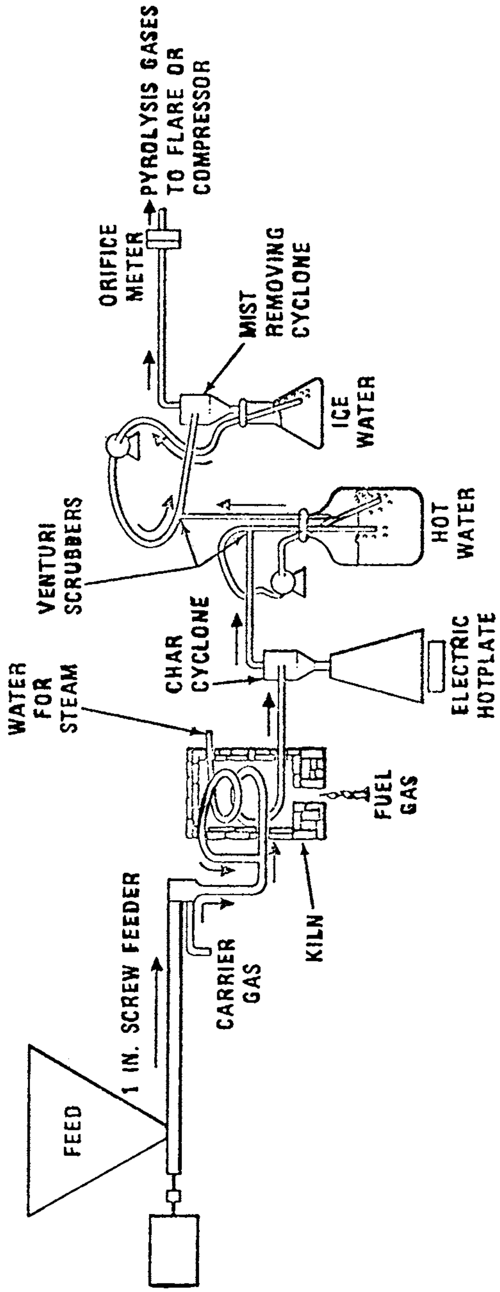

Figure 9. Pyrolysis schematic—China Lake project

TABLE XIII. Pyrolysis Feed Analysis (20)

Sample	Weight % moisture	Weight %, dry basis							Particle size, average	Dry basis heat of combustion HHV, cal/g (Btu/lb)
		Ash	C	H	O	N	S	Cl		
ECO II (composite NWC)	3.82	11.87	...	...	...	...	0.73	0.70	...	4,754 (8,558)
ECO II (drum T-47)	1.69	12.46	45.50	6.20	34.22	0.27	0.98	0.37	188µm	4,726 (8,507)
ECO II (drum T-51)	0.91	9.73	49.64	6.40	32.38	0.22	1.04	0.59	270µm	5,306 (9,550)
Altoona, PA	43.3	12.0	47.6	6.0	32.9	1.2	0.3	...	...	4,748 (8,546)
Dry cellulose +12% ash	0	12.0	39.11	5.4	43.45	...	...	...	...	3,677 (6,618)
IBM computer cards	3.8	0.68	43.21	6.46	49.45	0.10	0.023	0.080	...	3,920 (7,056)

TABLE XIV. Pyrolysis Summary (20)

Run	Date	% Moisture	Pyrolysis Temp., °C	Carrier Gas				Grams net product 100 grams dry feed										Volume % in gross gases	
				Mole %			M^3/kg dry feed												
				H_2	CH_4	CO		C_2H_2	C_2H_4	C_2H_6	C_3H_6	t-C_4H_8	CO_2	H_2	CH_4	Char	C_2+	C_2H_4	C_2+
1	11/5/75	0.01+	760	...	100	...	0.55	0.1	8.2	1.0	1.8	0.5	6.7	0.3	2.3	24.0	12.0	7.5	9.9
2	11/12/75	0.15	790	*a*	*a*	*a*	0.89	0.1	11.6	1.0	1.4	0.4	10.8	0.2	6.7	23.7	15.0	6.3	7.5
3	11/13/75	0.10	815	...	...	100	0.39	0.2	7.4	0.6	0.7	0.3	8.6	0.6	6.6	15.7	9.3	7.0	8.4
4[b]	11/18/75	0.10	850	...	...	100	0.25	0.2	4.3	0.2	0.2	0.1	4.1	0.4	4.1	15.0	5.1	6.5	7.5
5	12/5/75	0.10	815	16	44	40	0.30	0.2	7.8	0.6	0.5	0.1	8.8	0.2	8.8	18.6	9.2	8.2	9.3

[a] This carrier gas was a poorly mixed 20% H_2, 40% CO, 40% CH_4 mixture.

[b] Run No. 4 is suspected to have had a leak of pyrolysis gases since the gaseous product weights are all very low.

NOTE: These runs were made with ECO II (drum T-51) as feed. After run No. 1, there were two turns of 3/4-inch tubing (110 inches) prior to the feed entering the cyclone separator. The loss of carrier gas is thought to have occurred in all runs.

precursors (most notably, ethylene). It has been determined that in order to maximize hydrocarbon production during pyrolysis it is necessary to: (1) have a very rapid heating rate to temperatures in the range of 700°C; (2) a pyrolysis environment which keeps the free radical partial pressure low to prevent these radicals from reacting to form tars; and (3) a short residence time to prevent the formation of secondary reaction products. Achieving these conditions results in a gas stream with the appropriate hydrocarbon content necessary for gasoline polymerization.

After being generated in the pyrolysis operation, the gases are compressed, stored, and then purified. The purification system consists of a water scrubber which removes carbon dioxide along with water-soluble organics (acetic acid, etc.) and a hydrocarbon absorber and stripper system which utilizes recirculating benzene. The purification system takes the off-gases from the pyrolysis system and provides the "pure" gases coming from the benzene stripper to the gasoline polymerization unit.

The polymerization unit is designed to operate at temperature ranges of 750° to 950°F and at pressures of 700 to 1000 psi. These ranges are currently employed commercially in this type of process in the petroleum industry. Table XV shows experimental results from early tests utilizing pure ethylene as a feedstock (20).

An integrated system (i.e., pyrolyzer off-gas feeding the purification/polymerization section) has been built and is currently being debugged. It is planned that additional experimental data will be available in the near future.

A preliminary economic analysis of this system was completed by the project staff at China Lake. Municipal financing was assumed with a 25 year, 6.5 percent interest, amortization cost. The solid waste was assumed to contain 60.2 percent organic material, 27 percent moisture, 5.5 percent iron, 0.5 percent aluminum, 6.6 percent glass and 0.2 percent other materials. It was assumed that 46 gallons of liquid product would be produced per ton of solid waste and that the per capita generation of solid waste was 5 pounds per day. Figure 10 indicates the net operating costs per ton of solid waste as a function of population with negative costs indicating a profit (20). A more complete economic analysis is currently being done by an independent, "third-party" contractor of the EPA, Dow Chemical. Dow's analysis will be based upon their experience with the unit processes necessary to accomplish the required conversions and actual experimental data being generated on the latest integrated, bench-scale unit at China Lake.

TABLE XV. Polymerization Test Data Summary (20)

Run No.	Date	Coil volume, cm^3	Bath temperature, °F	Coil pressure, psia	Run time, min	Ethylene input, g/min	Gas out, std cc/min	Residence time, min	Gasoline production, cc/hr	% conv. to gasoline	% conv. (Theoret.)	Remarks
3	9/18/75	80	850	1,015	135	?	0	?	14.2	?		
4	9/23/75	80	840	1,065	120	?	0	?	41.0	?		Plugged
5	9/25/75	80	875	765	120	0.30	0	8.0	25.8	?		
6	9/26/75	80	910	515	270	0.02	0	75.6	1.8	?		
7	10/24/75	80	725	600	90	...	600/1,200	?	0	0		
8	10/30/75	260	890	815	90	1.16	600	4.5	38.5	42.0	46.4	
10	11/4/75	260	800	515	45	0.80	600	5.7	8.0	11.7	15.8	
11	11/5/75	260	810	765	90	1.22	600	5.3	43.3	42.7	23.3	
12	11/5/75	80	775	785	65	0.89	600	3.0	15.7	20.2	7.5	
13	11/6/75	260	765	785	120	0.38	0	17.7	32.5	?		
14	11/14/75	260	975	1,035	45	0.91	600	8.4	17.3	21.1	85.5	Plugged, leak
15	11/18/75	260	850	1,045	45	?	600/300	?	72.0	?		
16	11/18/75	260	835	970	45	0.83	150	9.4	56.0	79.0	51.8	
17	11/18/75	260	840	895	30	1.04	500	7.7	38.0	44.0	44.6	
18	12/5/75	260	875	1,015	45	?	150/600	?	89.0	?		
19	12/8/75	260	820	1,015	90	0.91	460	9.5	31.3	39.7	43.8	Leak

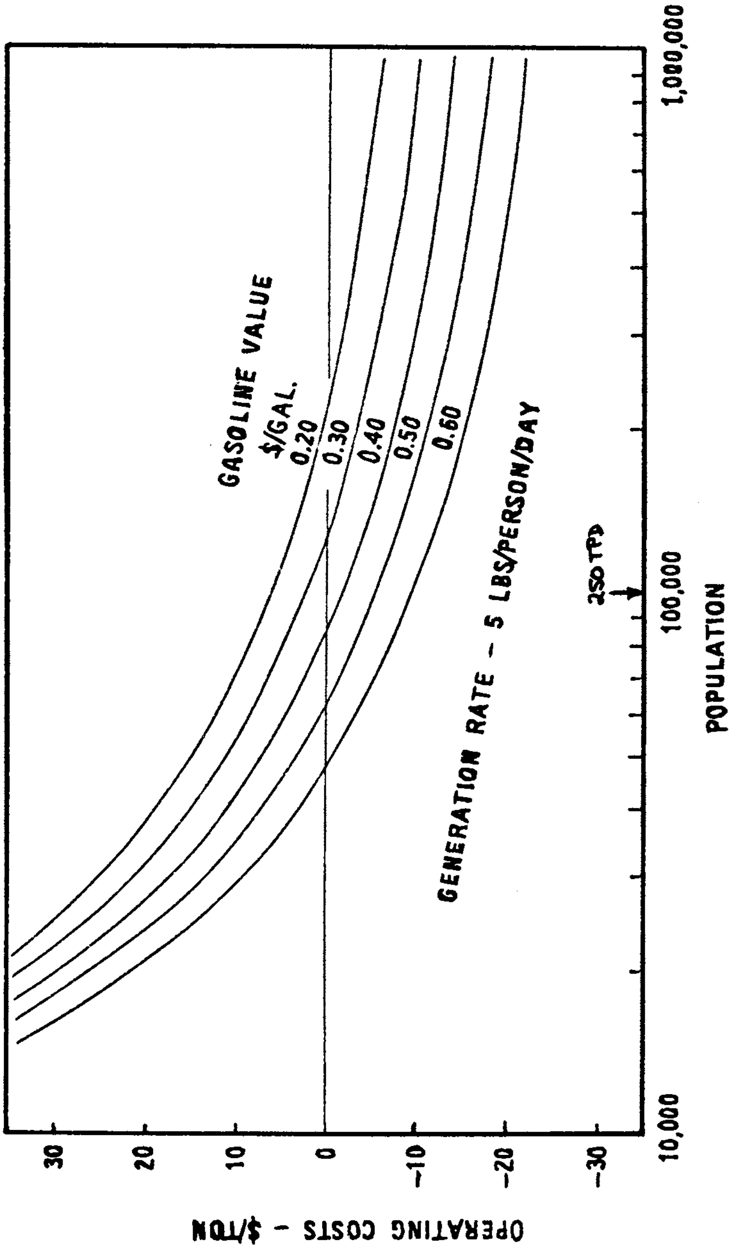

Figure 10. Net operating costs with materials recovery and conversion of organic fraction to gasoline—China Lake project

Conclusion

The projects discussed in this paper are examples of EPA's current efforts in the area of basic and applied research, and of their investigation of new and innovative concepts, as they relate to the thermochemical or pyrolytic conversion of solid waste into usable fuels. These efforts have the prime objective of providing process developers and potential technology users with information necessary to make sound technical assessments of these concepts. Other more advanced waste pyrolysis concepts supported by EPA that are currently underway include the development of a prototype mobile pyrolysis system for agricultural/industrial wastes (co-funded with State of California Solid Waste Management Board) and demonstration projects such as the Occidental Process at El Cajon, California, and Monsanto-Landgard Process at Baltimore, Maryland. It is intended that the data generated by our more basic or applied research projects (discussed herein) will lead to new commercial processes. The results of this EPA research will provide solid waste disposers with new, environmentally, technically, economically feasible options for converting their wastes into beneficial fuel products.

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The Syngas Process

HERMAN F. FELDMANN

Battelle–Columbus Laboratories, 505 King Ave., Columbus, OH 43201

JOSEPH K. ADLERSTEIN

Syngas, Inc., Suite 1260 East, First National Center, Oklahoma City, OK 73102

The Syngas Process concept and experimental data used in its development have been published.(1,2) The intent of this paper is to present additional information on calculated material balances for an integrated process, projected oxygen consumption, as well as to present preliminary projections of the economics of producing an intermediate-Btu gas by the Syngas Process.

Briefly, the Syngas Process, Figure 1, consists of a two-stage reactor system with a zone between the two stages where separation of metal and glass from organic char is accomplished by elutriating the relatively light carbonaceous char from the dense metal and glass phase. In the first stage, incoming solid waste contacts a hydrogen-rich synthesis gas generated from the gasification of residual carbonaceous char. In this first stage, the incoming waste is devolatilized and hydrogasified and the residual char is then gasified to produce the hot synthesis gas. Our experimental studies suggest that the product distribution (namely CH_4 versus organic liquids) is influenced by the hydrogen/solid waste feed ratio more than by the hydrogen partial pressure and, therefore, there is considerable flexibility in the selection of the system pressure.

