Final Report

DOE/CE/15612-T8

COMBUSTION OF MUNICIPAL SOLID WASTES WITH OIL SHALE IN A CIRCULATING FLUIDIZED BED

Grant NO.
DE-FG01-94CE15612

Prepared by:
Synfuels Engineering & Development Inc.,
11158 W. 68th Way
Arvada, CO 80004

30 June 1996

ep
DISTRIBUTION OF THIS DOCUMENT IS UNLIMITED

MASTER

DD final.

ACKNOWLEDGEMENT OF SUPPORT

The research described in this report was supported by a grant from the U.S. Department of Energy (DOE) under Grant No. DE-FG01-94CE15612. Such support does not constitute an endorsement by DOE of the views expressed in the report.

DISCLAIMER

**Portions of this document may be illegible
in electronic image products. Images are
produced from the best available original
document.**

TABLE OF CONTENTS

	<u>Page</u>
1.0 EXECUTIVE SUMMARY	1
2.0 STATEMENT OF THE PROBLEM	3
2.1 Introduction	3
2.2 Keys to Market Entry	7
3.0 PROJECT OBJECTIVES	10
4.0 METHODOLOGY	11
4.1 MSW/Oil Shale Test Plan	11
4.2 Conduct of Tests	17
4.3 Technical Analysis	19
4.4 Economic Analysis.....	32
5.0 FINDINGS AND CONCLUSIONS	45
6.0 RECOMMENDATIONS	47
7.0 REFERENCES	48
APPENDIX A. TEST PLAN	A-1
APPENDIX B. HAZEN RESEARCH, INC. TEST REPORT	B-1
APPENDIX C. ACERC EVALUATION	C-1

LIST OF TABLES AND FIGURES

Page

Tables

Table 4.1-I.	Waste Source Characterization	14
Table 4.1-II.	MSW Severe Case	15
Table 4.1-III.	Target MSW with Predicted Ultimate Analysis	16
Table 4.2-I.	Ultimate, Proximate, Heating Value, and Chlorine Analyses of MSW and Oil Shale	18
Table 4.3-I.	Summary of Pollutant Removal Test Results	22
Table 4.3-II.	Ash Stabilization and TCLP Testing Results	24
Table 4.3-III.	Summary of 6-Inch Unit Test Data	26
Table 4.4-I.	Base Case Cash Flow	37
Table 4.4-II.	Sensitivity Analysis	43

Figures

Figure 2.1-1.	Conceptual Diagram of Solid Waste Treatment System	4
Figure 4.3-1.	Performance Summary	21
Figure 4.3-2.	Effect of Temperature on Absorption	28
Figure 4.3-3.	6-Inch Tests	30
Figure 4.3-4.	15-Inch Tests	30
Figure 4.3-5.	Combined Test Plug-Flow Fit	31

1.0 EXECUTIVE SUMMARY

The authors of this report have invented an integrated process for the treatment of municipal solid waste (MSW). In this process, after recycling steps to save usable materials such as aluminum, other metals, and glass have been completed, the resulting refuse-derived fuel (RDF) is co-combusted with oil shale in a circulating fluidized bed. The oil shale not only removes SO_2 , Cl^- , and other effluents, but also adds significant fuel content and constituents that can produce a useful cementitious ash. This raises the possibility of creating an environmentally beneficial and financially viable industry in which the RDF helps produce electrical energy, the volume of solid waste is greatly reduced in both volume and weight, and its potentially hazardous components can be encapsulated in a non-hazardous cement that could serve useful purposes. This would eliminate the main environmental problems associated with MSW and its associated waste streams.

As critical steps toward the long-range goal, the objectives of Contract DE-FG01-94CE15612 were to:

- Demonstrate the complete combustion of oil shale/RDF in a circulating fluidized bed combustor (CFBC)
- Demonstrate the ability of oil shale to absorb the undesirable acid gas components (SO_2 and Cl^-) created during the combustion of RDF
- Demonstrate the ability of the cementitious oil shale ash to encapsulate the "hazardous" components of the RDF ash
- Determine whether there are major unexpected scale-up problems associated with our concept.

Test work performed on two pilot scale CFBC units at Hazen Research, Inc. successfully demonstrated the superior acid gas sorbent capabilities of oil shale as an additive to RDF. The test work also demonstrated that the addition of oil shale to the input stream improves the performance of the reactor from an operational standpoint. Analytical work performed at Hazen also confirmed the cementitious characteristics of the oil shale ash. The results of these tests were then used to project the economics of a typical commercial system that would use RDF/oil shale to produce electricity and dispose of MSW in an environmentally advantageous manner. The tests and analyses have shown:

1. That 96 to 98.5 percent of the acid gases produced during combustion are removed during operations at temperatures between 1475 and 1500° F with absorbent/pollutant (A/P) ratios ranging from 3.0 to 4.5. These results indicate that the

process can not only meet but surpass the requirements imposed by applicable acid gas emission standards.

2. That the resulting ash, when made into cement pellets, was non-leachable by standard EPA tests; therefore, the ash can be disposed of in standard landfills or used for other purposes.
3. Assuming a tipping (dump) fee of \$30 per ton (well below the average tipping fee in the geographical areas of interest), the process is economical (discounted cash flow return on investment of 22.65 percent with payout in six years from design commencement, i.e., less than four years of operation).

The above conclusions were based primarily upon results of tests using a 6-inch internal diameter circulating fluid bed combustion (CFBC) test unit. A larger 15-inch unit proved easier to operate, but the results of tests on the larger unit were inconclusive due to problems with residence time at operating temperature and probable residual contamination in the 15-inch test unit. However, no intrinsic scale-up problems were evident.

More tests are needed to:

1. Determine the effect of changes in residence time,
2. Determine the effect of reduced-temperature operations (1400-1450° F range),
3. Confirm scale-up at correct residence times and temperatures.

Once these tests have been successfully completed, the system will be ready for commercial demonstration. We are now pursuing options for creation of a demonstration plant.

2.0 STATEMENT OF THE PROBLEM

2.1 INTRODUCTION

The problem addressed by our invention is that of municipal solid waste utilization. The dimensions of the problem can be visualized by the common comparison that the average individual in America creates in five years' time an amount of solid waste equivalent in weight to the Statue of Liberty. The combustible portion of the more than 11 billion tons of solid waste (including municipal solid waste) produced in the United States each year, if converted into useful energy, could provide 32 quads per year of badly needed domestic energy, or more than one-third of our annual energy consumption (Energy Information Administration/Annual Energy Outlook 1995). Conversion efficiency and many other factors make such a production level unrealistic, but it is clear that we are dealing with a very significant potential resource.

Recent data (Statistical Abstract of the United States) indicate that only about 15 percent of MSW is being combusted for energy recovery. As is described under TECHNOLOGY OVERVIEW below, the situation appears to be worsening.

However, the problem is not simply that a potentially valuable resource is being wasted: perhaps more important, it is that a growing environmental/land use problem is not being solved. Landfills that can meet new standards are not available in many areas. The number of MSW landfills in the United States has declined from about 20,000 in 1970 to less than 5,000 in 1995 (Waste Age, January 1996).

Although landfills vary in terms of the length of time before they fail, no problem is permanently solved by conventional treatment and disposal, and the costs of environmental cleanup of contaminated sites continue to increase. A DOE report noted that "between 1982 and 1989 the cost associated with remediating a given amount of contaminated media increased by more than a factor of ten" (reported in Science, 21 February 1992, p. 901).

Our invention, which envisions an integrated recycling and electricity generation facility built around the unique characteristics of our abundant oil shale resource, could revolutionize the situation. We estimate that this process (conceptualized in Figure 2.1-1) would be economically competitive over a region defined by a 700-mile radius centered on two oil shale resource centers: (a) that of western Colorado and Eastern Utah, and (b) the New Brunswick oil shales bordering on Maine. Based on this criterion, the western resource could serve a population of over 42 million, and the eastern resource a U.S. population of more than 44 million. Thus, approximately one-third of the U.S. municipal waste market is within reach of our process.

Figure 2.1-1

CONCEPTUAL DIAGRAM OF SOLID WASTE TREATMENT SYSTEM

TECHNOLOGY OVERVIEW

We can classify municipal waste handling under the following labels:

1. Simple remote disposal (untreated waste into landfills)
2. Separation for recycling followed by remote disposal of residuals
3. Separation for recycling followed by incineration (mass burning) and remote disposal of residuals
4. Separation for recycling followed by conversion to non-energy uses and remote disposal of residuals
5. Separation for recycling followed by waste to energy conversion and remote disposal of residuals.

Remote disposal of untreated MSW has the advantages of simplicity and (sometimes) less initial cost than the other four categories. The disadvantages that appear to be driving this method toward extinction in this country are increasingly stringent environmental restrictions and the disappearance of convenient sites for landfills near the areas where most waste is generated. Also, there are significant problems posed by the production of harmful emissions from present and past (abandoned) landfills. Methods of minimizing such emissions are expensive and can create still other problems.

As commonly implemented, recycling followed by landfill has the same disadvantages as Class 1 except for diminution in volume of material landfilled and re-use of some materials, but usually has added cost because of the limited market for the recyclables and labor-intensive sorting and handling procedures.