Overall Material Balances

Overall material balances for the Syngas Process utilize experimental data generated in our small process development unit (PDU) to determine the key factors influencing carbon conversion and product distribution in the first stage methane production reactor (MPR). The char produced in the MPR is then fed to a second stage entrained gasifier where it is completely gasified with oxygen and steam. Because the char fed to the second stage is mostly devolatilized, the performance of the entrained gasifiers was simulated using published data for entrained systems feeding coal or coal char.(3)

Figure 2 shows a material balance calculated on the above basis. By separation of the MPR from the gasification zone, the

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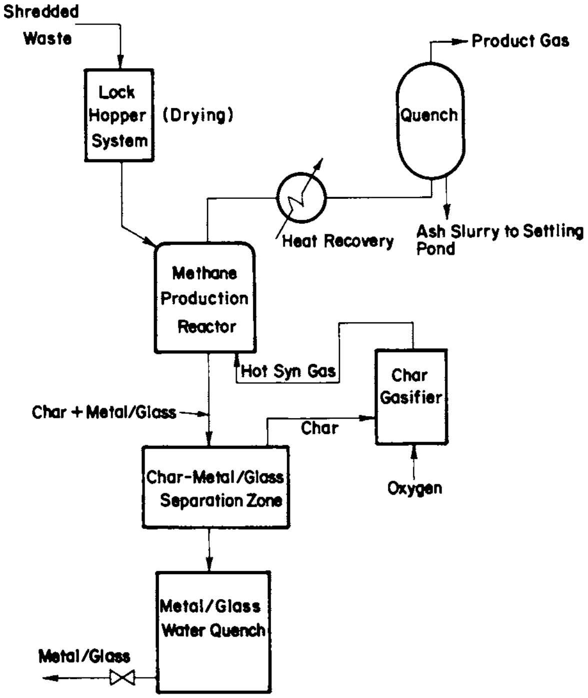

Figure 1. Simplified block diagram—Syngas reactor system

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

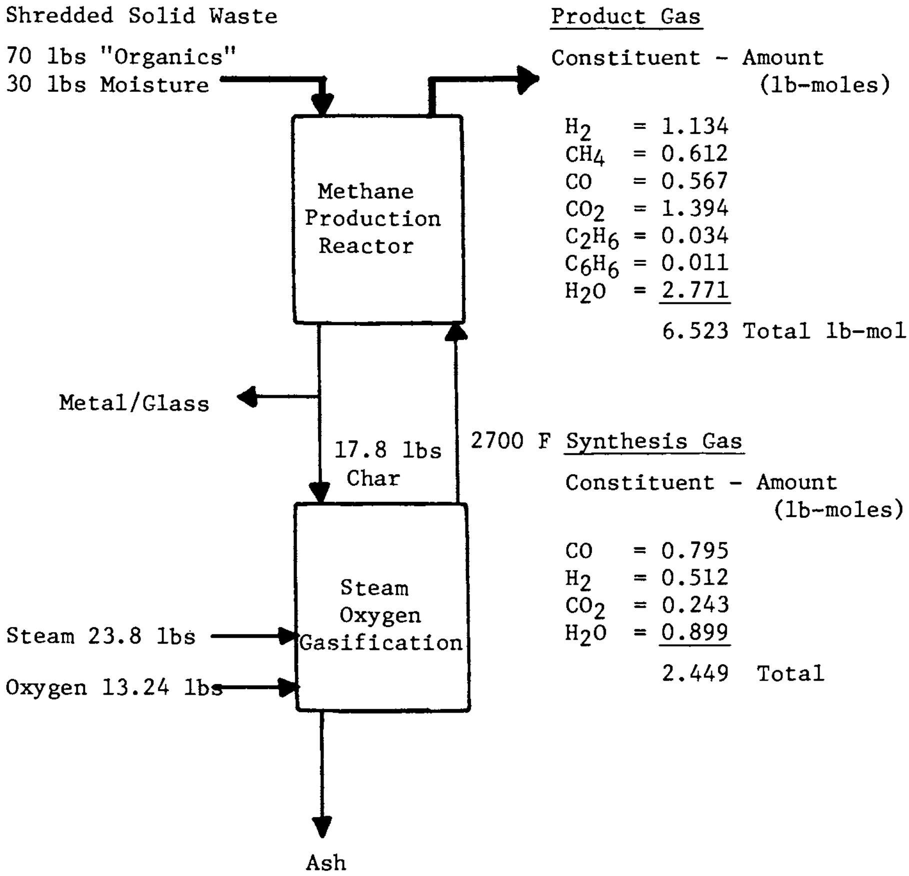

Figure 2. Material balance for the Syngas process

following key advantages critical to the overall process economics are achieved.

(1) Metal and glass can be recovered in a resalable state without being slagged or without requiring a complex front end separation and cleaning system. If the metal and glass entered the gasifier, fusion and possible oxidation of metal would result.

(2) Methane yields are increased because the methane is prevented from entering the gasifier where it would be reformed or combusted.

(3) Oxygen requirements are minimized because only the char from the MPR, a small fraction of the original solid feed, is gasified and because the char enters the gasifier at a temperature of 1000 to 1200 F. In addition, the overall process heat requirements are minimized because of the high methane yield possible from this type of staged system.

As mentioned, the balances shown in Figure 2 are based upon experimental data generated in the PDU studying the MPR.(1) For the steam/oxygen gasifier, experimental data presented by Von Fredersdorff and Elliott(3) for an IGT entrained gasifier was used. A comparison of the IGT entrained gasifier data with that from other entrained gasifiers indicated excellent consistency and it is therefore felt that the results are typical. Because the char from the MPR enters the gasifier at 1000 to 1200 F, oxygen consumption should be less than that calculated based on the above gasifier tests because the coal used in these tests was not preheated. It should be stressed that experimental data for an entrained gasifier was used and the gasifier was small compared to commercial size. Thus, these results include the effects of heat losses and conservatively represent the performance of a larger system.

After cooling and quenching, the product gas from the Syngas reactor will have the approximate composition and the yield shown below.

Fuel Gas Yield and Composition from Syngas Reactor System
Fuel Gas Composition, volume percent
H_2 = 30.3
CH_4 = 16.4 Heating Value = 329 Btu/SCF
CO = 15.1 Yield, SCF/lb-dry "organics" = 20.3
CO_2 = 37.3
C_2H_6 = $\underline{\ 0.9}$
100.0

The major uncertainty in the above gas composition is the distribution between H_2, CO and CO_2 which is determined by the water gas shift. The composition above is based on assuming an H_2/CO ratio approximately 2 because of the high steam content of

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

the gas. However, the Purox process reports an H_2/CO ratio slightly less than 1 in spite of high steam concentrations in the gas. A lower value for the H_2/CO ratio will increase the heating value of the dried product gas because the less water-gas shift occurring the lower the CO_2 content of the gas and the higher the H_2O content which is removed by drying while CO_2 is not.

Because oxygen consumption is important in determining gas production costs, the oxygen consumption projected for the Syngas Process should be compared with those reported for other gasification processes. A comparison between projected oxygen consumption for the Syngas Process, the Purox Process, and the Lurgi Process (for coal gasification) is given in Table 1.

TABLE 1. OXYGEN REQUIREMENTS PER MILLION BTU OF RAW PRODUCT GAS

Process	Oxygen Requirements (tons oxygen/MM Btu, raw product gas)
Syngas	0.014
Purox(1,2)	0.029
Lurgi(3)	0.019 (For a high-oxygen western coal)

(1) Solid Waste Disposal Resource Recovery, Brochure Published by Union Carbide's Environmental Systems.

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Thus, the Syngas Process offers the potential of a very substantial reduction in oxygen requirements over the Purox Process, which employs a fixed-bed slagging gasifier.

Reactor Design Concepts. Experimentally, we operated the MPR as both a free-fall and moving-bed reactor. However, these tests established that the conversion rate was limited only by heat transfer to the particle and that the conversion level was determined by the H_2 "seen" by the waste. Thus, there is considerable flexibility in reactor design.

For the purposes of estimating reactor throughput, a free-fall system is assumed. One major advantage of a free-fall MPR is that there should be no problems with bridging of solid waste. (For example, we were able to operate even our 2.8 inch I.D. MPR in the free-fall mode without bridging problems.) The presence of metal and glass should provide additional impetus for maintaining flow through the free-fall zone. Another advantage of the free-fall system is the increased utilization of sensible heat in the product gases, which allow feeding higher moisture content waste.

Crude measurements of the free-fall velocity of shredded paper, which will have the lowest terminal velocity of any significant fraction of solid waste, indicate that its free-fall velocity is about 4 ft/sec.

Operating this system in the free-fall mode at a gas velocity limited to 3 ft/sec allows a gas production rate in the MPR equivalent to about 7 million Btu/hr-ft^2 (1100 lbs dry organic material/hr-ft^2).

This specific Btu production rate is quite high compared to most coal gasification processes especially fixed or moving-bed systems. For example, the specific Btu output of a Wellman-Galusha coal gasifier has been reported to be about 1.5 million Btu/ft^2-hr.(4)

After the char/metal/glass mixture falls through the free-fall zone, the char will be stripped from the metal/glass mixture by a stream of steam and the char will be blown into an entrained gasifier into which oxygen is injected to complete gasification of the char.

Before a detailed design of the system is possible, additional data will have to be generated. For example, it will be necessary to make more accurate measurements of entrainment velocities of various solid waste constituents after primary shredding. In addition, the drying rates of solid waste will have to be determined in order to evaluate whether predrying is necessary of if flash drying can be carried out in the top of the MPR itself.

Integration of the MPR and gasifier should provide no technical problem but will require more detailed engineering analysis.

Preliminary Cost Estimates. Because we do not yet have a detailed reactor design, it is impossible to do a detailed cost estimate. By utilizing published costs for similar coal gasification processes, it is possible to determine a reasonable estimate of what gas costs are likely to be. Unfortunately, most of the published cost data on coal gasification is for base load pipeline gas plants which are far larger than is practical for a plant utilizing municipal solid wastes. However, the Culbertson, Kasper discussion on the economics of small coal gasifiers(4) provides a means of examining cost feasibility for plant sizes of greatest interest in gasifying municipal solid waste as well as other biomass feeds.

Economic feasibility was evaluated for a cattle manure feedstock in addition to the solid waste. Chemically, both are similar except that cattle manure is probably more reactive, more homogeneous, and, in many areas, available at a price that makes its utilization for conversion to gas economically attractive.

The cost projections are the utilization of commercially available Wellman-Galusha^{R} agitator type gasifiers 10 feet in diameter.

Though the Syngas system is different than the Wellman-Galusha^{R} gasifier, the higher throughput possible with the Syngas system makes the utilization of Wellman-Galusha^{R} cost data reasonable and probably conservative.

Included in the total capital requirements are:

(1) Estimated installed cost of both on-site and off-site facilities,
(2) Project contingency of 15 percent of facility cost,
(3) Initial charge of catalysis and chemicals,
(4) Paid-up royalties,
(5) Allowance for funds used during construction,
(6) Start-up costs, and
(7) Working capital.

The estimated gas price is the average price based on the following assumptions outlined in the ERDA Gas Cost Guidelines.

(1) 20-year project duration.
(2) 20-year straight-line depreciation on plant investment, allowance for funds used during construction, and capitalized portion of start-up costs.
(3) Debt/equity ratio of 75/25.
(4) 15 percent return on equity.
(5) 9 percent interest on debt.
(6) Federal income tax rate of 48 percent.

The major sections of the plant are:

(1) Coal storage and handling,
(2) Coal gasification,
(3) Particulate removal,
(4) Tar removal and disposal,
(5) Water treatment and disposal,
(6) Ash disposal, and
(7) Oxygen plant (for the intermediate-Btu gas case).

The only modifications of the above cost projections that were made were the following.

(1) The external oxygen consumption for producing intermediate-Btu gas from manure was adjusted for the difference in oxygen consumption between solid waste and coal.
(2) The difference in heating value between manure and solid waste and coal was compensated for in determining the manure requirements to produce an equivalent Btu output.
(3) The coal cost contribution was isolated to allow separate assessment of the effect of fuel cost.

Manure Case. No advantage was taken of the much higher reactivity of solid waste and manure compared to coal, the fact that these feeds do not agglomerate, the higher ethane yields resulting from their pyrolysis and the higher throughputs anticipated for the Syngas Reactor System. Thus, these cost projections summarized below should be regarded as conservative.

Comparison of Average Gas Prices Produced from Manure and Coal from a Single 10ft I.D. Fixed-Bed Gasifier

	Coal	Manure
Total Capital Requirements, millions	5.44	5.44
Manure or Coal Requirement, tons/day	132	254
Gas Production, 10^9 Btu/day	2.27	2.27
Cost Correction for Reduced Manure Oxygen Requirements, \$/MM Btu		-0.32
Average Gas Price without Coal or Manure Cost, \$/MM Btu	2.49	2.17
Average Gas Price Coal = \$35/ton, Manure = \$2.50/ton, \$/MM Btu	4.53	2.45

These cost figures establish that an intermediate-Btu gas can be produced from manure at a cost that is currently competitive with imported LNG (priced at 2.50 to 3.50/MM Btu) and No. 2 heating oil (priced at \$2.70/MM Btu as of Summer 1977) and at a considerably lower price than it can be produced from coal.

Wherever manure can be combined with solid wastes, the costs will be even more attractive because of the lower cost to the plant of solid waste compared to manure and the larger plant size. Of course, other biomass can also be combined to allow larger plants. The quantitative effect of combining other forms of biomass will depend on its cost relative to manure.

Solid-Waste Case. Conversion of municipal solid waste to either an intermediate or low-Btu gas will be extremely cost effective because a nominal land-fill fee can be charged to dispose of the solid waste and the contained metal and glass can be recovered in resalable form by the Syngas Process.

In the production of intermediate-Btu gas, another important factor in costs is oxygen consumption. As previously pointed out, the Syngas Process should allow a reduction of approximately \$0.32/MM Btu compared to coal (because of the higher oxygen content of cellulosic type materials compared to coal). As mentioned, a gravitating, fixed-bed, slagging bottom gasifier such as the Purox will require substantially more oxygen according to published oxygen consumption data.(5)

In addition to the increased oxygen consumption, the fusion of metal and glass to slag reduces the revenue that could otherwise be obtained from their sale. For the slagging gasifier case, it is assumed that their is no net value for the slag and for the Syngas Process a net value of $3.00 MM/Btu is assumed because metal and glass are recovered in resalable form.

In order to maintain these cost projections on as common a basis as possible, the same investment costs were assumed for the slagging gasifier and the Syngas Process with an adjustment made in oxygen consumption, which for the coal case is estimated by Culbertson and Kasper(4) to contribute about $1.00/MM Btu of product gas.

The overall thermal efficiency of both the Syngas Process and the slagging gasifier is assumed the same (approximately 72 percent) as estimated by Culbertson and Kasper for the coal gasification case.

No low-Btu gas cost projections are made for the slagging gasifier because it is impossible to achieve slagging temperatures with an air-blown gasifier unless excessively high air preheat temperatures are employed. Achieving these temperatures requires auxiliary firing of the air preheater with a clean fuel.

Also, for the solid-waste cases, it was arbitrarily decided to increase the investment cost by 20 percent over the coal case to account for the greater hetrogeneity of solid waste compared to coal or manure. The gas prices presented in the following table are:

(1) The average gas price with coal as a feedstock.

(2) A base case gas price for both the Syngas and Slagging Gasifier Systems which assumes that waste, already shredded at various transfer stations, is delivered to the gasification plants. The disposal charge for this waste is assumed to be $2.00/ton. For both Syngas and the slagging gasifier a 20 percent increase in plant investment was assumed for the same Btu output.

(3) A gas price assuming a net recycle value of $3.00/ton for recovered metal and glass is calculated for the Syngas Process. The justification is that the Syngas Process recovers the metal and glass in a resalable form as part of the gasification process.

(4) Prices for low-Btu gas are presented only for the Syngas Process because it can operate air blown as well as oxygen blown.

These cost projections indicate:

(1) That either solid wastes or manure can produce intermediate and low-Btu gas at prices that are either currently competitive or substantially cheaper than other clean fuels and, in the case of a solid-waste feedstock, even cheaper than coal (coal at $35/ton is approximately $1.46/MM Btu).

Table II. **COST PROJECTIONS FOR INTERMEDIATE AND LOW-BTU GAS FROM SOLID WASTES**

No. of Gasifiers	Coal Req. tpd	Gas Output 10^9 Btu/day	Gas Price Coal Feed $/MM Btu	Equivalent Solid Waste Requirements tpd	Syngas Base Case Gas Price @ $2.00/ton Solid Waste fee $/MM Btu	Syngas Gas Price $3/ton Metal and Glass Value $/MM Btu	Slagging Gasifier $2/ton Solid Waste Fee No additional Value for Metal and Glass
Oxygen Blown Coal Cost Contribution = $2.03/MM Btu Gas Heating Value = 285 Btu/SCF for coal; about 400 Btu/SCF from solid waste							
1	132	2.27	4.53	284	2.25	1.30	2.77
5	660	11.35	3.75	1420	1.47	0.52	1.99
Air Blown Coal Cost Contribution = $1.66/MM Btu Gas Heating Value = 158 Btu/SCF							
1	78	1.64	3.44	168	1.68	1.37	--
5	390	8.20	2.62	839	0.86	0.55	--

(2) That the oxygen reduction made possible by the Syngas Process allows a very significant reduction in gas price over a slagging gasifier.
(3) The additional net revenue because of the separation of metal and glass that accrues to the Syngas Process is an extremely important factor in process economics.
(4) That a plant producing low or intermediate-Btu gas from manure or solid waste need not be a giant to be economic. In fact, if low-Btu gas is produced, plant sizes capable of serving small communities (170 tpd) are economically very attractive. This greatly increases the solid waste and manure that is economically available for gasification.
(5) Because one can produce a low-Btu gas from solid waste at a lower price per Btu than is available for coal, at an overall thermal efficiency of 88 percent, one should seriously examine the option of selling this gas as supplementary fuel to a larger fossil fuel power plant rather than co-firing waste and refuse.

Obviously the combination of attractive economics, an ability to utilize a currently wasted resource, and the solution of a growing envionmental problem make the Syngas Process attractive for commercialization. These economic projections also indicate that gasification of waste and transportation of the gas to end users who can utilize the gas is probably a much more attractive option than using the solid waste directly. For example, the option of gasifying the waste and using the intermediate-Btu gas in utility boilers will probably be a more attractive option than firing waste directly. For larger plants (~3000 tpd), the production of SNG from the raw product gas may also present an attractive option in many locales.

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March 3, 1978

Investigations of the PERC Process for Biomass Liquefaction at the Department of Energy, Albany, Oregon Experimental Facility

T. E. LINDEMUTH

Research and Engineering, Bechtel National, Inc., 50 Beale Street,
San Francisco, CA 94119

Background of the Project

Earliest efforts in the liquefaction of biomass as it is currently practiced at the Albany experimental facility took place during the late 1960's and early 1970's at the U.S. Bureau of Mines Pittsburgh Energy Research Center (PERC). The study there showed that a wide variety of biomass materials, including wood municipal solid waste and cattle manure, could be turned into an oil-like material by reaction with carbon monoxide in the presence of an alkaline catalyst under conditions of moderate temperature and high pressure.[1-3] This work was an outgrowth from earlier experiments of coal liquefaction using a similar reaction scheme. On the basis of the PERC results, the flow scheme for both a commercial size plant and what was to eventually become the Albany facility was developed in 1973.[4] Figure 1 shows the process portion of the Albany experimental facility. Detailed engineering specifications were prepared in 1974[5] and construction of the facility proceeded soon thereafter. In late 1976 Bechtel National, Inc. was contracted to monitor the completion of construction and then to commission the facility. The following paper describes some of the activities involved with commissioning of the facility as well as the initial results and preliminary economic assessments of the process.

Objectives

The major objectives in the first year's activities at Albany were to verify, as well as possible, the original process data, to evaluate the process equipment involved, and to develop a preliminary conceptual design and economic assessment.

The plant itself is located within the city limits of Albany, Oregon immediately adjacent to the U.S. Bureau of Mines Metallurgical Research Center. This facility is set in the middle of a residential area instead of the more isolated setting that one might expect. The location has created a special concern for new adverse environmental effects. In addition to the PDU shown in Figure 1, the site has a modern control room, an office building for technical and administrative staff, a repair and maintenance shop, a small process

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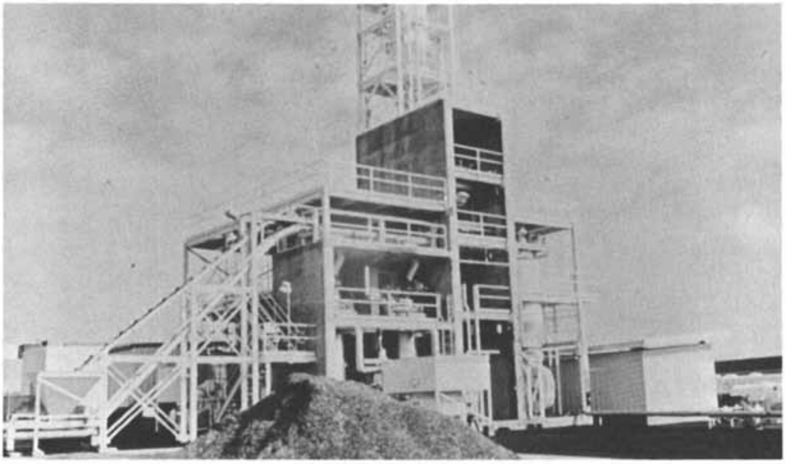

Figure 1. Albany, Oregon biomass liquefaction PDU

control laboratory, and simple pollution control facilities.

Figure 2 is a simplified flow scheme for the Albany facility as it was originally built. In the original process scheme, wood in the form of wood chips, such as are used in the manufacture of pulp and paper, are brought to the plant and conveyed into a storage hopper. First, moist wood chips are dried in a rotary drum drier. Here, the moisture content is reduced from 50 percent down to 3 percent or 4 percent. After drying, the wood is ground in a hammer mill to approximately 50 mesh. This gives the material the consistency of ordinary flour. In order to bring the wood into the reaction system, it is first blended with plant recycled oil in a multibladed continuous blender, similar to those used in making bread dough or compounding rubber. The wood/oil slurry, containing approximately 25 percent wood flour is then pumped up to the reaction system pressure which is in the range of 100 to 250 atmospheres. The slurry is then routed to a scraped surface heat exchanger where it is heated to approximately 350°C. At that point a catalyst solution of aqueous sodium carbonate is added and the mixture is routed to a 375 L. stirred autoclave. In the autoclave the mixture is further heated to 370°C and contacted with carbon monoxide gas. This gas is provided from storage tanks and delivered by multistage diaphragm compressors. Both the preheater and stirred tank reactor are electrically heated. Offgases from the reactor containing carbon monoxide, carbon dioxide, and hydrogen are cooled, condensed, and combusted in a flare.

The liquid mixture that remains from the CSTR is first cooled and then depressurized for removal of water and noncondensible gases. These gases are also routed to the flare system. Liquid from the flash tank is sent to a centrifuge for the removal of water and solid sludge. From the centrifuge, part of the oil is recycled to the wood/oil blender, thus completing the process loop. The remaining oil is removed as product.

Plant Commissioning

Before any process testing could begin, the entire facility had to be checked out and each individual piece of equipment commissioned. To give you an idea of the overall complexity of the plant, Table 1 lists the number of major process items in the plant, each of which required some sort of check-out. To further illustrate the complexity of the facility, Figure 3 shows the central control panel.

During the commissioning of the plant a number of difficulties were encountered with the process equipment itself. This is not unusual for the first development unit in a new process. Table 2 highlights the areas where most difficulties were encountered. Of particular importance were pump and agitator shaft seals on high pressure equipment. In order to use this equipment, many modifications had to be made in this area. In addition, problems showed up with the gas compressors, wood handling equipment, and the preheater scraper. To illustrate the magnitude of some of the problems, Figure 4

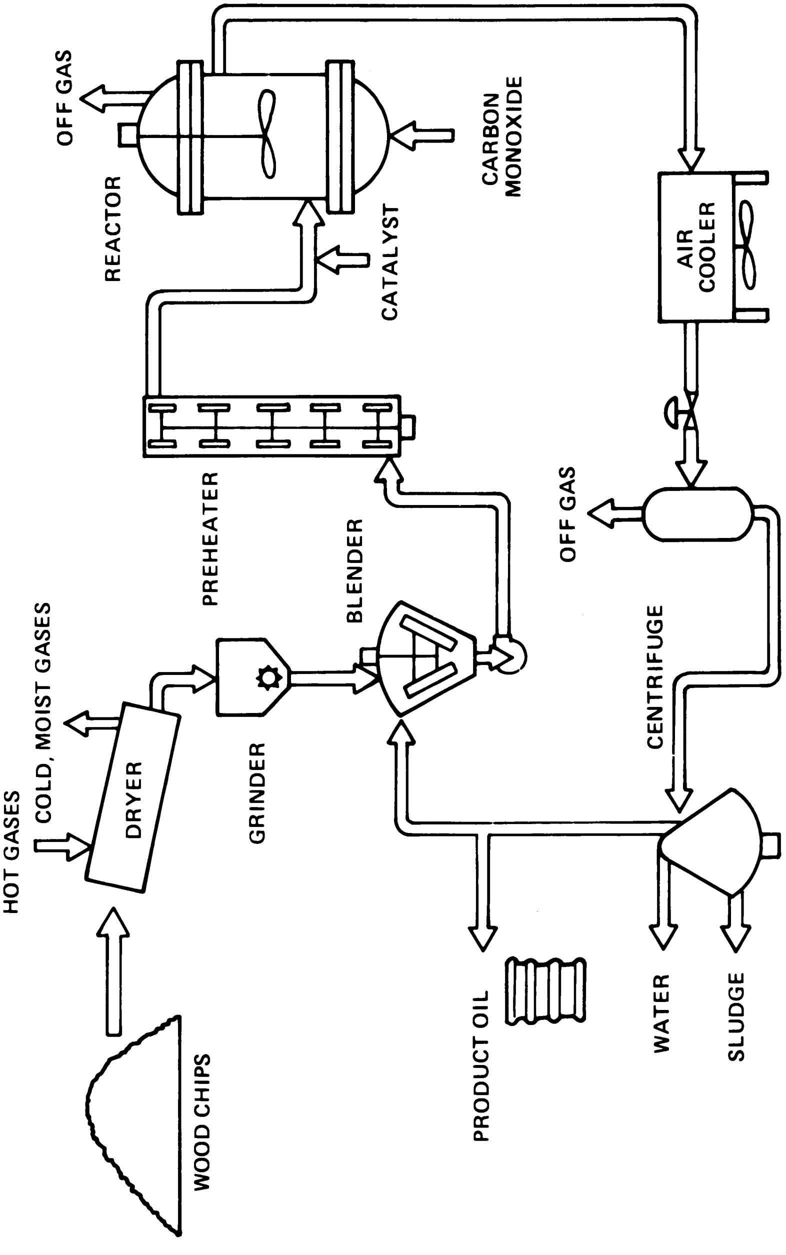

Figure 2. Albany PDU schematic (as built)