Class 3 (incineration) has the advantage of further reduction in volume of residuals, but (a) poses real and perceived environmental problems in air emissions and potentially hazardous residual qualities, and (b) is wasteful in terms of the potential value of the materials destroyed by incineration (though not more wasteful than landfills).

Class 4 (conversion to non-energy uses) commonly offers further reduction in volume of residuals, has fewer real or perceived environmental disadvantages than simple incineration, and attempts to take some advantage of the potential value of the materials that would otherwise be destroyed or lost. The disadvantages of this class include: (a) from an economic standpoint, this is not the highest use of the organic portion of the waste; and (b) the low value of the product, combined with the slowness of the conversion

process and the aforementioned inefficiencies of the separation process, usually result in an uneconomic process that requires high tipping fees to be viable.

Even though Class 5 (recycling and cogeneration) offers the conceptual advantages of a large reduction in volume of residuals and highest (most economically attractive) use of the organic portion of the waste, the most recent evidence is that all waste-to-energy methods are losing ground. An article in the November 1993 Waste Age magazine notes that the number of resource recovery projects in conceptual or early planning stages decreased from 139 in 1988 to 27 in 1992, and at least 77 facilities have been "scratched" or canceled since 1991. Mandated improvements in ash disposal practices and air pollution control equipment have been so expensive that the tipping fees of newer resource recovery plants have risen to about \$70 per ton. The total number of waste-to-energy plants peaked in 1988 at 368, fell to 294 by 1991, and dropped to 248 by 1993. This illustrates the need and opportunity for our invention.

Our invention would fall into the Class 5 general category, but could minimize the critical ash disposal, pollution control, and volume problems by making a quantum jump forward in terms of:

- Air quality advantages resulting from the absorbent qualities of the oil shale additive and the combustion efficiency of the fluid bed process, i.e., acid gas emissions will be reduced to levels below EPA guidelines.
- Water quality advantages resulting from the cementitious properties of the oil shale ash (leachable materials will be encapsulated).
- Land use advantages resulting from virtual elimination of the residue entering the landfill. The small amounts that need to be deposited will be rendered non-hazardous.
- Economic advantages resulting from the fuel value of the oil shale additive and the combustion efficiency of the fluid bed process. For example, our calculations indicate that a typical installation would be economically viable at a tipping fee of \$30 per ton versus the almost \$70 per ton mentioned in the Waste Age article.

COMPETITIVE ENVIRONMENT

The size of the solid waste handling market and the serious nature of the problems faced by the industry ensure that the marketplace will be highly competitive for the foreseeable future. In terms of the broader national interest, this is a favorable factor. Given both market pressures and such governmental regulations as those requiring best available control technology, the demonstration of

a significantly improved method such as that embodied in our invention will spur both adoption of our concept and the search for truly competitive methodology.

Portions of the market are quite fragmented. Many landfills are owned by municipalities, but most are under private ownership. There are many small waste handling organizations and a few very large ones. There is a rather large number of suppliers of equipment. Industry publications and telephone books list numerous consultants. The Resource Recovery Yearbook published by Governmental Advisory Associates, Inc. (NY: telephone No. 212-410-4165) gives a good cross-section of vendors, designers, builders, and manufacturers of waste handling systems and equipment.

Though a competitive environment tends to lessen the market share for a given invention, such an environment also lessens the barriers posed by inertia and vested interests. We note that the potential market for our concept in the U.S. alone comprises over 300 facilities with an average cost of about \$70 million, or about \$21 billion over its life span. From an investment standpoint, it is necessary to capture only a small fraction of that market to justify the expenditures involved in a commercial-scale demonstration project. Given the advantages offered by our invention, its proper exploitation should result in a very large market share.

2.2 KEYS TO MARKET ENTRY

An invention will be successful if it satisfies a need in society and if the proper elements of society become convinced that:

- The process or concept is environmentally equal or superior to other options
- The process or concept is economically competitive with all other options
- The need for the process or concept is great enough to overcome the inertia associated with the status quo.

The environmental and economic advantages of the process are highlighted in the Technology Overview, and the need for the concept is evident from the statistics quoted herein. It has been reported that, with the new EPA landfill regulations, 34 states do not have acceptable landfill sites; previously approved sites are becoming more remote, and disposal costs are increasing. Thus, the need is already great and may become more critical within the next few years. Therefore, the key to market entry is the widespread demonstration of these advantages. To do this, we advocate a five-year program culminating in the operational testing of a facility that would demonstrate a unique, environmentally superior, economically attractive waste processing system. The program would

progress through five phases, as follows:

I: Preliminary system design and critical concept demonstration (the essence of this work has been performed under the grant program)

II: Detailed test and evaluation program

III. Design of an initial demonstration facility, site selection, and establishment of a joint venture

IV. Construction and start-up of demonstration facility

V. Operation of demonstration facility.

Assuming success in each phase, we will possess convincing evidence that implementation of our invention would be advantageous throughout the target regions. The demonstration and the creation of a joint venture with sufficient resources to accomplish the construction and operation of full-scale plants will give us the needed credibility and stability, and will enable us to do the advertising and public relations work needed to disseminate our findings throughout the various elements of society (government regulators, public officials, special interest groups, media representatives, etc.) whose approval of our concepts will ensure success.

BARRIERS

The potential barriers to widespread application of the process (and to most inventions in general) could be classified under the following headings:

- Simple inertia
- Vested interests
- Potential public mis-perceptions
- Unforeseen technological breakthroughs by competitors.

Inertia

Herein we use the term inertia simply to mean resistance to change on the part of the general public, as opposed to those (the "vested interests") who might perceive their own livelihood to be threatened by implementation of the invention. Inertia arises from a complex mixture of causes with both psychological and physical components. Inertia in the present case is represented by the attitude that "MSW is always handled by landfill." However, we believe that the public is becoming aware that this is no longer feasible near large metropolitan areas where most of the MSW is

generated. There will be great resistance in rural areas, but there is comparatively little municipal waste generated in such areas; we will implement the solution where the problem is greatest.

Vested Interests

An industry as large as that involved in municipal waste handling, dependent upon the goodwill or at least the acquiescence of governmental officials and regulators at every level, has a huge investment to protect and many powerful friends and advocates. However, the Government is now wrestling with the problem of finding and/or preserving landfill sites. Waste collection will not be affected by our process; therefore, we will not pose a threat to a large percentage of the current workforce.

Potential Public Mis-perceptions

Since the invention effectively addresses the outstanding environmental and economic concerns of the public, it is only necessary that we have sufficient resources to enable us to expend the effort needed to ensure that these advantages are understood by the public. From our work with synfuels and other industries, we recognize the importance of working with regulators and public interest groups from the outset of any program, and we are confident that our regulatory/permitting efforts will be as successful in this instance as they have been in other programs.

Unforeseen Technological Breakthroughs

Historically, technical breakthroughs are long in coming and longer in implementation. Successfully demonstrated technology always has an advantage over concepts that are still in the research stage. The five-step program outlined above is designed to culminate in successful demonstration of the technology; once this is accomplished we expect that the process will be able to accommodate and profit from other technological breakthroughs if and when they occur.

3.0 PROJECT OBJECTIVES

In brief, the project objectives were to:

- Demonstrate the complete combustion of oil shale/refuse derived fuel (RDF, the combustible portion of MSW) in a circulating fluidized bed combustor (CFBC)
- Demonstrate the ability of the oil shale to absorb the undesirable acid gas components (SO_2 and CL^-) created during the combustion of RDF
- Demonstrate the ability of the cementitious oil shale ash to encapsulate the "hazardous" components of the RDF ash
- Determine, to the extent possible, the potential scale-up problems associated with our concept.

We view the work performed under the grant as the first indispensable step in making a major contribution to the solution of the problems outlined in Section 2 of this report. We can restate this long-range goal as the creation of numerous facilities which use municipal solid waste (MSW) to generate electricity while recycling all usable components thereof and eliminating the main environmental problems associated with MSW and its associated waste streams. As noted in Section 2, we will have to demonstrate to a wide audience both the environmental and economic advantages of the concept. Each of the four objectives of the current work relates to this long-range goal. The following sections of this report describe how we conducted the investigation and the degree of success we had in reaching the limited objectives of the work performed under the grant.

4.0 METHODOLOGY

This phase of the project consisted of testing the technical viability of the process and determining workable operating conditions. We determined that a 6-inch CFBC could be used most efficiently to establish the ratio of oil shale to refuse-derived fuel (i.e., the absorbent-to-pollutant or A/P ratio), the temperature range for best absorption, and a workable residence time. This enabled us to scale up the tests to a 15-inch CFBC. Considering the limited number of tests we could conduct under the financial constraints of the grant, it was not possible to develop data on all design parameters (e.g., residence time, full temperature range, particle size effects). However, the number of tests was sufficient to demonstrate the technical feasibility of the process and define workable operating conditions, thereby enabling us to prepare a preliminary CFBC design adequate for economic estimation purposes.

Using the data produced during the CFBC tests, we were able to update economic projections and develop an improved understanding of the conditions governing economic viability. The details of the methodology employed in the study are outlined in the following subsections.