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

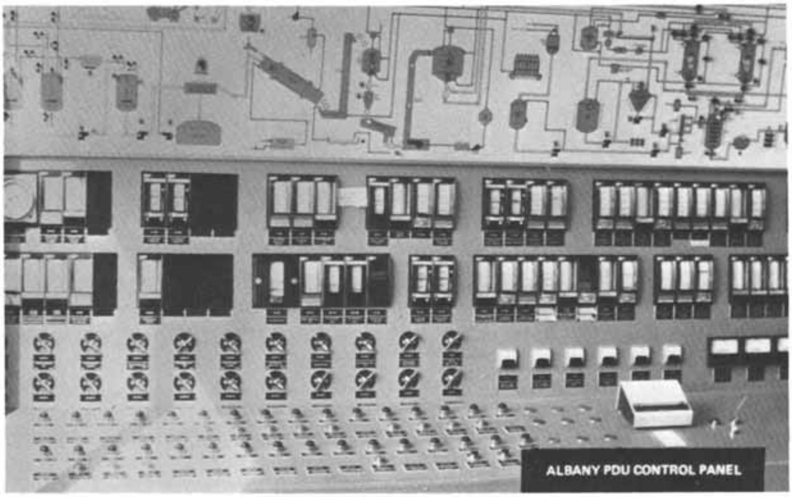

Figure 3. Albany PDU control panel

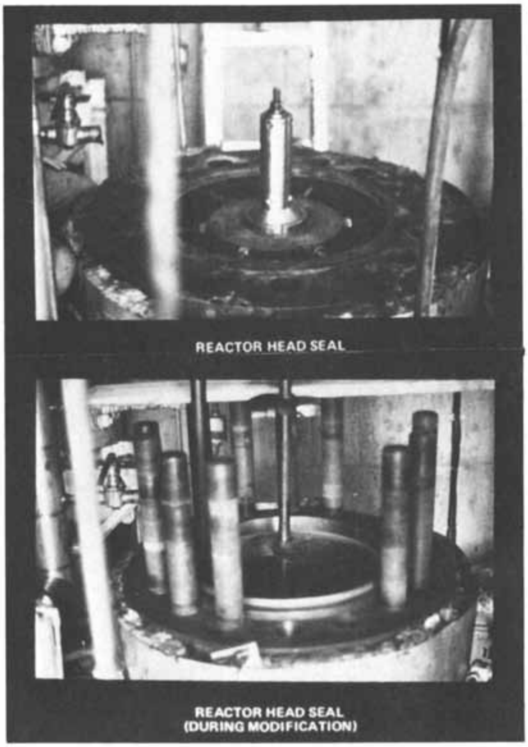

Figure 4. Reactor head seal

shows modification work on the head to shell closure of the stirred tank reactor.

This vessel is approximately 600 millimeters in diameter, and designed for pressures of up to 300 atomospheres. The problems uncovered required complete removal of the original seal mechanism, replacement of the vessel lining, and remachining for a new seal ring. All of the work was done in place, including the machining of the reactor head, which weighs over 1,000 kilograms and was located 15 meters above the ground.

The initial operation of the preheater uncovered difficulties with internal coking and eventual damage to the scraper blades themselves. Figure 5 illustrates the before and after condition of some of the 50 scraper blades that are used in this heat exchanger.

Degradation of polymeric materials caused numerous problems in the plant. The product oil or the anthracene vehicle oil which was used for startup rapidly attacked gaskets, seal parts, and O-rings, as illustrated in Figure 6. Replacement of those parts was required in almost every seal and pressure boundary part in the plant.

On the basis of information acquired during the commissioning, a number of equipment and process modifications were made. Figure 7 illustrates the changes that were made in the flow scheme as follows:

- The stirred tank reactor was bypassed due to continuing problems with the agitator shaft seal
- The preheater vessel alone was then used for heating and reaction
- Carbon monoxide and catalyst solution were introduced ahead of the preheater
- Liquid effluent bypassed the centrifuge since the centrifuge was unable to handle the reactor effluent

Process Results

In spite of obstacles that were encountered, valuable operating time has been logged - recently, exceeding 50 percent — and important process results are being accumulated. The first area studied was reaction characteristics which included the effects of system temperature, pressure, carbon monoxide concentration, residence time, and catalyst use. These have been evaluated for their effect on wood conversion, overall yield, and product quality. Figure 8 illustrates preliminary results on the rate of wood conversion per pass through the preheater vessel.

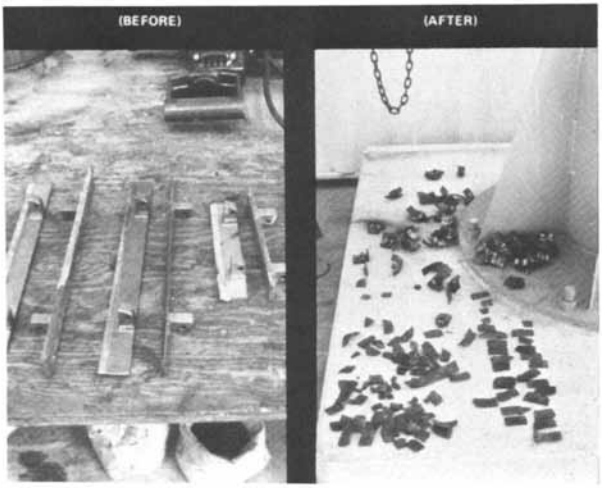

Figure 5. Reheater scraper blades

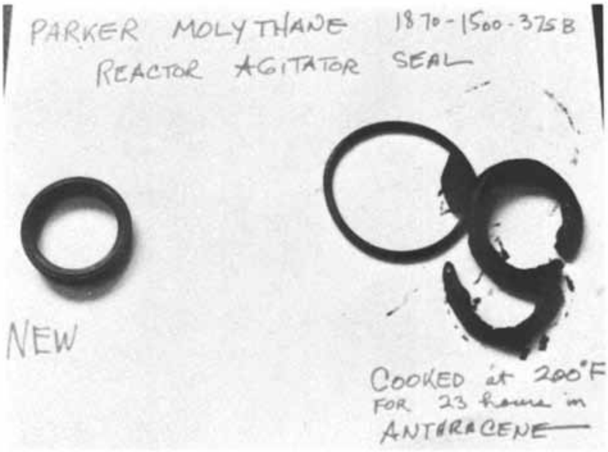

Figure 6. Decomposed reactor rubber seal material

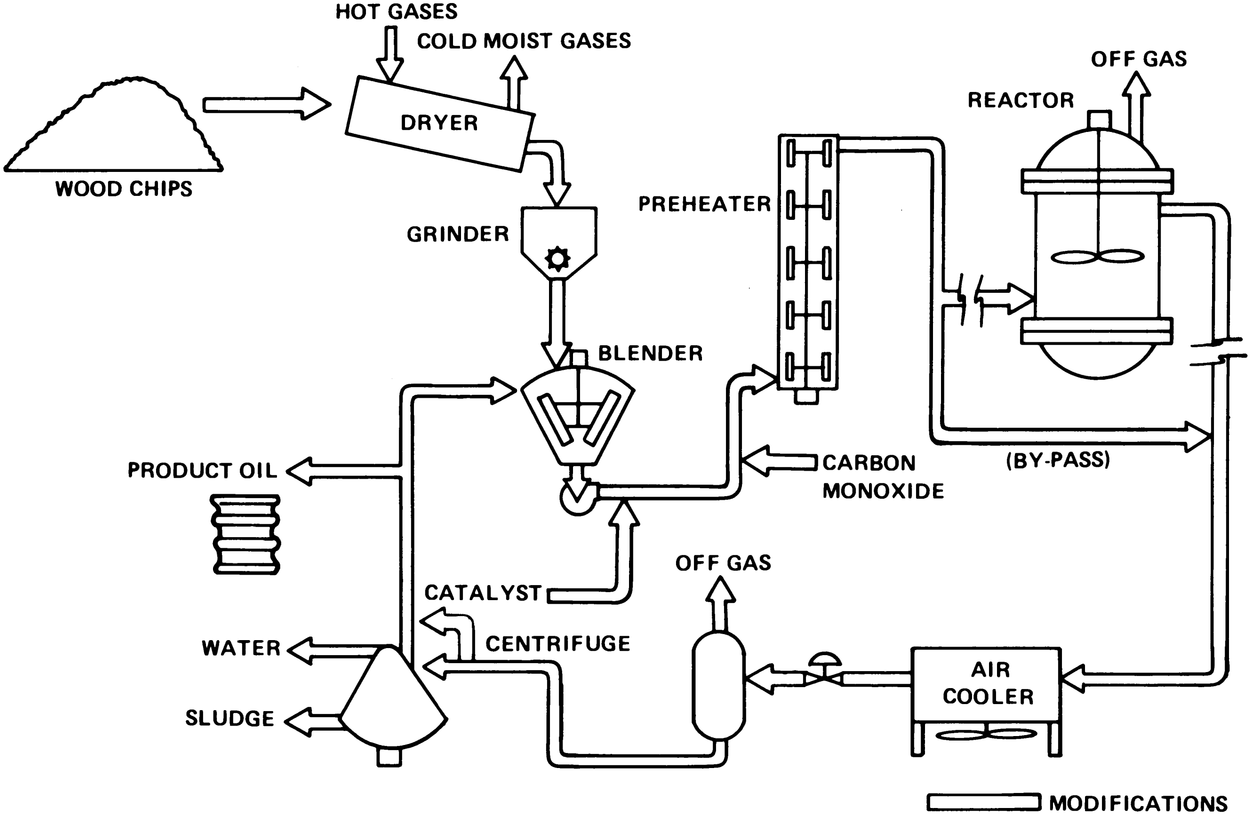

Figure 7. Albany PDU schematic (modified)

Wood conversion, K, is defined as:

$$K = \frac{Cf - Co}{Cf} \times 100$$

where K is wood conversion in %

Cf is the undissolved solids in the feed

Co is the undissolved solids in the outlet.

It should be noted that conversion does not necessarily mean oil production but simply the disappearance of wood. Residence time is the overall residence time in the preheater and not time at temperature. It should be noted that the temperature range in the preheater is from 100° to approximately 350°C. As Figure 8 illustrates, at approximately 25 minutes residence time, 70 percent conversion per pass is usual. Since the slurry is recycled on an approximate 5 parts recycle to one part product ratio, overall wood conversion is over 90 percent. Under the conditions tested so far, the approximate capacity of the plant is 1,000 kilograms a day of wet wood feed with 400 to 500 kilograms of oil produced. This is shown schematically in Figure 9, which illustrates the overall plant mass balance. As shown, the main reactants are carbon monoxide and wood while the main products are oil and flue gases. In a commercial sized plant, as will be illustrated later, only wood would be used as a reactant.

To date, the product oil that has been produced has been quite viscous and resembles heavy heating oil in its consistency. Table 3 illustrates a few of the more important properties of the product oil itself. At times, increasing viscosity of the product oil has necessitated termination of test runs. Methods to control this phenomenon are now being examined and will have high priority in future experiments.

As can be seen, the heating value is similar to a bunker fuel oil. The elemental analysis shows that the oxygen content is approximately 8 percent as compared to the wood flour used to feed which has over 40 percent oxygen. The net effect of conversion is that the heating value is approximately twice as high as wood on a weight basis and over four times as high on a volume basis.

Commercial Plant Concepts

With the information learned at Albany, a preliminary conceptual design has been developed. Some of the criteria that have been incorporated into this conceptual design are the amount of carbon monoxide consumed, the use of the slurry recycle operation, the technique of CO addition ahead of the preheater, and a knowledge of the slurry and off-gas characteristics. Figure 10 illustrates the main features of a commercial flow scheme.

While the carbon monoxide gas is brought in trucks to the Albany plant, a commercial facility will use a synthesis gas contain-

Table I. Equipment Commissioning and Proof Testing

• PRESSURE VESSELS	7
• PUMPS	15
• HEAT EXCHANGERS	6
• WOOD HANDLING UNITS	6
• ELECTRIC MOTORS	40
• CONTROLS	60

Table II. Mechanical Difficulties

- PUMP AND AGITATOR SHAFT SEALS
- GAS COMPRESSORS
- ELECTRICAL PROCESS HEATING
- WOOD HANDLING EQUIPMENT
- REACTOR HEAD SEAL
- PREHEATER SCRAPER BLADES

Table III. Product Oil Properties

• VISCOSITY	200-1000 cp	
• SPECIFIC GRAVITY	1.05-1.10	
• BOILING POINT	280°C	
• HEATING VALUE	15,000 Btu/lb	
• ELEMENTAL ANALYSIS	% C	77.2
	% H	6.5
	% N	0.4
	% O	8.4

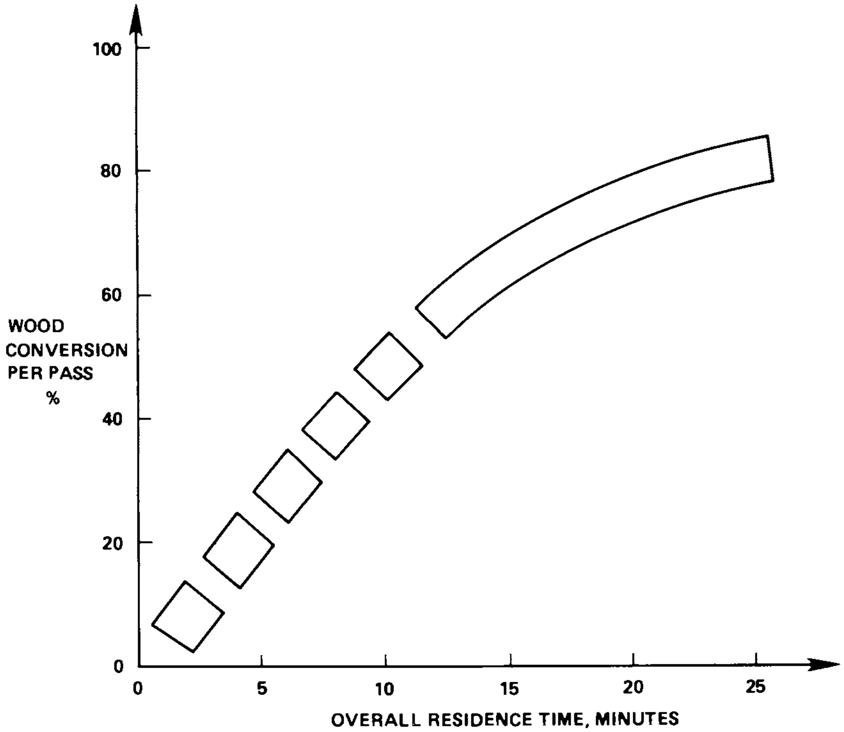

Figure 8. Rate of wood conversion

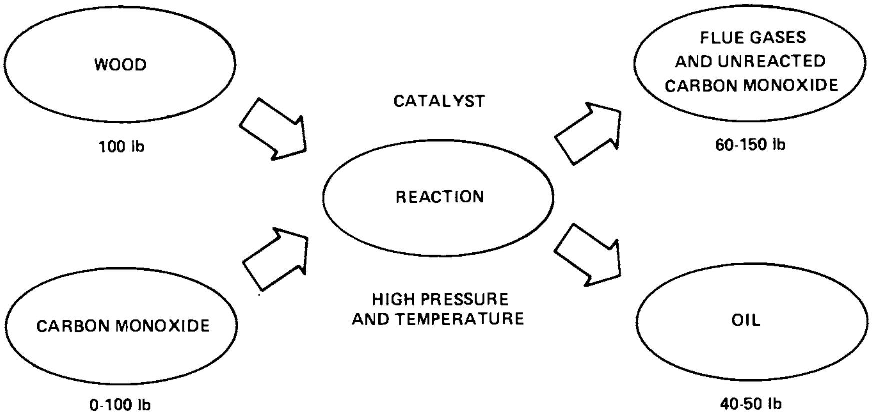

Figure 9. Albany PDU mass balance

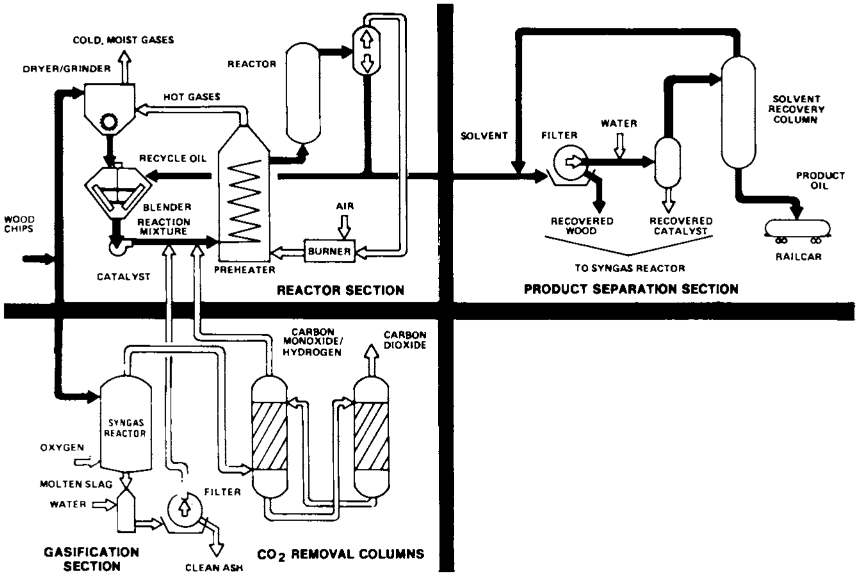

Figure 10. Biomass liquefaction conceptual commercial process

ing both hydrogen and oxygen. A synthesis gas plant and a product separation system have been added to the commercial facility design.

The synthesis gas is produced in an oxygen blown gasifier and is treated for removal of water and carbon dioxide. The synthesis gas also treats the catalyst solution for recovery of the alkali. In a commercial plant about half the wood that is introduced as feed goes for synthesis gas production. The other half of the wood is treated by drying and grinding and then blending with recycled oil prior to pumping into the reactor system as in the Albany plant.

A helical coil preheater followed by a holdup tank is also currently part of the design. The preheater is heated by combustion of system offgases. After pressure letdown from the reactor hold tank, some of the slurry is recycled to the blender while the product portion is routed for purification. Here the product, containing some unreacted wood as well as water and catalyst, is first diluted with a light solvent. After dilution, the mixture is filtered for the removal of solids. The liquid stream is then further diluted with water to wash out any remaining catalyst. Both the solids and the aqueous waste stream are routed to the synthesis gas reactor for treatment. The organic stream is sent to a fractional column for recovery of the solvent. At this point the still bottom becomes the product.

Figure 11 is an artist's rendering of a commercial plant envisioned in a Pacific Coast setting. In the foreground a large pile of wood chips is shown. These are assumed to be prepared in the forest. From the pile, the wood is distributed by conveyors to the synthesis gas reactors and to the drier-grinder apparatus. In addition to the major facilities for oil production, administration building repair shops, cooling towers and wastewater facilities are also illustrated.

It is important to note that the liquefaction plant should be an environmentally good neighbor. Some of the highlights of the design will serve to illustrate:

- Biomass and air are the only raw materials
- Oil is the only product
- The effluents consist of only water, clean flue gases, and inert ash
- The process is efficient from an energy standpoint

Calculations to date show that overall yield for a commercial plant should be approximately 35 percent wood to oil, while energy balance would indicate approximately 54 percent, as illustrated in Figures 12 and 13.

Economic Analysis

In order to develop operating and capital costs, several plant sizes were costed out. Figure 14 illustrates the effect of plant

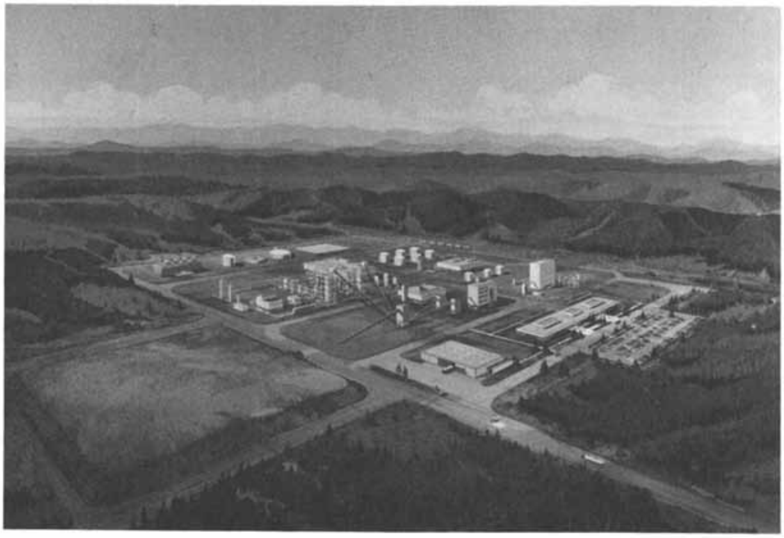

Figure 11. Artist's rendering of a commercial biomass liquefaction plant

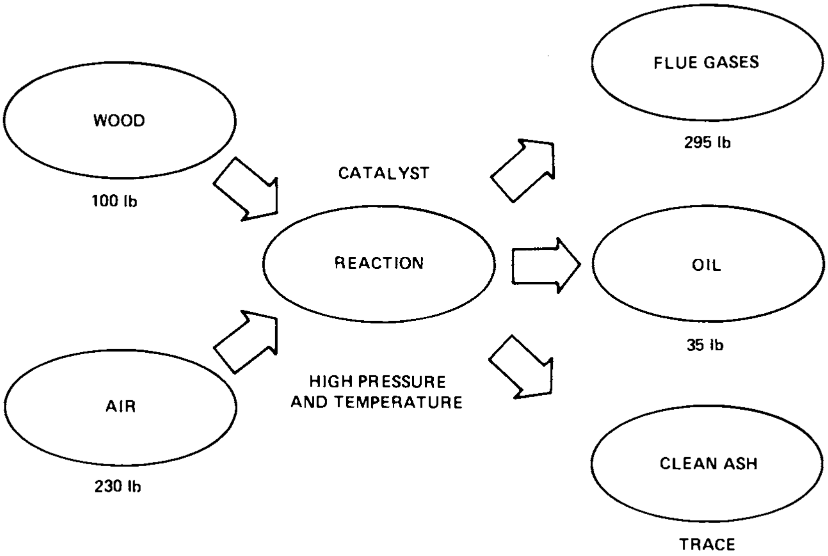

Figure 12. Biomass liquefaction mass balance

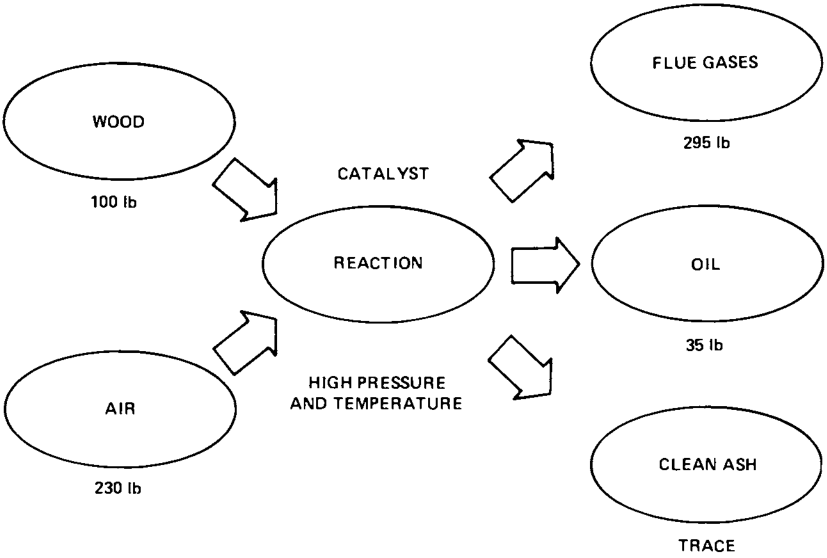

Figure 13. Biomass liquefaction energy balance

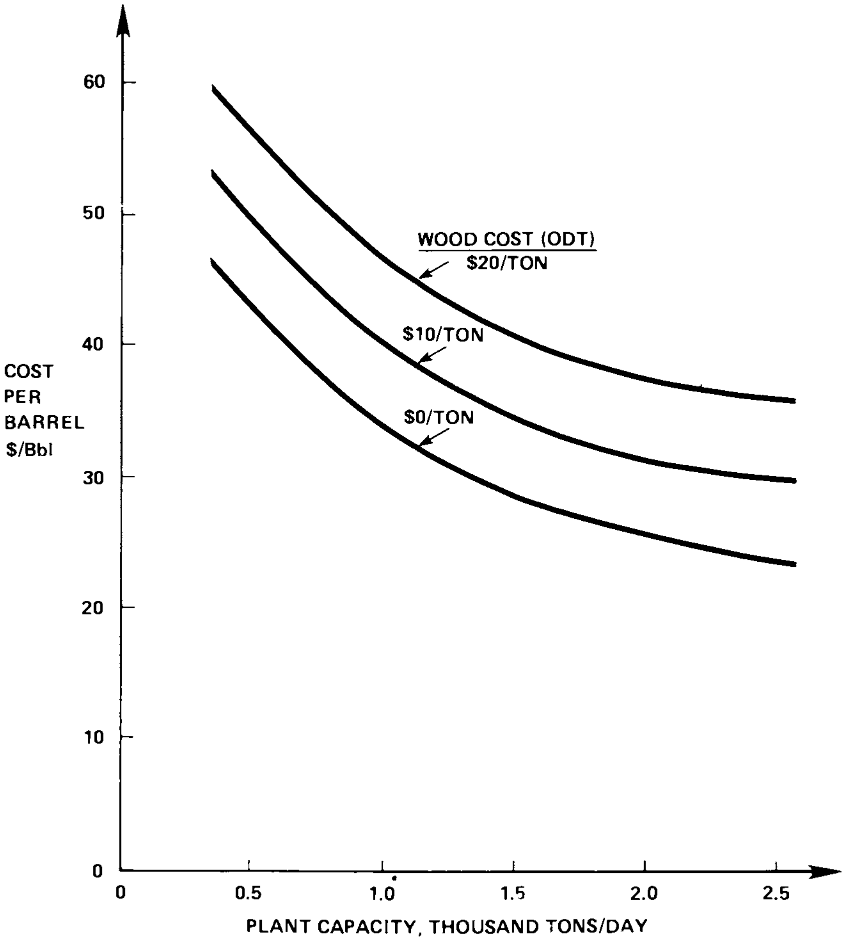

Figure 14. Effect of plant size on production cost

size on product cost. As can be seen, due to the plant complexity, the economies of scale are quite strong. It would appear that only liquefaction plants of large production (in excess of 1,000 tons per day) will be attractive. Figure 14 illustrates the relative portion of plant capital attributable to the major process sections of a commercial plant. This breakdown was based on a plant of approximately 1,000 metric tons of oven-dried wood a day. The costs of the wood handling and synthesis gas plant are the largest single items in the plant.