4.1 MSW/OIL SHALE TEST PLAN

4.1.1 Summary of Planned Tests

Our test plan (presented in detail in Appendix A) was designed to demonstrate that oil shale co-combusted with municipal solid waste (MSW) can reduce gaseous pollutants (SO_2 , CO) to acceptable levels (90%+ reduction) and produce a cementitious ash which will, at a minimum, be acceptable in normal landfills. The small-scale combustion testing was to be accomplished in a 6" circulating fluid bed combustor (CFBC) at Hazen Research Laboratories. This work was to be patterned after the study we conducted in 1988 when coal and oil shale were co-combusted in a program sponsored by the Electric Power Research Institute.

The specific purposes of the current test program were determined to be as follows:

1. Determine the required ratio of oil shale to MSW by determining the ratio of absorbent to pollutant (A/P),
2. Determine the effect of temperature and residence time in the reactor, and
3. Determine whether the chemical kinetic model previously developed for coal/oil shale mixtures is applicable to the present case.

The main features of the test plan were:

Equipment: Primary test unit a 6" circulating Fluid Bed Combustor with

monitors for off-gas for SO_2 , NO_x , Cl^- , O_2 , CO . Feeders for MSW and oil shale. Follow-up test(s) on larger-scale (15-inch) unit. One or more of the "best" scenarios identified in 6-inch unit tests to be replicated in the larger unit (scale-up factor of six). A plug-flow model would be used to predict the results of the confirmatory tests.

Tests: Conduct about 15 tests, having first obtained ultimate, proximate, heating value, sulfur content, and chlorine data on the MSW, and Ca, Mg, S, Cl, and grade data on the oil shale. Vary the A/P ratio to establish sensitivity of emissions to feed ratio of oil shale to RDF. Then, vary the temperature to establish an appropriate operating temperature range. Increase sulfur concentration to simulate maximum condition likely to be encountered in RDF. Increase Cl^- concentration to simulate maximum condition likely to be encountered in MSW. Pick preferred operating condition at the "best" A/P ratio for gas pollutant reduction. Produce at steady state sufficient ash for leachability tests. Repeat selected tests on the larger test unit.

4.1.2 Selection of Desired MSW Sample

Task 3, Attachment A to Grant Instrument DE-FG01-94CE15612, stated that "Municipal solid waste from industrial, residential, commercial and rural areas will be obtained to ensure that all components likely to be found in rural areas will be present in the test samples." That is, we wished to assemble test samples based on actual wastes from the above sources but specially mixed or synthesized to make possible a realistic evaluation of our concept's "ability to absorb all the undesirable acid gas components derived during the combustion of MSW" (Application for Federal Assistance, Section 2.0, item 2).

For MSW, we use the definition from Ref. 4.1-1, p. ES-2: "Municipal solid waste includes wastes such as durable goods, nondurable goods, containers and packaging, food scraps, yard trimmings, and miscellaneous inorganic wastes from residential, commercial, institutional and industrial sources ... MSW does not include wastes from other sources, such as construction and demolition wastes, municipal sludges, combustion ash, and industrial process wastes that might also be disposed of in municipal waste landfills or incinerators." For our purposes, MSW also excludes hospital wastes, not explicitly mentioned in the foregoing definition. The MSW to be tested differs from the general stream described above in that it does not include items that will be removed in steps prior to injection into a fluidized bed unit, i.e., we wish to test what Ref. 4.1-1 terms discards (refuse derived fuel or RDF), or MSW remaining after recovery for recycling.

Our representative sample therefore met, insofar as possible, the criteria listed in Tables 4.1-I (waste source characteristics) and 4.1-II (waste ultimate analysis). Where conflict might arise between source distribution and ultimate analysis, priority was to be given to ultimate analysis. Key components of the ultimate analysis (chlorine and sulfur) were to be synthesized as necessary to bring the representative waste into conformity with the ultimate analysis.

In Table 4.1-I we define two "representative" samples, one of which is identified as "severe" and the other as "typical." The "severe" sample is so designated because it has the highest concentration of acid-gas-producing materials that are considered likely to be encountered in ordinary practice, whereas the "typical" sample is averaged over all locations and seasons as determined from review of the literature (e.g., References 4.1-1 through 4.1-3). Note the large difference in yard trimmings between the "severe" and typical" samples.

Table 4.1-II shows the results of combining the "severe" waste composition with the highest reported sulfur and chlorine concentrations by type of waste. The two numbers at the bottom right of the table show that this case, designed to yield a maximum combined acid-gas test for our concept, would have a weight percent of .213 sulfur and .705 chlorine, or less than one percent total of the two acid gas contributors. Note also that the worst conceivable case for sulfur, based on available data, would be 1.5% sulfur if a batch were composed entirely of rubber and leather waste, whereas the worst case for chlorine would be 3.48% if a batch were composed entirely of textile waste. These numbers are not considered likely if large amounts of mixed MSW were to be processed, but they place a theoretical upper limit on the acid-gas problem facing our process.

Table 4.1-III shows the results of combining the "typical" waste composition with the highest reported sulfur and chlorine concentrations by type of waste. Since high-sulfur and high-chlorine concentrations come from different types of waste, the "typical" case shows a slightly higher sulfur concentration but a much lower chlorine concentration than does the "severe" case.

The third item to be considered is the highest-concentration sample of the two elements as found in the literature. For sulfur, the highest reported value is .70 wt percent (a case in Baltimore), and for chlorine it is an extremely high 1.51 wt percent (a case in San Diego). Considering the individual ultimate-analysis maxima shown in the table, it is difficult to visualize how these extreme values could arise.

TABLE 4.1-I
Waste Source Characterization

ITEM	Part A - Typical				Part B - Severe			
	WASTE STREAM (Nat. Avg. E6 Tons)	POST-SEP. (95% eff.)	TO FBC (lb)	TO FBC (%)	INPUT (Assum.)	POST-SEP. (95% eff.)	TO FBC (lb)	TO FBC %
Paper & Paperboard	73.30	73.30	1,763.08	44.08%	100.7	100.7	2,129.53	53.24%
Glass	13.20	0.66	15.87	0.40%	4	0.20	4.23	0.11%
Metals				0.00%				0.00%
Ferrous	12.30	0.62	14.79	0.37%	4	0.20	4.23	0.11%
Aluminum	2.70	0.14	3.25	0.08%	1	0.05	1.06	0.03%
Other Nonferrous	1.20	1.20	28.86	0.72%	1	1	21.15	0.53%
Plastics				0.00%				0.00%
Rigid	9.82	9.82	236.16	5.90%	20	20	422.94	10.57%
Film	6.38	6.38	153.50	3.84%	10	10	211.47	5.29%
Rubber & Leather	4.60	4.60	110.64	2.77%	2	2	42.29	1.06%
Textiles	5.60	5.60	134.70	3.37%	18	18	380.65	9.52%
Wood	12.30	12.30	295.85	7.40%	12	12	253.77	6.34%
Food Wastes	13.20	13.20	317.50	7.94%	5	5	105.74	2.64%
Yard Trimmings	35.00	35.00	841.85	21.05%	16	16	338.36	8.46%
Misc. Inorganic Wastes	2.90	0.29	6.98	0.17%	1	2.00	42.29	1.06%
Other	3.20	3.20	76.97	1.92%	2	2	42.29	1.06%
	195.70	166.30	4,000.00	100.00%	195.70	189.15	4,000.00	100.00%

TABLE 4.1-II
MSW Severe Case

ITEM	Severe Case			Ultimate Analysis			
	INPUT (Assumed)	POST-SEPAR. (95% efficiency)	INTO FBC (lb)	Sulfur (wt %)	Chlorine (wt %)	Sulfur (lb)	Chlorine (lb)
Paper & Paperboard	100.7	100.7	2,129.53	0.140	0.115	2.981	2.449
Glass	4	0.2	4.23	0.000	0.000	0.000	0.000
Metals			0.00			0.000	0.000
Ferrous	4	0.2	4.23	0.000	0.000	0.000	0.000
Aluminum	1	0.05	1.06	0.000	0.000	0.000	0.000
Other Nonferrous	1	1	21.15	0.000	0.000	0.000	0.000
Plastics			0.00			0.000	0.000
Rigid	20	20	422.94	0.600	2.250	2.538	9.516
Film	10	10	211.47	0.083	0.528	0.176	1.117
Rubber & Leather	2	2	42.29	1.500	0.000	0.634	0.000
Textiles	18	18	380.65	0.050	3.480	0.190	13.247
Wood	12	12	253.77	0.250	0.170	0.634	0.431
Food Wastes	5	5	105.74	0.150	0.320	0.159	0.338
Yard Trimmings	16	16	338.36	0.220	0.320	0.744	1.083
Misc. Inorganic Wastes	1	2	42.29	0.09	0.410	0.038	0.000
Other	2	2	42.29	0.09	0.410	0.410	0.000
TOTAL	196.7	189.15	4000.000			8.505	28.182
						0.213	0.705