The portion of the plant cost for the high pressure reaction equipment is only 16 percent of the total. This would tend to indicate that future development work should center on reduction in the size of the synthesis gas plant and a simplification of wood handling.

Based on a plant of this size, Figure 15 shows the relative cost portions of the product that can be allocated to amortization of capital, feed stock, electric power, taxes and insurance, labor and consumables. Amortization of capital is by far the largest factor followed by the cost of feed stock. In this instance, wood is assumed to carry a value of $20 per oven-dried ton. Overall, this results in a breakeven production cost of just under $35 per barrel.

Future Activities

The past year's efforts and the development of the first conceptual designs have shown a number of areas of fruitful continued efforts as illustrated in Figure 16. The overall production costs are very sensitive to plant yield and capital cost. In the following program substantial efforts will be aimed at reducing the amount of synthesis gas required, raising overall yield and efficiency, and reducing energy input.

Figure 17 is a projection of the effect of future activities on the overall cost of biomass produced oil.

It is expected that continued developments at the Albany facility will yield improvements that should show a reduction in projected product price to approximately $25 a barrel. This should take place over the next 2-1/2 years. At this point it is suggested that a pilot plant of approximately 100 to 300 tons per day capacity be built and operated. This pilot plant would have all the facilities now envisioned for a commercial plant but of much smaller size. Information gained by the operation of this plant should reduce the product cost as expressed in constant 1978 dollars to $20 per barrel. This would take approximately five years. The final step in the development would be the design and construction of a demonstration plant utilizing full-size process trains. Information gained here should provide the means to be commercially competitive by approximately 1990.

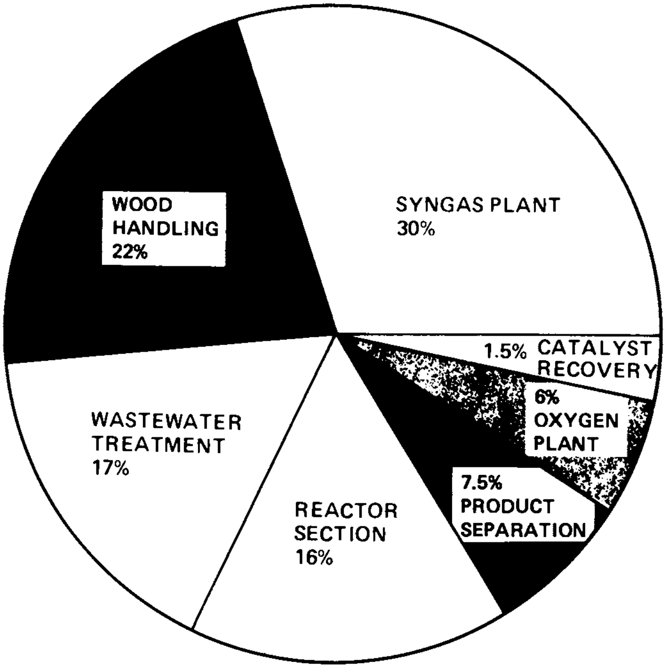

Figure 15. Capital cost breakdown—commercial biomass liquefaction plant

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

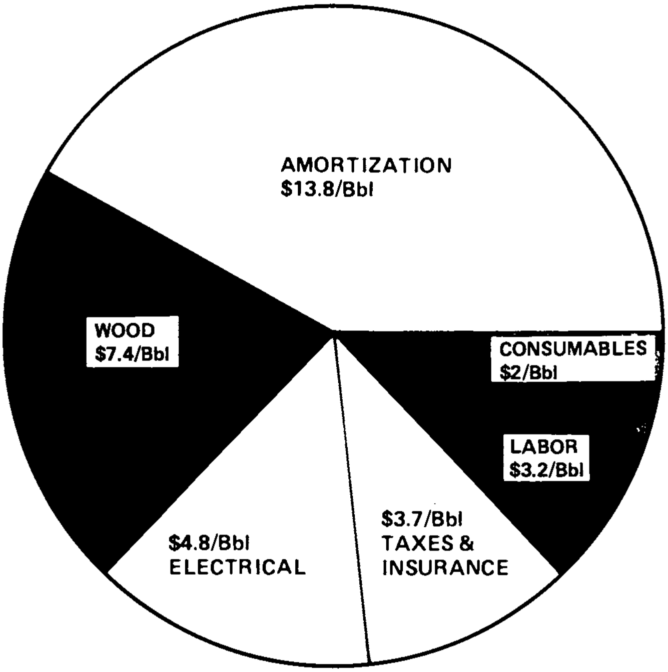

Figure 16. Oil production (break-even) costs—commercial biomass liquefaction plant

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

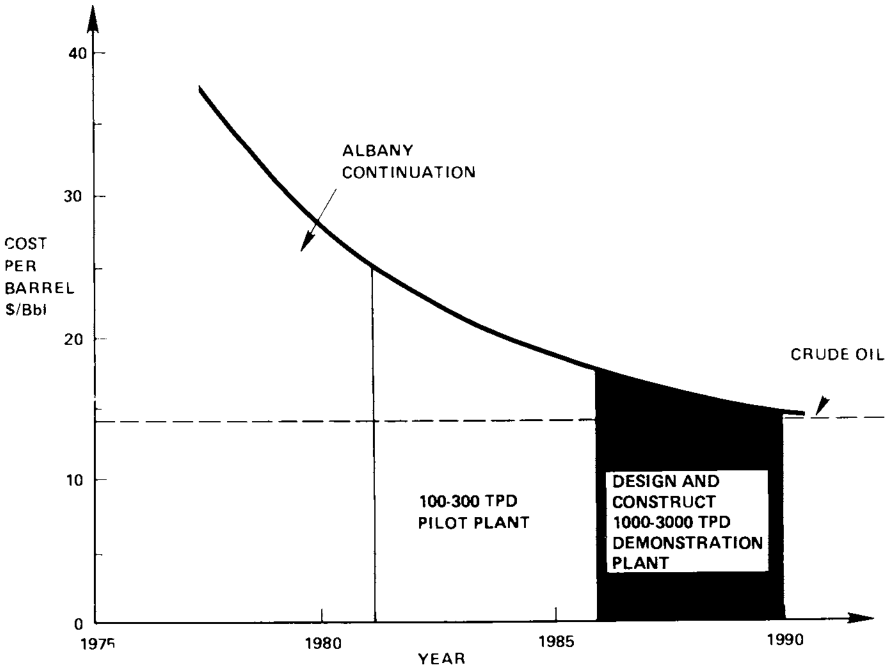

Figure 17. Projected process economics

Summary

Investigations to date have led the authors to be optimistic about the possibilities of oil production from biomass. While difficulties in bringing the current facilities on-stream have somewhat limited information to date, it is felt that a vigorous activity in the future can eventually provide a new source of energy for the country in the form of oil from biomass.

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MARCH 15, 1978.

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Pyrolysis for the Production of Activated Carbon from Cellulosic Solid Wastes

PAUL V. ROBERTS and JAMES O. LECKIE
Department of Civil Engineering, Stanford University, Stanford, CA 94305
PAUL H. BRUNNER
EAWAG, Swiss Federal Institutes of Technology, CH 8600 Dübendorf, Switzerland

Municipal solid wastes can be pyrolyzed to produce a carbonaceous, solid char (1). Such a char potentially could be activated to generate a product with large internal surface area and favorable adsorptive properties. The production of activated carbon in this manner offers the combined benefits of reducing the volume of solid wastes with a minimum of air pollutant emissions and producing a valuable product for a growing market in water and wastewater treatment. Presently, carbonaceous raw materials such as bituminous coal, lignite and peat are used in the manufacture of activated carbon (2). The properties of the final product are strongly influenced by the chemical composition and physical structure of the raw material (2, 3).

If solid wastes are to be used as raw materials for activated carbon manufacture, it is important to understand the effects of waste composition, pyrolysis conditions, and activation conditions on the yield and properties of the intermediate product, char, and the final product, activated carbon. Municipal solid wastes vary widely in composition both geographically and temporally (4, 5). Their potential for activated carbon preparation can be gauged crudely from their content of organic material. Municipal solid wastes contain from 50 to 75 percent organic material, the primary constituent being paper (5). Thus, cellulose is a logical choice for a model substance to investigate the behavior of solid wastes in pyrolysis. The elemental composition of the organic fraction of solid wastes can be given approximately as $C_6H_{10}O_4$ (4, 5, 6), which agrees closely with the empirical formula for cellulose $(C_6H_{10}O_5)_n$ (6).

Objectives

This work was undertaken to:

1. Determine the effects of pyrolysis and activation conditions on the yields and properties of char and activated carbon prepared from cellulose and municipal solid waste.

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2. Ascertain the influence of the composition of the educt on the properties and yield of the final product.
3. Evaluate the effectiveness of air classification in removing inorganic impurities and improving the quality of the final product.
4. Evaluate the feasibility of producing from municipal solid waste an activated carbon comparable in quality to commercially available products.

Experimental Procedure

Twenty-gram samples of purified α-cellulose were pyrolyzed in a 500-ml quartz glass retort which was placed in a temperature-controlled muffle furnace and purged with argon (6). Two analytical grade cellulose products were compared: Merck microcrystalline cellulose for packing of chromatographic columns (designated hereafter as Cellulose A), and Schleicher and Schuell No. 2181 filter paper for preparative chromatography (designated as Cellulose B). The characteristics of the two α-cellulose products are given in Table I. The samples were first dried at 65°C. The ash content was estimated as the residue after heating at 550°C for 24 hours. The metal contents were determined after digestion with perchloric acid.

The char formed by pyrolysis of cellulose was activated in a steam-CO_2 atmosphere at temperatures in the range of 700 to 1000°C. Char samples weighing one to three grams were placed on a glass frit in the neck of an inverted 750-ml quartz bottle, which was heated in the same furnace used for pyrolysis.

Table I
Characteristics of Cellulose Used in Pyrolysis Experiments
(Reference 6)

	Cellulose A	Cellulose B
	Merck microcrystalline chromatographic cellulose	Schleicher and Schuell 2181 filter paper
Carbon content, % w/w	44.4	44.3
Hydrogen content, % w/w	6.2	6.2
Ash content, ppm	65	750
Na	not determined	175
Ca	"	67
Mg	"	21
K	"	5
Particle size	90% w/w < 100 μm	4.6 mm thick × 10.5 mm square
Density, g/cm^3	0.38	0.69

Thermoanalytic investigations of cellulose pyrolysis were conducted with a Mettler Thermoanalyzer, Model 1, which simultaneously registered signals for thermogravimetry (TG), differential thermogravimetry (DTG) and differential thermal analysis. In these experiments, 40 mg of sample were heated in an argon atmosphere at a controlled rate of 0.5°, 1.5°, or 4°C/min to a temperature of 550°C.

Mixed solid waste from the City of Schweinfurt, Germany, was pyrolyzed and the char was activated under the conditions described above for comparison with cellulose (6). This solid waste was chosen by virtue of its relative homogeneity, owing to its preparation that included hand and magnetic sorting, shredding, and screening through a sieve having 50-mm diameter openings. The sorted and screened waste contained 23.9 percent carbon, 3.2 percent hydrogen, and 51.2 percent ash on a dry weight basis. The water content was 38.6 percent.

Air-classified solid waste from the City of Palo Alto, California, was chosen for study because of its relatively high fraction of organic material. The average ash content was slightly less than 20 percent on a dry weight basis. Large sample sizes (approximately 1 kg) were a necessity because of the heterogeneity of the material. Pyrolysis was carried out in a stainless steel retort heated in a nitrogen atmosphere. The organic content of the char was then beneficiated by air classification in the apparatus shown in Figure 1 (7). The air-classified char was then activated in an atmosphere of steam and/or CO_2 in the same apparatus used for pyrolysis (7). The activated carbon prepared from air-classified solid waste was ground in a ball mill and screened through a 100-mesh sieve before being characterized.

The specific surfaces of char and activated carbon samples were estimated by B.E.T. analysis of nitrogen adsorption data obtained with a Ströhlein Area Meter in the experiments with cellulose and unclassified waste (6). Products derived from classified solid waste were characterized by means of an Orr Surface Area and Pore Size Analyzer, Model 2100 (7).

Results and Discussion

Pyrolysis of Cellulose and Solid Waste. The effect of pyrolysis temperature and rate of heating on the yield, composition and specific surface area of char was investigated (6). The objective of these experiments was to find a set of pyrolysis conditions resulting in a favorable combination of yield and properties of char, considering it as an intermediate product to be activated.