TABLE 4.1-III
TARGET MSW WITH PREDICTED ULTIMATE ANALYSIS

ITEM	Typical Comp.			Ultimate Analysis/High Conc.			
	INPUT (Assumed)	POST-SEPAR. (95% efficiency)	INTO FBC (lb)	Sulfur (wt %)	Chlorine (wt %)	Sulfur (lb)	Chlorine (lb)
Paper & Paperboard	73.3	73.3	1,743.06	0.140	0.115	2.440	2.005
Glass	13.2	0.66	15.69	0.000	0.000	0.000	0.000
Metals			0.00			0.000	0.000
Ferrous	12.3	0.2	4.76	0.000	0.000	0.000	0.000
Aluminum	2.7	0.05	1.19	0.000	0.000	0.000	0.000
Other Nonferrous	1.2	1	23.78	0.000	0.000	0.000	0.000
Plastics			0.00			0.000	0.000
Rigid	9.82	9.82	233.52	0.600	2.250	1.401	5.254
Film	6.38	6.38	151.72	0.083	0.528	0.126	0.801
Rubber & Leather	4.6	4.6	109.39	1.500	0.000	1.641	0.000
Textiles	5.6	5.6	133.17	0.050	3.480	0.067	4.634
Wood	12.3	12.3	292.49	0.250	0.170	0.731	0.497
Food Wastes	13.2	13.2	313.89	0.150	0.320	0.471	1.004
Yard Trimmings	35	35	832.29	0.220	0.320	1.831	2.663
Misc. Inorganic Wastes	2.9	2.9	68.96	0.09	0.410	0.062	0.000
Other	3.2	3.2	76.10	0.09	0.410	0.410	0.000
TOTAL	195.7	168.21	4000.000			9.180	16.860
WT %						0.229	0.421

Based on the data analyzed, we planned to:

1. Use the "typical" composition identified in Tables 4.1-I and 4.1-III as a target starting place.
2. Measure the sulfur and chlorine components therein, and then synthesize the worst probable case (defined to be .5 wt percent sulfur and 1.1 percent chlorine as the rough average of the results obtained in Tables 4.1-II and 4.1-III and the highest recorded values) by addition of sulfur and chloride components as needed.

4.2 CONDUCT OF TESTS

4.2.1 Introduction

The tests were performed in conformity to the test plan (Subsection 4.1), with minor adjustments as desirable in view of preliminary test results. The tests are described in detail in Appendix B. The discussion in Appendix B is summarized and clarified as needed in this subsection.

Five days of combustion testing were performed using Hazen's 6-inch inside diameter (I.D) circulating fluidized bed combustor (CFBC) during February and March 1995. A scale-up test on Hazen's 15-inch I.D. CFBC was performed on March 23, 1995. The demonstration runs were coordinated and witnessed by Harry E. McCarthy and Reed L. Clayson, with the initial test runs also being witnessed by Professor Joann Lighty of ACERC.

Subsequent tests of the cementitious properties of the oil shale ash were conducted in late 1995 and early 1986. These tests consisted of stabilization tests and Toxicity Characteristic Leaching Procedure (TCLP) analyses on three ash samples. All tests were successful in terms of the objectives of the project.

4.2.2 Process Materials

The two process feeds for our test program were RDF and oil shale. A "typical" composition of both MSW and its derivative RDF, obtained from averaging data from many sources, appears in the left-hand columns of Table 4.1-I. A "severe case" composition was also derived from the literature (severe in the sense of having an acid gas precursor concentration thought not to be found in more than one case in 20) resulted in the target concentrations of sulfur and chlorine shown in the right-hand columns of Table 4.1-I.

The MSW sample used in the program was obtained from BFI of Eden Prairie. Since the BFI sample, which was pelletized, did not reach the concentrations of pollutants specified in our plan, cylinders of SO₂ and HCl were used to spike the process gas to reach the desired concentration of sulfur and chlorine (0.21% sulfur and 0.70% chlorine).

The oil shale used in the tests was lower in calcium and magnesium than that normally found in the Piceance Basin. This deficiency was offset

by the MSW sample, which contained significant levels of calcium and magnesium, determined to be residual binder from the pelletizing process. The data on the process materials are summarized in Table 4.2-I and detailed in Appendix B.

Table 4.2-I

Ultimate, Proximate, Heating Value, and Chlorine Analyses of
MSW and Oil Shale

	MSW per BFI		MSW per Hazen		Oil Shale per Hazen	
	As Received	Dry Basis	As Received	Dry Basis	As Received	Dry Basis
Ultimate, %						
Moisture	12.08	0.00	8.20	0.00	0.62	0.00
Carbon	14.43	16.41	41.90	45.64	12.14	12.22
Hydrogen	1.56	1.78	5.35	5.83	1.34	1.35
Nitrogen	0.11	0.12	0.43	0.47	0.29	0.29
Sulfur	0.06	0.07	0.17	0.19	1.35	1.36
Ash	17.68	20.11	9.86	10.74	75.28	75.75
Oxygen ^{1/}	54.08	61.51	34.09	37.13	8.98	9.03
Total	100.00	100.00	100.00	100.00	100.00	100.00
Proximate, %						
Moisture	12.08	0.00	8.20	0.00	0.62	0.00
Ash	17.68	20.11	9.86	10.74	75.28	75.75
Volatile	59.68	67.88	n/a	n/a	n/a	n/a
Fixed Carbon	10.56	12.01	n/a	n/a	n/a	n/a
Total	100.00	100.00	18.06	10.74	75.90	75.75
Heating Value						
Btu/lb	5660	6438	7445	8110	1630	1640
Chlorine, %			0.50	0.54	0.01	0.01

^{1/} Oxygen determined by difference

n/a Not assayed

4.2.3 Test Equipment

Hazen's externally heated, 6-inch I.D. CFBC test facility was used for the first series of tests. This facility, which has been used for a number of customers (including ourselves) on related work, is equipped with numerous instruments and recorders to permit control of the combustion operation and to obtain all needed operating information. The 6-inch test CFBC is described and illustrated in detail in Appendix B.

The second series of tests utilized Hazen's 15-inch I.D. CFBC. This system includes the combustor, an external cyclone and baghouse, and an exhaust gas scrubber. This system is also equipped with numerous instruments for combustion control and on-line reporting and recording of all needed data (Appendix B).

4.2.4 Testing Program

In all, 14 combustion tests were performed, with nine tests involving the 6-inch facility and five being performed on the 15-inch unit. In these tests, molar absorbent/pollutant ratios were varied from 1.77:1 up to 5.0:1, and average fluid bed operating temperatures were varied from 1438 to 1519°F. All tests were carefully monitored.

4.3 TECHNICAL ANALYSIS

4.3.1 Introduction

In our test program, we were primarily interested in the ratio of absorbent (the total calcium and magnesium present in the oil shale and RDF expressed in moles per hour) to pollutant (total sulfur and chlorine expressed in consistent units) needed to reduce the acid gases by a given percentage as a function of run parameters (primarily temperature in the CFBC). We found in the literature considerable data indicating that ordinary CFBC processes using state of the art techniques (using crushed limestone as an additive) commonly reduce sulfur emissions by 90%, but little data indicating that reductions of much more than 90% are obtained with limestone. However, the most recent source available to us specifically relating to MSW indicates that "Limestone can be added to the bed of FB units to achieve a nominal reduction in sulfur dioxide (40-50% reduction) and some HCl reduction" (Ref. 4.3-1, 1995, p. 309).

Our belief was that, if the oil shale could perform at least as well as limestone as an additive, the other advantages of the oil shale (its own heating value as an additive versus the energy penalty associated with limestone, plus the cementitious nature of the oil shale ash and its attendant service in processing of the solid residue) would be sufficient to establish our plentiful oil shale resource as the additive of choice for fluid bed operations with MSW.

The results of the CFBC tests bore out our expectations, both when

considering absorbent/pollutant (A/P) ratios as a whole and in terms of absorption of both sulfur and chlorine. However, the results from the 6-inch and 15-inch CFBC tests were not as consistent as expected, especially in the low A/P ratio range. This effect is noticeable in the top portion of Figure 4.3-1. Analysis of data indicates that two factors are probably responsible for the inconsistency. First, data from the 15-inch tests indicates that the reduced residence time and/or reduced temperature in the freeboard resulted in reduced acid gas absorption. Second, the 15-inch test data without oil shale shows an excess sulfur output compared with the input; the excess probably came from contamination in the 15-inch system. If we compute the effects of contamination, and if we assume that the contamination is constant, we note that the effect of the excess output moves the 15-inch unit performance closer to that of the 6-inch unit, especially at high A/P ratios. This is shown in Table 4.3-I and Figure 4.3-1. However, such inconsistency (further discussed in Subsection 4.3.3 below) should not obscure the central finding of the tests, which is, for both test units, that:

- Oil shale as an additive to an RDF stream in small CFBC test units is capable of reducing pollutant levels by extremely high percentages
- Injection of oil shale with RDF has other notable advantages, including making it significantly easier to control the fluid bed operating conditions.

Physical test work is expensive and time-consuming, and funds under the grant were limited. Therefore, it was necessary to (a) establish realistic performance goals which would guide the test work, and (b) use modelling and analysis to generalize upon the test results insofar as possible. As described in the following subsections, we based our performance goals upon the environmental requirements which apply to this type of process, and our modelling and analysis upon combined practical data and theoretical considerations.

4.3.2 Performance Standards

Informal communications with EPA Region VIII specialists and documents referred to by them have led us to conclude that implementation of our invention would involve satisfaction of environmental requirements embodied in New Source Performance Standards for New Stationary Sources: Municipal Waste Combustors (40 CFR Part 60), and Standards of Performance for New Stationary Sources and Guidelines for Control of Existing Sources: Municipal Solid Waste Landfills (40 CFR Parts 51, 52, and 60).

Figure 4.3-1
Performance Summary

Note: With contaminant added in

Table 4.3-I
SUMMARY OF POLLUTANT REMOVAL TEST RESULTS*

CFBC	RUN NO.	A/P RATIO	% SO2 REDUCTION	% ACID GAS REDUCTION
6"	1	1.70	80.2	92.0
6 "	4	2.89	89.8	94.5
6 "	3	3.61	93.8	96.1
6"	2	4.48	98.2	98.6
15"	1	1.65	-38.8	43.8
15"	2	3.37	59.3	74.8
15"	3	3.87	69.0	79.0
15"	4	4.29	86.0	90.0
15"	5	4.57	90.5	92.9

*See table below for corrections to 15" tests assuming amounts produced by prior contamination (residue) of pollutant

**TEST RESULT MODIFICATIONS BASED ON ASSUMPTION OF PRIOR
CONTAMINATION OF 15" TEST UNIT**

CFBC	RUN NO.	A/P RATIO	% SO2 REDUCTION	% ACID GAS REDUCTION
15"	1	1.10	38.5	62.6
15"	2	2.49	74.3	81.3
15"	3	2.99	78.5	83.8
15"	4	3.42	89.7	92.1
15"	5	3.72	92.7	94.2

We learn the following from the regulatory sources:

- SO₂ emissions should not exceed 30 ppmv or 80-percent reduction, whichever is less stringent (40 CFR Part 60).
- HCl emissions should not exceed 25 ppmv or 95-percent reduction, whichever is less stringent (40 CFR Part 60).
- Landfill leaching is listed as the primary potential source of water pollution from an uncontrolled landfill. We therefore infer that, if the solid residue from implementation of our invention is placed in a landfill, its ability to pass standard leachability tests is the criterion on which it should be judged.
- MSW landfill emissions consist primarily of methane and CO₂, with trace amounts of more than 100 different NMOC's (nonmethane organic compounds). Emission regulations relate to on-site landfill treatments to reduce total emissions from landfills. Therefore, the degree to which implementation of our invention reduces the amount of MSW residue placed in a landfill is a primary criterion consistent with official government pollution reduction priorities (Pollution Prevention Act of 1990).

Work reported herein has provided plentiful evidence that (a) compliance with the SO₂ emissions limit is readily attainable (as indicated in Table 4.3-I, control in the high 90 percent range is attained in the test CFBCs), (b) compliance with the HCl emissions criteria is readily attainable (emissions were virtually eliminated - consistently below the threshold (5 ppmv) of the monitoring equipment), and (c) the ash as tested passed the leachability and TCLP tests (Table 4.3-II) and should therefore minimize landfill leaching to allow safe ash disposal at normal landfills.

4.3.3 Data Analysis

Methodology. Our data analysis was based upon Hazen's test work, but was carried out independently from that performed by Hazen. For example, our analysis used 379 ft³/lb mole for standard conditions. We assumed that the SO₂ and HCl flows were measured at standard conditions. Since the temperature and pressure of these gases were not measured, these assumptions may differ slightly from those employed by Hazen. However, these small differences do not alter the overall analysis nor the conclusions reached. Herein we first describe the analyses we performed and then present our results. These results can be compared with those reported by Hazen (Appendix B).

Table 4.3-II
ASH STABILIZATION AND TCLP TESTING RESULTS

A. Fluid-bed Ash Stabilization Testing Formulations

Sample No.	Fluid-bed Ash, g		Reagent Additives, g		Water Added
	Bed	Baghouse	Type II Cement	Blast Furnace Slag	
1	9	141	50	0	87
2	8.4	131.6	6	54	78
3	22.5	127.5	0	0	~80

B. Toxicity Characteristic Leaching Procedure (TCLP) Leachate Analysis Results

Criterion No.	RCRA Metal	Sample 1, mg/l	Sample 2, mg/l	Sample 3, mg/l	EPA Regulatory Limit, mg/l
1	Arsenic	<0.01	<0.01	<0.03	5
2	Barium	0.2	0.2	<0.5	100
3	Cadmium	0.03	0.04	0.03	1
4	Chromium	0.49	0.08	0.7	5
5	Lead	0.4	0.4	<0.5	5
6	Mercury	<0.0001	<0.0001	<0.001	0.2
7	Selenium	<0.01	<0.01	<0.03	1
8	Silver	0.03	0.03	0.05	5

We took the air flow and feed rates of oil shale and RDF from the Hazen data. We then computed the pound-moles/hour (lb moles/hr) of absorbents for each test run. These absorbents were defined as the CaO and MgO contained in the RDF and the oil shale. We then computed the lb moles/hr of SO₂ and Cl⁻ from the three sources (RDF, oil shale, and HCl and SO₂ gases). The total absorbent was calculated. The pollutant totals were computed by adding SO₂ and Cl^{1/2}. The reason for dividing Cl⁻ by two is that two moles of Cl⁻ would be absorbed by each mole of absorbent. The A/P ratio was then computed by dividing the lb moles/hr of absorbent by the lb moles/hr of pollutant.

In computing the A/P ratios for the 15-inch unit tests, it was noted that, in a test made without oil shale, more pollutant SO₂ was exiting from the 15-inch unit than was being input (15-inch test No. 1). In fact, 38% more output was measured than input, even though both CaO and MgO were present in the RDF feed. (This absorbent was probably due to the binder used to make the RDF pellets.) See Appendix B for a more complete description.

The reasons for the excess SO₂ could have been:

- (1) More sulfur was present in the RDF in the 15-inch tests than in the 6-inch tests,
- (2) The test instruments were reading incorrect SO₂ output, or
- (3) The 15-inch system contained some sulfur contamination from earlier tests.

Potential reasons (1) and (2) were checked out and eliminated as possibilities, leaving reason (3) as the most logical possibility.

Using these data, we assumed that the RDF performed the same in the 15-inch as in the 6-inch test; thus, a contamination output of 3.462×10^{-3} moles/hour of SO₂ was computed. If we then assume that this contamination output was constant, we can compute its effect. This effect is shown in the lower part of Table 4.3-I.

Using these data to account for the contamination, we can compare the results of the 6-inch and 15-inch tests. Figure 4.3-1 shows a plot of A/P ratio versus percent removal of SO₂ and total acid gas. At low A/P ratios, we find 40% to 60% reduction in removal for the 15-inch unit. At higher (3.5) A/P ratios, we find only 4% to 5% reduction in absorption for the 15-inch tests.

Referring again to the Hazen tests (Appendix B), we note that the freeboard temperature in the 15-inch unit was 200°+ F below the desired temperature. If we assume that the quenching effect for the absorption of the acid gas component is of the same magnitude as the effect at higher temperatures, we find that the residence time for the 16-inch unit is only about 30% of the residence time of the 6-inch unit. Thus, we can explain the data differences due to reduced time for absorption. If this is true, we then note that the effect is much more pronounced at

the lower A/P ratios (Figure 4.3-1). We can in effect override the effect of reduced performance in the 15-inch unit by increasing the A/P ratio. Although this is not the best way to correct this problem, we note that residence time in commercial units is about seven seconds versus the two seconds used in the 6-inch unit. The longer residence time will be a built-in factor on the comparatively large commercial units, thereby presumably minimizing any absorption problems when moving to the larger units.

System Performance. From the above, we can conclude that the process is technically viable, and that the oil shale with RDF can produce energy in an environmentally acceptable manner. In addition, the data obtained on leaching shows that the ash can be disposed of in an environmentally benign manner.

We can also conclude that an A/P ratio of 3.5 is sufficient to remove 96% or more of the acid gas pollutants. This ratio is equivalent to adding 15.8 pounds of oil shale per 100 pounds of normal (as received) RDF. For the "severe" case, we compute that 21.04 pounds of oil shale per 100 pounds of RDF would be required to achieve the same result. These amounts would be sufficient to absorb the acid gas. The minimum amount necessary to make the ash environmentally benign will be determined by further tests. We will update the economic analyses using the computed oil shale quantities required.

Temperature Considerations. System performance appeared to be more dependent on operating temperature than we had noted in similar tests on coal/oil shale mixtures. Therefore, we analyzed this effect as outlined in this subsection.

Tests 1 through 5 of the series of runs performed on 6 February 1995 showed a strong relationship between acid gas absorption and operating temperature. We have re-computed the A/P ratios to be consistent with our other data. We have also computed the percent removal of the SO₂ and total acid gases (Table 4.3-III).

Table 4.3-III Summary of 6-Inch Unit Test Data 2/6/95				
Test No.	A/P	Percent Removal		Temperature (deg. F)
		SO ₂	Acid Gas	
1	4.45	98.5	98.7	1438
2	3.47	95.7	97.2	1449
3	3.61	84.4	90.2	1465
4	4.39	83.5	86.2	1504
5	4.98	60.4	70.1	1578

To improve our understanding of these data, we plotted acid gas removal versus temperature (Figure 4.3-2). These data show that the temperature effect is greater than the effect of increasing the A/P ratio. In studying this effect, limitations in the number of tests performed have

made it necessary to take certain liberties with the data. Note that the A/P ratios in Tests 2 and 3 are nearly equal, as are the A/P ratios of Tests 1 and 4. Using these two sets of points to define straight lines (Figure 4.3-2), we find that the temperature effect is much more pronounced at an A/P ratio of about 3.5 than it is at an A/P ratio of about 4.5. Also, note that at an A/P ratio of 4.98 the data point is below the 4.5 curve, indicating that the temperature effect is even more pronounced at still higher temperatures. This indicates a faster-than-linear fall-off in absorption as the operating temperature increases. Since this effect was not as noticeable in our prior coal/oil shale tests (Ref. 4.3-2), one must speculate on possible explanations.