The char yield is shown as a function of pyrolysis temperature in Figure 2A. The behavior of solid waste as well as cellulose can be represented by sigmoid curves, with a sharp decrease in yield between approximately 225°C and 375°C. This agrees with observations that cellulose undergoes dehydration and cleaving of -(OH) and -(CH_2OH) groups at temperatures below 250°C, followed by

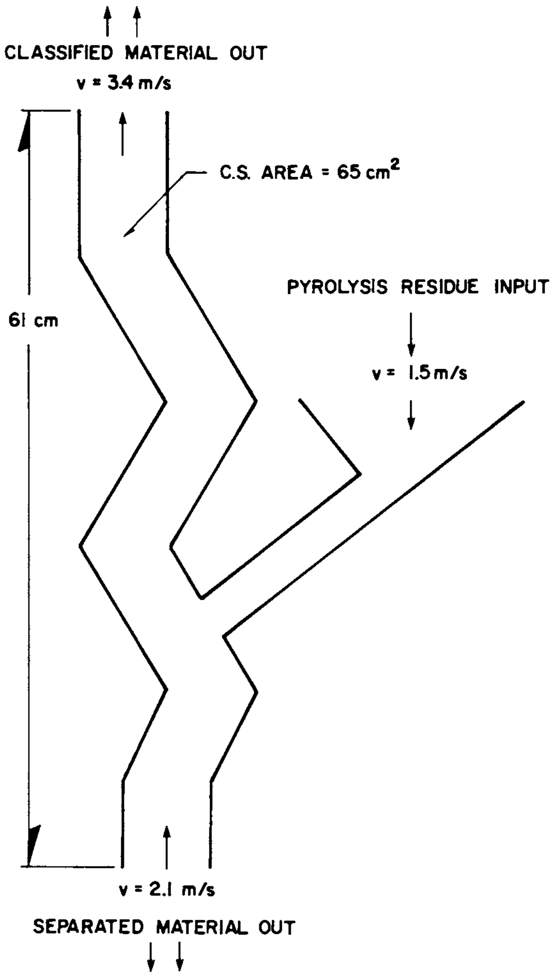

Figure 1. Configuration of and air velocities in the modified zigzag air classifier used for separation of inerts from char

depolymerization in the range of 250°C to approximately 370°C (8, 9). Accordingly, the carbon content of the char increases most rapidly with increasing temperature between 250°C and 370°C, as shown in Figure 2B. Above 400°C, the ring structures resulting from depolymerization are further dehydrated and condensed to 6-member rings with a C/H ratio of 2.5 (8). Above 550°C, graphitization proceeds, resulting in a rise in the C/H ratio to values in excess of ten at 750°C (8).

The organic fraction of solid waste behaves similarly to cellulose, as seen in Figures 2A and 2B. The principal differences are that the onset of pyrolysis of solid waste occurs at a slightly lower temperature and the yield of char is higher than for cellulose at temperatures above 500°C. Of course, the yield of char from solid waste on a total weight basis is substantially higher than from cellulose because of the difference in ash content. However, when compared using the weight of volatile solids as a measure of potentially pyrolyzable, organic material, the yields are only slightly higher for solid waste than for cellulose. Furthermore the carbon content of the char derived from solid waste shows much the same dependence on temperature as is the case for cellulose chars, after adjustment for the difference in volatile solids content of the educt. The carbon content of the volatile solids in unpyrolyzed solid waste is higher than in cellulose. This accounts for the relatively high carbon content of the solid waste "char" at temperatures below 300°C. The higher carbon content (on an adjusted basis) of the solid waste char compared to cellulose chars above 600°C may be explained by bias introduced in using the weight of volatile solids as a basis. A portion of the material remaining as "inert residue" in the 550°C ignition of the volatile solids determination is actually decomposed to volatile products during pyrolysis at temperatures above 600°C.

The most direct measure of the efficiency of pyrolysis in converting carbonaceous materials into an activated char is given by the yield based on carbon. The percent yields based on carbon for cellulose and solid waste are shown in Figure 2C. The yields are comparable below 400°C, but at temperatures above 600°C the carbon yield from solid waste is greater than from cellulose by a factor of two.

To ascertain the effect of the rate of heating during pyrolysis, experiments were conducted with a controlled rate of temperature rise. The results are summarized in Table II. Increasing the heating rate resulted in reduced yields expressed as either total weight or weight of carbon, and in approximately the same proportions. The carbon yield in cellulose pyrolysis was reduced by half when the heating rate was increased by a factor of 100, whereas the carbon yield from solid waste was less strongly affected. The fraction of carbon in the char was not affected appreciably by changing the heating rate. The dependence of char yield on heating rate is shown in Figure 3. For cellulose, a linear relationship was found of the following form:

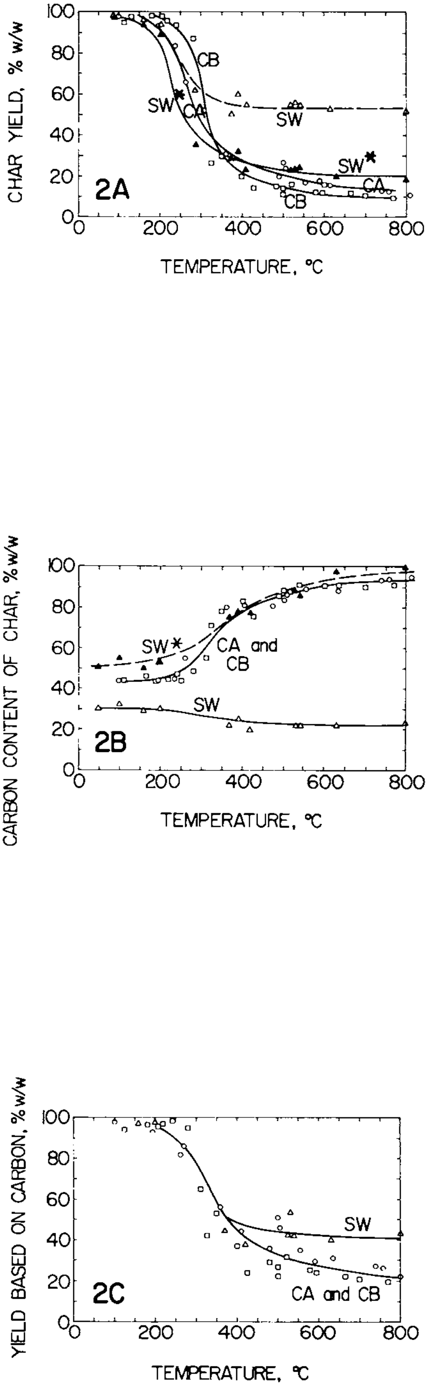

Figure 2. Dependence of char yield (2A), carbon content of char (2B), and carbon yield (2C) on the pyrolysis temperature. Data are shown for Cellulose A (CA, ○); Cellulose B (CB, □); and unclassified solid waste on the basis of total weight (SW, △) as well as adjusted to the weight of volatile solids (SW, ▲), expressed as g volatile solids in char per g volatile solids in educt in Figure 2A or g C per g volatile solids in char in Figure 2B. These data were obtained in batch pyrolysis experiments conducted at a high heating rate, $\phi \approx$ 100°C/min.*

Table II
Effect of Heating Rate on the Yield and Composition of Char
(Reference 6)

Educt	Final Temperature °C	Heating Rate, °C/min	Yield, % Based on		Carbon Content of Char, % w/w
			Total Weight	Carbon	
Cellulose A	540	0.13	31.1	61.9	88.5
		0.14	31.9	65.0	90.6
		0.48	29.3	61.4	93.0
		0.89	28.4	59.7	93.5
		1.59	26.6	55.8	93.3
		2.20	26.5	54.7	91.6
		19.0	19.8	40.4	90.8
		max*	17.2	34.6	89.3
Cellulose A	710	0.2	29.2	62.2	94.8
		0.42	27.0	57.3	94.2
		1.2	25.9	56.4	96.7
		1.7	24.7	53.3	95.9
		43.0	12.2	25.7	94.3
		max*	13.0	27.3	93.5
Cellulose B	710	0.5	23.7	52.9	99.4
		1.7	21.9	47.3	95.8
		5.35	20.3	43.1	94.6
		78.0	12.3	26.4	95.2
		max*	10.2	20.6	89.6
Unclassified Solid Waste	710	0.2	65.3	47.2	16.9
		0.5	65.4	46.9	16.8
		1.1	64.4	42.3	15.4
		32.0·	63.2	39.6	14.7

*Samples were heated to the pyrolysis end temperature at the maximum possible rate by introducing them into the furnace after it had been preheated to the designated temperature. The initial heating rate was approximately 100°C/min.

$$y = -2.4 \ln \phi + C$$

where

y = weight of char expressed as percent of educt
ϕ = heating rate, °C/min
C = empirical constant
= 27.9 for Cellulose A at 540°C end temperature
= 25.3 for Cellulose A at 710°C
= 22.9 for Cellulose B at 710°C.

The specific surface of the char was likewise found to be dependent on both end temperature and heating rate. Data for Cellulose A are shown in Figure 4. The specific surface increases with increasing temperature above 400°C, reaching a maximum of approximately 350 to 400 m^2/g at 700°C. Above 700°C, the specific surface declines with further increases in temperature. The decline is more pronounced at high heating rates than at low heating rates. For chars derived from solid wastes, the specific surface at 700°C was approximately 10 m^2 per gram of char, or 30 to 80 m^2 per gram of volatile solids.

The kinetics of pyrolysis were investigated using thermogravimetry (10, 11). Interpretation of overall rate data from complex systems of solid phase reactions is perilous; however, the comparison of activation energies so derived is permissible among experments that are conducted under comparable conditions as in this work (11). The results were interpreted by the approach suggested by Freeman and Carroll (10, p. 44, 12), who proposed that the activation energy and order of a decomposition reaction can be estimated using the equation

$$\frac{\partial[\ln(d\alpha/dt)]}{\partial T} = -\frac{E}{R}\frac{\partial(1/T)}{\partial T} + n\,\frac{\partial[\ln(1-\alpha)]}{\partial T}$$

where α is the fractional weight loss; T is the absolute temperature; R is the gas constant; n is the order of reaction; and E is the activation energy. The activation energy can be obtained from plots of

$$\frac{\Delta \log(d\alpha/dt)}{\Delta \log(1-\alpha)} \quad \text{against} \quad \frac{\Delta(1/T)}{\Delta \log(1-\alpha)}$$

The activation energies for pyrolysis of cellulose and solid waste are summarized in Table III. The values in this work range from 20 to 70 kcal/mol, agreeing approximately with activation energies previously reported for pyrolysis of cellulose (13, 14, 15, 16). The heating rate was found to have no effect on activation energy. The activation energies differed substantially for the different educts. The highest value was found for the educt having the lowest ash content, Cellulose A, while the educt with highest ash content, i.e. the solid waste, exhibited the lowest activation energy.

The differences in activation energies presumably can be attributed to a catalytic effect of inorganic impurities present in

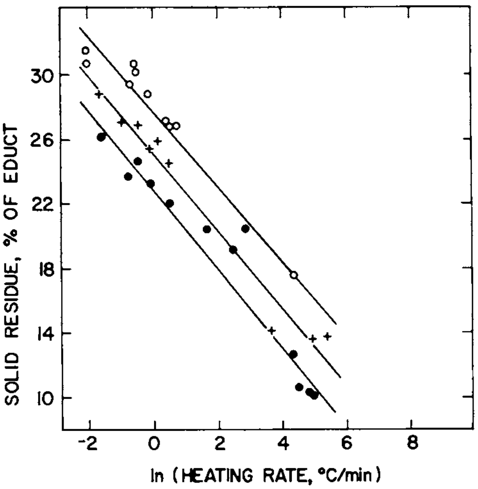

Figure 3. Effect of heating rate during pyrolysis on the yield of char: ●, Cellulose B at 710°C; +, Cellulose A at 710°C; ○, Cellulose A at 540°C

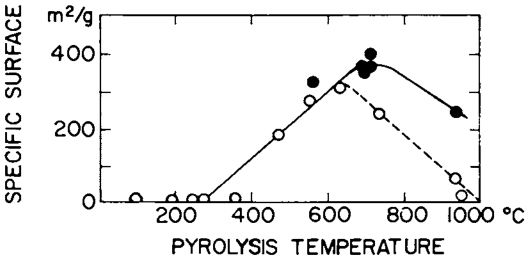

Figure 4. Specific surface of char prepared from Cellulose A, expressed as m² per g of char: —●—, ~ 1°C/min; - -○- -. >> 1°C/min

Table III
Activation Energies for Pyrolysis
(Reference 6)

Educt and Ash Content	Heating Rate, °C/min.	Activation Energy kcal/mol	
Cellulose A (65 ppm ash)	4.0	69.9	average = 70.6
	1.5	70.0	
	0.5	71.8	
Cellulose B (750 ppm ash)	4.0	38.3	average = 38.0
	1.5	37.3	
	0.5	38.8	
Unclassified Solid Waste (51.2% ash)	4.0	21.2	average = 21.0
	1.5	20.1	
	0.5	21.7	

the educt. The addition of inorganic substances that are Lewis acids or bases has been shown to reduce the activation energy for pyrolysis (13, 14, 16, 17, 18). It has been postulated that such additives promote the rate of dehydration of cellulose compared to depolymerization, consequently increasing the yield of char at the expense of volatile pyrolysis products (18). When the concentration of additives is raised, a saturation limit is reached beyond which there is no incremental catalytic effect (17). No further reduction in activation energy and increase in char yield were observed when the additive content was increased from two to eight percent (18).

This explanation also may account for the higher yield of carbon from solid waste compared to cellulose, as was observed at temperatures above 600°C in Figure 2C. The inorganic matter present in solid waste contains appreciable amounts of phosphorus, aluminum, potassium, sodium and other elements that might have a catalytic effect (4).

Activation. Activation of pyrolysis chars with steam and CO_2 resulted in an increase in specific surface at temperatures above 700°C, as shown for cellulose-derived chars in Figure 5. The rise in specific surface is more pronounced in the case of Cellulose B than for Cellulose A. The specific surface of activated carbon derived from Cellulose B at an activation temperature of 950°C was 1550 m^2 per gram, nearly twice as great as for activated carbon made from Cellulose A. The corresponding value for activated carbon from unclassified solid waste was comparable to that from Cellulose A when compared on the basis of carbon content, i.e., approximately 850 m^2 per gram carbon.