The two most obvious possibilities for the unexpectedly pronounced dependence of absorption on temperature are (a) chemical reactions and (b) ash fusion of the oil shale. These possibilities are discussed below.

(a) Chemical reaction. The absorption reactions must be at equilibrium, i.e., $\text{CaO} + \text{SO}_2 \rightleftharpoons \text{CaSO}_3$. If the equilibrium shifts with temperature, we would expect a rather smooth temperature effect which could be shifted by adding more absorbent. This explanation fits the data on Figure 4.3-2.

(b) Ash fusion of oil shale. Although data on the ash fusion temperature of the oil shale used in our tests was not measured, typical Colorado oil shale fuses at between 1700 and 2000° F (927 to 1093° C). Some oil shales may have lower fusion temperatures; however, we have not found data indicating fusion at temperatures as low as 1600° F. Also, if fusion were occurring, we would expect to see a catastrophic change when fusion temperature was reached (i.e., an increase of only a few degrees would have a severe impact on absorption). Also, we would expect to find that changing the A/P ratio would have little impact on performance. Thus, our test data do not support the fusion hypothesis; however, in the next test series we will measure ash fusion temperature.

 * The data analyses performed by ACERC personnel show that the equilibrium shift with the decomposition of CaSO_3 to CaO and SO_2 will occur at temperatures above 800°C (1562°F). We noted that this shift occurs at temperatures somewhat lower than that. This could be explained by the presence of metals in the ash catalyzing the decomposition reaction. We did not note the same dramatic temperature effect in our studies with coal/oil shale mixtures; thus, it is likely that any catalyzing effect would be characteristic of the MSW. The variability of MSW would then lead one to expect that not all MSW would behave exactly the same in this respect; however, based upon all data available to date, we would not expect any stronger effect than that noted in the data presented herein. Again, as noted above, the temperature effect can be offset to some extent by increasing the A/P ratio.

Figure 4.3-2
Effect of Temperature on Absorption

See Table 4.3-III for data on (Runs) 1,2,3,4,5

4.3.4 Synthesis

In earlier testing of a similar nature, we found that the CFBC was operating over the range of conditions tested in plug-flow mode (i.e., with no back mixing of gas within the reactor). A chemical reaction model based on plug flow assumes that: (a) the flow of fluid through the reactor is orderly, with no element of fluid overtaking any other element; (b) the two phases are a pseudo continuum, behaving as a homogeneous substance for computational purposes; (c) temperature in the combustor is constant; and (d) the reaction rate for a given input is entirely a function of temperature.

Integration of the applicable reaction equations leads to the finding that the relationship between two quantities (one directly related to the absorbent/pollutant molar input ratio and the other related to both that ratio and the pollutant input/output concentration ratio) should plot as a straight line on semilog paper. The equation so derived is:

$$(1 + R) K_1 Y = \ln \frac{R - \Delta_o}{R(1 - \Delta_o)}$$

where R = input absorbent/input pollutant,
 Y = dimensionless residence time,
 Δ_o = output pollutant/input pollutant, and
 K_1 = reaction rate constant times pollutant concentration at point when residence time goes to zero.

If this relationship does indeed hold over all tests, one is justified in concluding that:

- (1) The model's assumptions are valid for the conditions tested, so that one can estimate the performance of the system for absorbent/pollutant conditions not tested as long as the other conditions are held constant, and
- (2) The tests have been conducted and reported correctly, since the probability of precisely compensating errors must become vanishingly small over a large number of trials.

For the work performed under the current grant, we find reasonable agreement with the hypothesis that the CFBC was in plug-flow mode. Figures 4.3-3 and 4.3-4 illustrate the relationships for the 6-inch and 15-inch tests, respectively. The goodness of fit with the plug-flow hypothesis is indicated by the correlation coefficients between the data sets (-.927 for the 6-inch tests and -.941 for the 16-inch tests, where a value of -1.00 would indicate the best fit obtainable). This is a reasonably good data fit.

6-Inch Tests

Figure 4.3-4

15-inch Tests

The similarity in the slopes of the regression lines for the 6-inch and 15-inch unit runs led to an additional step in the analysis. We first multiplied the $(1+R)$ values from the 15-inch unit tests by a factor (0.75) which approximates the effective difference in residence times between the 15-inch test unit and the 6-inch test unit. We then combined the 6-inch and 15-inch test data and obtained a good approximation of a straight line on log paper (with a correlation value of .969) for the combined data (Figure 4.3-5). This further supports the assumption that we can use the plug-flow model to estimate the performance of the system under A/P ratios and some operating conditions (e.g., residence times) that were not tested under this contract. However, more data are required to enable us to develop a full picture of the model fit and its implications.

Figure 4.3-5

4.4 ECONOMIC ANALYSIS

We used the approach of relying primarily upon the discounted cash flow return on investment of a representative project as the best means of assessing the economic viability of the concept.

4.4.1 Methods and Objectives

The purposes of this discussion are to establish the framework for the economic and financial evaluation of a representative project based on our invention, to explain the evaluation methods used, and to summarize the results of the evaluation. The analysis assumes successful prior completion of Phase I (Subsection 4.2), and covers the post-Phase I work involved in designing, constructing, and operating the first facility based on the subject invention. In developing this preliminary economic analysis, we used engineering, taxation, and cost estimating factors which apply to the western United States.

The general economic analysis methodology consists of three basic parts: inputs, modeling, and outputs. In turn, inputs are divided into four parts: engineering estimates, economic assumptions, financial assumptions, and sensitivity parameters. Engineering estimates include the project schedule, yearly capital costs, operating costs, working capital requirements, and production rates. Economic assumptions include product selling price, inflation rates or escalation factors, taxation rates, and treatment of tax benefits and credits. Financing assumptions include the fractions of the project that are funded by debt and equity interests, interest rates on borrowed funds, payment provisions, and general loan structure. Sensitivity parameters provide the framework for evaluation of the effects of uncertainties in previously described estimates and assumptions.

The above-described inputs are incorporated into the economic model to generate yearly project income statements for the life of the project. This method consists of taking the gross revenues; subtracting direct operating costs, royalties, severance taxes (where appropriate), property taxes, and interest payments to obtain net income; subtracting depreciation to get taxable income; subtracting federal and state income taxes less tax credits to find the net cash flow after taxes; and adding depreciation back in to obtain operating cash flow. From operating cash flow, subtracting capital expenditures, working capital, bonus bid payments (where appropriate), any debt principal paid, and adding back any borrowed funds yields the net cash flow for each year. Subsequent analysis of the project's viability is based on net cash flow.

4.4.2 Evaluation Criteria

The final output of any economic analysis is the measure of profitability and risk. Four approaches to measuring a project's viability are used in this analysis. Project risk is assessed using parametric sensitivity analysis to measure the impact of departures from our estimates and assumptions.

The four analysis measures used are:

1. Discounted cash flow return on investment (DCFROI) - the discount rate at which the sum of the net cash flows is zero.

DCFROI = i , where

$$\sum_{n=0}^N \frac{C_n}{(1+i)^n} = 0$$

C_n = net cash flow in year n
 i = discount rate
 and N = project life

2. Net Present Value (NPV) - the present worth of all future cash outlays and returns at a specified discount rate.

3. Cash Flow Analysis - an undiscounted evaluation showing investment, maximum equity exposure, average cash flows, cumulative cash flow, minimum yearly cash flow, and principal and interest payments. It provides a measure (a quantitative indication) of project profitability to equity participants and a measure of risk to prospective lenders.

4. Payback Period - the time required for equity participants to recoup their after-tax expenditures.

4.4.3 Economic Inputs and Assumptions

This subsection lists the baseline cost, revenue, inflation, financial, tax, scheduling, production, and modeling inputs and assumptions.

a. Engineering Estimates

For the baseline case, expected increases in prices and costs are not included. Some of the income and expenses for this case are in today's dollars, but others are in dollars for the latest available year. The project development schedule for a representative facility and the engineering cost factors used in the analysis are based on work we have performed for energy companies in the United States.

The economic analysis is restricted to a 24-year period of development, construction, and operation, even though production could continue for some years beyond that point. Because operating costs are low in comparison to revenues, extending the analysis beyond year 24 would increase the expected return on investment. However, for purposes of economic analysis the project is conservatively considered to cease operations at the end of year 24 with no salvage value.

The economic analysis (1) relates to a single plant which would satisfy the waste disposal needs of a population of 250,000 to 300,000, so that a number of such plants would be needed for any major metropolitan area; and (2) excludes preliminary test work, whose costs should be pro-rated or apportioned over all of the plants to be constructed.

b. Royalties

No royalties are anticipated.

c. Property Tax

Property tax (or ad valorem tax) estimates are based on the structure of a western U. S. state. The tax is computed on the basis of fair market value of all equipment, structures, and property improvements. Yearly property tax is collected in arrears (i.e., this year's tax is based on last year's assessment), and is computed as follows:

$$\begin{aligned} \text{Property tax} &= \text{Mill levy} \times \text{assessment rate} \times \text{assessment} \\ \text{where mill levy} &= 64.38 \text{ mills } (0.06438), \text{ and} \\ \text{assessment rate} &= 20 \text{ percent } (0.20). \end{aligned}$$

d. Depreciation

Depreciation is computed on an asset-specific basis as a function of the asset tax life and the year the asset was placed into service in accordance with the depreciation schedules of the 1986 U. S. Tax Reform Act.

e. State Income Tax

The state income tax is computed at the rate of 5 percent of net after depreciation.

f. Federal Income Tax

Federal income tax is computed as 27 percent of pre-tax net profit minus state income tax.

g. Tax Credits

No special (e.g., energy) tax credit is projected to be available. A tax loss carry-forward is computed for the first years of project operation.

h. Treatment of Tax Benefits during Development

In our analysis, it has been assumed that the project's equity sponsors will have sufficient other income to absorb all tax credits and operating losses (in the form of negative income taxes) in the year they are shown during the construction and start up period. Therefore, no losses forward are shown in the income statements. This also serves to reduce the real after-tax investment of the equity participants.

4.4.4 Financial Assumptions

a. Project Financing

In the sensitivity analyses (subsection 4.4.6), one financial assumption was that the sponsors would carry all project costs (i.e., there would be no borrowing, so the debt-equity ratio would be zero). This is the most severe test from a financial standpoint. However, for the base case covered in this subsection, a more typical arrangement of 70-percent borrowing ($DER = .700$) at an annual interest rate of 10 percent is assumed.

b. Income Stream Factors

The representative project is assumed to derive revenue from three main sources: (1) sale of electricity (export power), (2) tipping fees from all MSW received, and (3) sale of recyclable materials (aluminum, other metals, glass). Prices for these items are extremely variable, either geographically, seasonally, or by category of operating agency. We base our factors on the northeastern and far western metropolitan areas where appropriate oil shale resources are available, and upon current experience as reported in the literature.

Sale of Electricity. The key question here is whether the plant would have to sell electricity at the so-called "avoided-cost" rates under PURPA (where the results of reported negotiations have varied from 3 to 7 cents per kilowatt-hour), or whether the plant would function as, say, part of a municipality's public utility structure, in which case the current sales price averages 6.97 cents per kw-h (Ref. 4.4-1). Given current pressures to modify or abolish the PURPA legislation, there is a high degree of uncertainty regarding this factor. We note one recent waste-to-energy case in which the sales price was reported to be 11 cents per kw-hr. Herein we use a conservative estimate of the amount of electric power produced by the process in our representative case, and we use 6.9 cents per kw-hr as our base case.

Tipping Fees. In the regions of interest, the current tipping (disposal) fees at privately owned or operated landfills average \$73.17 per ton in the Northeast and \$37.69 in the West (Ref. 4.4-2). We used a value of \$30 per ton as the base case, and applied the more typical current factors in our sensitivity analyses (Subsection 4.4.6).

Sale of Recyclables. It was difficult to find a large enough sample of prices to derive averages for the highly variable prices being paid for aluminum, other metals, and glass which would be recovered at the representative facility. We used prices of 20 cents per pound for aluminum, 10 cents per pound for other metals, and 2 cents per pound for glass.

4.4.5 Expected Returns

The objectives of the economic evaluation of a process are to determine the ability of a representative project based on the process to support

all credit obligations, plus earn attractive returns for the equity investors. A major concern of both equity participants and lenders for a "first generation" project of this type is the ability of the project to meet its debt and equity obligations under adverse deviations from current and assumed economic, technical, and operating conditions. We have addressed these concerns using parametric sensitivity analysis. Factors examined for deviation from our conservative base case assumptions include:

- Increases in Capital Costs
- Increases in Operating Costs
- Changes in Electricity (Sales) Prices
- Different Tipping (Waste Disposal) Fees
- 100% Project Financing

The results of the economic evaluation criteria using the baseline assumptions are shown in Table 4.4-I. These findings are summarized and clarified in the subsequent discussion.

- After-tax DCFROI: 22.65% on equity invested
- After-tax Net Present Value:
 - at 0% = \$132.9 million
 - at 10% = \$27.6 million
 - at 20% = \$3.0 million
 - at 30% = \$ -5.3 million
- Cash Flow Summary:
 - a. Capital Investment (including project engineering and working capital) = \$68.5 million
 - b. Loan Amount = \$52.4 million (@ DER=0.7)
 - c. Maximum After Tax Equity Exposure = \$26.7 million
 - d. Average Yearly Net Cash Flow during full production = \$7.1 million
 - e. Minimum Yearly Net Cash Flow during full production = \$3.7 million
 - f. Annual operating cost during full production = \$3.7 million
- Years to payout: 6 (4 years of full production)

Table 4.4-I (page 1 of 5)

BASE CASE CASH FLOW

(UNITS IN THOUSANDS)

Major Production of	4080 Mw	/day	for \$	6.90	Debt/Equity Ratio: 0.7000	
Project Year	1	2	3	4	5	
PRODUCTION						
Exp Pwr Mw	0	0	803	1428	1428	
Tip Fee Tons	0	0	89	158	158	
Recycle Lbs	0	0	277	541	541	
REVENUES						
Exp Pwr Mw	0	0	5543	9853	9853	
Tip Fee Tons	0	0	2658	4725	4725	
Recycle Lbs	0	0	1870	3658	3658	
GROSS REVENUE	0	0	10071	18236	18236	
ROYALTY	0	0	0	0	0	
OPERATING COSTS						
OPERATIONS	3624	4693	1954	2286	2286	
MAINTENANCE	0	0	791	1407	1407	
TOTAL OPERATING COSTS	3624	4693	2745	3693	3693	
INSURANCE	0	0	67	67	67	
SEVERANCE TAX	0	0	0	0	0	
INTEREST	0	0	0	5236	4800	
PROPERTY TAXES	0	0	0	1150	729	
NET AFTER COSTS	-3624	-4693	7259	8089	8947	
DEPRECIATION	0	0	9875	13683	13010	
NET AFTER DEPRECIATION	-3624	-4693	-2616	-5594	-4063	
DEPLETION	0	0	0	0	0	
PRE-TAX PROFIT	-3624	-4693	-2616	-5594	-4063	
STATE INCOME TAX	-181	-235	-87	-186	-135	
NET TAXABLE INCOME (FED)	-3443	-4458	-2529	-5407	-3927	
FEDERAL INCOME TAX	-930	-1204	-683	-1460	-1060	
TAX CREDITS	0	0	4712	0	0	
NET AFTER TAXES	-2513	-3255	2866	-3947	-2867	
PLUS: DEPRECIATION	0	0	9875	13683	13010	
DEPLETION	0	0	0	0	0	
OPERATING CASH FLOW	-2513	-3255	12741	9736	10143	
CAPITAL EXPENDITURES	19541	47719	0	0	0	
CAPITALIZED INTEREST	0	1452	4894	0	0	
BONUS BID PAYMENTS	0	0	0	0	0	
BORROWED FUNDS	14519	34420	3426	0	0	
DEBT PRINCIPAL	0	0	0	4364	4364	
WORKING CAPITAL	1200	0	0	0	0	
NET CASH FLOW	-8736	-18006	11273	5372	5779	
SUNK COSTS	0	0	0	0	0	
COVERAGE RATIOS						
INTEREST	0.00	0.00	0.00	-0.07	0.15	
DEBT SERVICE	0.00	-0.22	0.26	0.19	0.21	
CUMULATIVE CASH FLOW	-8736	-26741	-15469	-10097	-4317	

Table 4.4-I (page 2 of 5)

BASE CASE CASH FLOW

(UNITS IN THOUSANDS)

Major Production of 4080 Mw /day for \$ 6.90 Debt/Eqty Ratio: 0.7000					
Project Year	6	7	8	9	10
PRODUCTION					
Exp Pwr Mw	1428	1428	1428	1428	1428
Tip Fee Tons	158	158	158	158	158
Recycle Lbs	541	541	541	541	541
REVENUES					
Exp Pwr Mw	9853	9853	9853	9853	9853
Tip Fee Tons	4725	4725	4725	4725	4725
Recycle Lbs	3658	3658	3658	3658	3658
GROSS REVENUE	18236	18236	18236	18236	18236
ROYALTY	0	0	0	0	0
OPERATING COSTS					
OPERATIONS	2286	2286	2286	2286	2286
MAINTENANCE	1407	1407	1407	1407	1407
TOTAL OPERATING COSTS	3693	3693	3693	3693	3693
INSURANCE	67	67	67	67	67
SEVERANCE TAX	0	0	0	0	0
INTEREST	4364	3927	3491	3055	2618
PROPERTY TAXES	591	454	332	277	591
NET AFTER COSTS	9521	10095	10653	11144	11267
DEPRECIATION	11437	11241	1807	1807	1807
NET AFTER DEPRECIATION	-1916	-1146	8846	9337	9460
DEPLETION	0	0	0	0	0
PRE-TAX PROFIT	-1916	-1146	8846	9337	9460
STATE INCOME TAX	-64	-38	295	311	315
NET TAXABLE INCOME (FED)	-1852	-1108	8551	9026	9145
FEDERAL INCOME TAX	-500	-299	2309	2437	2469
TAX CREDITS	0	0	0	0	0
NET AFTER TAXES	-1352	-809	6242	6589	6676
PLUS: DEPRECIATION	11437	11241	1807	1807	1807
DEPLETION	0	0	0	0	0
OPERATING CASH FLOW	10085	10432	8049	8396	8483
CAPITAL EXPENDITURES	0	0	0	0	0
CAPITALIZED INTEREST	0	0	0	0	0
BONUS BID PAYMENTS	0	0	0	0	0
BORROWED FUNDS	0	0	0	0	0
DEBT PRINCIPAL	4364	4364	4364	4364	4364
WORKING CAPITAL	0	0	0	0	0
NET CASH FLOW	5721	6069	3686	4032	4119
SUNK COSTS	0	0	0	0	0
COVERAGE RATIOS					
INTEREST	0.56	0.71	3.53	4.06	4.61
DEBT SERVICE	0.23	0.27	0.23	0.27	0.32
CUMULATIVE CASH FLOW	1404	7473	11158	15190	19309