The weight loss during activation increases sharply between 700°C and 950°C, as is evident in Figure 6. At 950°C and activation times of 30 minutes or more, the weight loss exceeded 50 percent. For the activation of cellulose chars under the conditions in this work, the optimum combination of temperature and time of activation appears to be approximately 850 to 900°C and 30 to 60 minutes. In this range the weight loss is 35 to 55 percent and the surface area is 800 to 1100 m^2/g.

The activation yield for char derived from unclassified solid waste was much higher than for cellulose when compared on a gross weight basis. A lower limit on the yield from solid waste was imposed by the ash content of the char, approximately 80 to 85 percent by weight. At activation temperatures above 800°C, the carbon content of the activated product from unclassified solid waste was less than ten percent by weight.

Air Classification

Air classification of both raw, wet solid waste and of pyrolysis char was evaluated as a means of separating out inorganic components that are not amenable to carbonization and activation, with a view to increasing the carbon content and specific surface of the activated product (7).

Air classification segregated approximately half of the inorganic material in the raw wet waste. The organic content of the reject material was approximately ten percent by weight. Air classification of the char achieved a further reduction of 60 percent by weight of the ash remaining after pyrolysis. The inert fraction segregated in air classification of the char contained approximately 90 percent ash, as did that from raw waste classification.

The yield based on carbon was approximately 40 percent by weight in pyrolysis at 700°C. The yield based on carbon in activation with steam and CO_2 at 910°C and an activation time of 20 minutes was 40 percent by weight. These yields are virtually identical to those observed with unclassified solid waste at the same conditions. Nevertheless, the properties of activated carbon prepared from solid waste were improved markedly by utilizing air classification. The ash content was reduced from 90 percent to 60 percent by weight. The specific surface was increased from 80 m^2 per gram gross weight to 350 m^2 per gram. The specific surfaces of activated carbons derived from both classified and unclassified solid wastes were approximately equal when compared on the basis of carbon content, 800 m^2 per gram of carbon.

The efficacy of air classification in reducing the fraction of inorganic impurities in waste-derived activated carbon is demonstrated in Table IV. The overall yield of carbon in pyrolysis and activation is the same, approximately 25 percent, regardless of whether or not air classification is practiced. Nevertheless, the carbon content of the final product is 40 percent with air

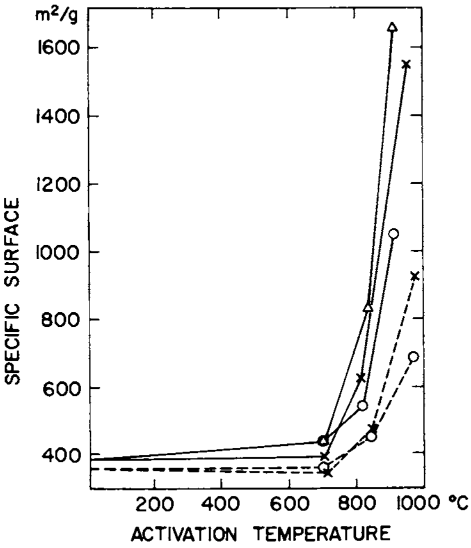

Figure 5. Effect of activation conditions on the specific surface of activated carbons prepared from cellulose. Specific surface is expressed as m² per g of activated carbon. Cellulose A, (– – –); Cellulose B, (——); ○, 15 min; ×, 30 min; △, 60 min.

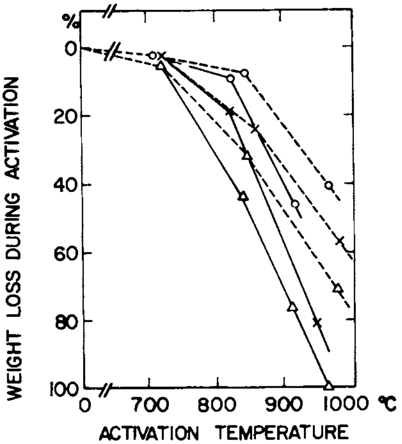

Figure 6. Weight loss during activation of chars derived from cellulose. Cellulose A, (– – –); Cellulose B, (——); ○, 15 min; ×, 30 min; △, 60 min.

Table IV
Yields and Product Compositions in Pyrolysis and Activation of Air-Classified and Unclassified Solid Waste

		Weight and Change in Weight, g					
		With Air Classification			Without Air Classification		
Material	Process	Total Weight, g	Ash, g	Carbon, g	Total Weight,* g	Ash, g	Carbon, g
Raw, wet waste*		100	30	20	100	30	20
	Air Classification	-16	-15	-0.2	n.a.	n.a.	n.a.
Air-Classified Waste		84	15	19.8	100	30	20
	Pyrolysis†	-58	-0	-11	-59	-0	-11
Char		26	15	9	41	30	9
	Air Classification	-10	-9	-1	n.a.	n.a.	n.a.
Beneficiated Char		16	6	8	41	30	9
	Activation†	-6	-0	-4	-6.5	-0	-4.5
Activated Carbon		10	6	4	34.5	30	4.5

*Basis: 100 g of raw, wet waste: 30% w/w H_2O; 40% w/w volatile matter with an empirical formula ($C_6H_{10}O_4$); and 30% w/w ash.

†Assumed yields based on carbon: 45% in pyrolysis and 50% in activation.

n.a.: not applicable.

classification of raw waste and char, compared to only 13 percent without air classification.

Comparison with Conventional Activated-Carbon Products. Powdered activated carbon derived from air-classified solid waste was compared to several commercially available, powdered activated carbons in terms of specific surface, ash content, and empirical measures of adsorption capacity. The results are shown in Table V.

Table V
Properties of Powdered Activated Carbon

Activated Carbon	Ash Content % w/w	Surface Area m^2/g	Iodine Number	Phenol Number	Methylene Blue Number	Relative COD Reduction Factor[†]
Carbon from Air-Classified Solid Waste[*]	60	350	500	15	120	1.00
Aqua Nuchar A	3.5	625	560	28	150	1.00
Calgon BL	8	1100	920	30	280	0.95
Darco M	16.5	630	455	21	125	0.95

[*] Activated with steam and CO_2 at 900°C.
[†] See definition in text.

The waste-derived material exhibited a substantially higher ash content and lower specific surface than any of the commercial products tested. Moreover, it was rated inferior according to conventionally accepted criteria for adsorption capacity such as the phenol and methylene blue numbers. However, the waste-derived carbon proved equivalent to the others when judged according to the criterion of relative capacity of reduction of chemical oxygen demand (COD) in primary-treated municipal wastewater. The relative capacity for COD removal was defined operationally as g COD removed from Palo Alto primary effluent (COD = 60 mg/l) per gram of adsorbent, averaged over a range of carbon dosages from 400 to 1200 mg/l and normalized with respect to Aqua Nuchar A as a reference product (7). The satisfactory performance of the solid waste carbon despite its low specific surface is surprising. The apparent anomaly can be explained in part by differences in pore size distribution. The waste-based carbon contains a higher proportion of its internal void volume in large-sized pores. A comparison with the reference carbon, Aqua Nuchar A, can be made with the data in Table VI. Because of the relatively low proportion of

Table VI
Distribution of Pore Volume

Activated Carbon	Percent of Pore Volume			
	Pore Radius $<25 \times 10^{-10}$m	Pore Radius $<50 \times 10^{-10}$m	Pore Radius $<100\times10^{-10}$m	Pore Radius $<500\times10^{-10}$m
Derived from Classified Solid Waste	15	30	45	75
Aqua Nuchar A	85	90	98	99

small pores ($< 25 \times 10^{-10}$m), the carbon from waste has a low specific surface, and hence a low capacity for adsorption of small molecules such as phenol. However, the large proportion of pore volume in the transitional size range (25×10^{-10} to 500×10^{-10}m) assures that the mass transport resistance will be low compared to that in carbon such as Aqua Nuchar A that has only a small fraction of such larger pores. This is an important factor affecting the extent and rate of adsorption from solution of large molecules such as are represented by the COD content of wastewater. According to the data of Dewalle and Chian (19), nearly sixty percent of the soluble organic matter in biologically treated effluents lies in the range of molecular weight greater than 500. Substances of such high molecular weight are hindered in transport through pores of radius less than 25×10^{-10}m, since the dimensions of the molecule are of the same order of magnitude as those of the space through which it is diffusing.

Feasibility of Activated Carbon from Solid Waste

It has been shown in this work that the quality of activated carbon prepared from unclassified municipal solid waste is so poor in terms of specific surface that further investigation is scarcely justified. On the other hand, there is promise that an activated carbon of acceptable quality can be made from municipal refuse if the carbon content is enhanced by a selective separation technique such as air classification.

In order for an activated carbon to be considered of high quality for water treatment use, the ash content should be less than 15 percent by weight. The corresponding maximum allowable ash content in the educt can be estimated. Optimistically assuming a 55 percent carbon yield in pyrolysis and 70 percent in activation, the limiting ash content in the educt would be 3.3 percent based on dry weight if there were no separation of inert material from the char. If a 60-percent rejection of the inert material in the char could be achieved by air classification or some other means, the permissible ash content of the educt would be 8 percent.

In Solid Wastes and Residues; Jones, J., et al.;
ACS Symposium Series; American Chemical Society: Washington, DC, 1978.

Thus, if it is desired to produce from solid waste an activated carbon that will be considered equivalent to the highest quality products available for water and wastewater treatment, attention should be centered on solid wastes having an ash content of 10 percent or less. An activated carbon of acceptable quality can be produced from solid waste having an ash content as high as 20 percent of dry weight, provided that inert materials are removed from the char. Pilot plant studies of air classification of a large number of municipal wastes indicate that the inorganic content can be easily reduced to 20 percent (4). Combining extensive preshredding with air classification, an ash content of less than 10 percent by weight is attainable (20).

It is to be anticipated that pyrolysis systems for the general purpose of municipal solid waste management will be designed for maximum reduction of waste volume and maximum production of energy values in the form of gaseous fuels. To fulfill those objectives, an approach based on rapid heating of the educt is rational. This will result in lower carbon yields than were obtained in this work using low heating rates. Highly efficient preclassification of the solid waste is a prerequisite for obtaining chars with even a modest potential for activated carbon manufacture from an operation oriented toward volume reduction and gas production.

Hence, the chars obtained from pyrolysis of municipal solid wastes under the conditions that probably will be used in practice are not ideal for conversion to activated carbon having properties suitable for water and wastewater treatment. Special attention must be devoted to the selection of the educt and the overall processing scheme should be optimized with activated carbon in mind.

Conclusions

Cellulose was found to be a suitable model compound that behaves similarly to the organic fraction of solid waste during pyrolysis and activation. Conversely solid waste behaves in a general way as expected for a mixture of cellulose and an inorganic ash.

The maximum specific surface of cellulose char was 350 to 400 m^2/g at 710°C. Activation of cellulose char with steam and CO_2 resulted in a product with a specific surface as large as 1650 m^2/g. An acceptable combination of activation temperature and time was found in the range of 850°C to 900°C and 30 to 60 minutes, where a product having a specific surface of 800 to 1000 m^2/g was obtained with a yield of 50 percent by weight. The overall yield of carbon in pyrolysis and activation was approximately 25 percent.

Inorganic constituents at low concentration in the educt appear to have a catalytic effect enhancing pyrolysis and activation. The inorganic impurities in cellulose and solid waste contain constituents that raise the char yield and increase the rate of pyrolysis. Hence, inorganic impurities present at levels of a few percent are beneficial for producing activated carbon. However, concentrations of inorganic material greater than ten percent in the educt adversely affect the properties of the final product, causing high ash content and low specific surface.

The yield and composition of char from unclassified solid waste were similar to those from cellulose when adjusted for the differing contents of inert ash. However, because of the high ash content of the educt, the activated carbon prepared from municipal solid waste without segregation of the inert material has a composition that is unacceptable for water and wastewater treatment. The carbon content of the final product is ten percent or less and the specific surface is less than 100 m^2 per gram.

Air classification of solid waste and char has a significant, beneficial effect in enhancing the carbon content of the final product. Activated carbon so prepared has a carbon content of 40 percent and a surface area of 350 m^2/g. It is comparable in quality to several low-priced commercial products for reduction of chemical oxygen demand in wastewater.

Manufacture of activated carbon from solid waste appears feasible under certain conditions. The educt should be selected for low ash content, heating during pyrolysis should be gradual, and inert material should be segregated from the char by air classification or an equivalent method. The char thus prepared can be activated by conventional thermal techniques. The optimization of pyrolysis for maximum volume reduction and gasification of solid waste at a minimum cost conflicts with the manufacture of high-quality activated carbon.

Acknowledgements

The experimental work concerning pyrolysis and activation of air-classified solid waste was conducted by M. K. Stevenson. The financial support of the U.S. Environmental Protection Agency, Contract No. CPE-70-129 and Grant No. R-803188, as well as the Canton of Zürich is gratefully acknowledged. R. Braun, P. Erni, P. Grabner, and J. Wetzel of the Swiss Federal Institutes of Technology and D. M. Mackay of Stanford University provided helpful advice.

Abstract

The preparation of activated carbon by pyrolysis and thermal activation of municipal solid waste was investigated. The organic fraction of the solid waste behaves similarly to cellulose. Satisfactory yields and specific surface were obtained by pyrolysis at 700°C and a heating rate of 1°C/min, followed by activation at 900°C with steam and CO_2. The overall yield based on carbon was 25 percent. The surface area of the activated product was 800 m^2 per gram carbon. Segregation of inert material by air classification of the municipal solid waste and its derivative char improved the carbon content of the activated product from ten to forty percent and increased the specific surface from 80 to 350 m^2/g. The

activated carbon so prepared was comparable to commercially available, powdered activated-carbon products in removing COD from wastewater.

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