Table 4.4-I (page 3 of 5)

BASE CASE CASH FLOW

(UNITS IN THOUSANDS)

Major Production of	4080 Mw	/day	for \$	6.90	Debt/Eqty Ratio: 0.7000
Project Year	11	12	13	14	15
PRODUCTION					
Exp Pwr Mw	1428	1428	1428	1428	1428
Tip Fee Tons	158	158	158	158	158
Recycle Lbs	541	541	541	541	541
REVENUES					
Exp Pwr Mw	9853	9853	9853	9853	9853
Tip Fee Tons	4725	4725	4725	4725	4725
Recycle Lbs	3658	3658	3658	3658	3658
GROSS REVENUE	18236	18236	18236	18236	18236
ROYALTY	0	0	0	0	0
OPERATING COSTS					
OPERATIONS	2286	2286	2286	2286	2286
MAINTENANCE	1407	1407	1407	1407	1407
TOTAL OPERATING COSTS	3693	3693	3693	3693	3693
INSURANCE	67	67	67	67	67
SEVERANCE TAX	0	0	0	0	0
INTEREST	2182	1745	1309	873	436
PROPERTY TAXES	568	559	550	541	561
NET AFTER COSTS	11726	12171	12616	13062	13479
DEPRECIATION	1807	1612	1612	977	977
NET AFTER DEPRECIATION	9919	10559	11005	12085	12501
DEPLETION	0	0	0	0	0
PRE-TAX PROFIT	9919	10559	11005	12085	12501
STATE INCOME TAX	331	352	367	403	417
NET TAXABLE INCOME (FED)	9589	10207	10638	11682	12085
FEDERAL INCOME TAX	2589	2756	2872	3154	3263
TAX CREDITS	0	0	0	0	0
NET AFTER TAXES	7000	7451	7766	8528	8822
PLUS: DEPRECIATION	1807	1612	1612	977	977
DEPLETION	0	0	0	0	0
OPERATING CASH FLOW	8807	9063	9377	9505	9799
CAPITAL EXPENDITURES	0	0	0	0	0
CAPITALIZED INTEREST	0	0	0	0	0
BONUS BID PAYMENTS	0	0	0	0	0
BORROWED FUNDS	0	0	0	0	0
DEBT PRINCIPAL	4364	4364	4364	4364	4364
WORKING CAPITAL	0	0	0	0	0
NET CASH FLOW	4443	4699	5014	5141	5435
SUNK COSTS	0	0	0	0	0
COVERAGE RATIOS					
INTEREST	5.55	7.05	9.41	14.85	29.65
DEBT SERVICE	0.40	0.52	0.72	1.09	2.25
CUMULATIVE CASH FLOW	23752	28452	33465	38607	44042

Table 4.4-I (page 4 of 5)

BASE CASE CASH FLOW

(UNITS IN THOUSANDS)

Major Production of 4080 Mw /day for \$ 6.90 Debt/Equy Ratio: 0.7000					
Project Year	16	17	18	19	20
PRODUCTION					
Exp Pwr Mw	1428	1428	1428	1428	1428
Tip Fee Tons	158	158	158	158	158
Recycle Lbs	541	541	541	541	541
REVENUES					
Exp Pwr Mw	9853	9853	9853	9853	9853
Tip Fee Tons	4725	4725	4725	4725	4725
Recycle Lbs	3658	3658	3658	3658	3658
GROSS REVENUE	18236	18236	18236	18236	18236
ROYALTY	0	0	0	0	0
OPERATING COSTS					
OPERATIONS	2286	2286	2286	2286	2286
MAINTENANCE	1407	1407	1407	1407	1407
TOTAL OPERATING COSTS	3693	3693	3693	3693	3693
INSURANCE	67	67	67	67	67
SEVERANCE TAX	0	0	0	0	0
INTEREST	0	0	0	0	0
PROPERTY TAXES	551	542	542	542	598
NET AFTER COSTS	13925	13934	13934	13934	13878
DEPRECIATION	977	977	0	0	0
NET AFTER DEPRECIATION	12948	12957	13934	13934	13878
DEPLETION	0	0	0	0	0
PRE-TAX PROFIT	12948	12957	13934	13934	13878
STATE INCOME TAX	432	432	464	464	463
NET TAXABLE INCOME (FED)	12516	12525	13469	13469	13415
FEDERAL INCOME TAX	3379	3382	3637	3637	3622
TAX CREDITS	0	0	0	0	0
NET AFTER TAXES	9137	9143	9833	9833	9793
PLUS: DEPRECIATION	977	977	0	0	0
DEPLETION	0	0	0	0	0
OPERATING CASH FLOW	10114	10120	9833	9833	9793
CAPITAL EXPENDITURES	0	0	0	0	0
CAPITALIZED INTEREST	0	0	0	0	0
BONUS BID PAYMENTS	0	0	0	0	0
BORROWED FUNDS	0	0	0	0	0
DEBT PRINCIPAL	0	0	0	0	0
WORKING CAPITAL	0	0	0	0	0
NET CASH FLOW	10114	10120	9833	9833	9793
SUNK COSTS	0	0	0	0	0
COVERAGE RATIOS					
INTEREST	0.00	0.00	0.00	0.00	0.00
DEBT SERVICE	0.00	0.00	0.00	0.00	0.00
CUMULATIVE CASH FLOW	54156	64276	74109	83941	93734

Table 4.4-I (page 5 of 5)

BASE CASE CASH FLOW

(UNITS IN THOUSANDS)

Major Production of 4080 Mw /day for \$ 6.90 Debt/Equity Ratio: 0.7000

Project Year	21	22	23	24
PRODUCTION				
Exp Pwr Mw	1428	1428	1428	1428
Tip Fee Tons	158	158	158	158
Recycle Lbs	541	541	541	541
REVENUES				
Exp Pwr Mw	9853	9853	9853	9853
Tip Fee Tons	4725	4725	4725	4725
Recycle Lbs	3658	3658	3658	3658
GROSS REVENUE	18236	18236	18236	18236
ROYALTY	0	0	0	0
OPERATING COSTS				
OPERATIONS	2286	2286	2286	2286
MAINTENANCE	1407	1407	1407	1407
TOTAL OPERATING COSTS	3693	3693	3693	3693
INSURANCE	67	67	67	67
SEVERANCE TAX	0	0	0	0
INTEREST	0	0	0	0
PROPERTY TAXES	596	596	596	596
NET AFTER COSTS	13880	13880	13880	13880
DEPRECIATION	0	0	0	0
NET AFTER DEPRECIATION	13880	13880	13880	13880
DEPLETION	0	0	0	0
PRE-TAX PROFIT	13880	13880	13880	13880
STATE INCOME TAX	463	463	463	463
NET TAXABLE INCOME (FED)	13417	13417	13417	13417
FEDERAL INCOME TAX	3623	3623	3623	3623
TAX CREDITS	0	0	0	0
NET AFTER TAXES	9795	9795	9795	9795
PLUS: DEPRECIATION	0	0	0	0
DEPLETION	0	0	0	0
OPERATING CASH FLOW	9795	9795	9795	9795
CAPITAL EXPENDITURES	0	0	0	0
CAPITALIZED INTEREST	0	0	0	0
BONUS BID PAYMENTS	0	0	0	0
BORROWED FUNDS	0	0	0	0
DEBT PRINCIPAL	0	0	0	0
WORKING CAPITAL	0	0	0	0
NET CASH FLOW	9795	9795	9795	9795
SUNK COSTS	0	0	0	0
COVERAGE RATIOS				
INTEREST	0.00	0.00	0.00	0.00
DEBT SERVICE	0.00	0.00	0.00	0.00

CUMULATIVE CASH FLOW	103529	113324	123118	132913
Internal Rate of Return: 22.65%	Payout Years 6			
Loan Amount 52364	Total New Capital 68460	Total Capital 68460		
Max Equity Exposure -26741	Total Capitalized Interest 6346			
Avg Yearly Cash Flow (Prod) 7066	Min Yearly Cash Flow 3686			
Net Present Value Table				
@0%	@5%	@10%	@15%	@20%
132913	59986	27586	11645	3013
				-2079
				-5309

By all financial evaluation criteria evaluated, this is an attractive project. Its economic desirability stems in part from its relatively low capital and operating costs. Our experience is that this type of project needs a 15% DCFROI, and 20% is generally considered very good. In Table 4.4-I, the monetary values are in thousands of dollars per year. The performance shown in the table may be overly conservative in several respects. There has been no attempt to optimize the schedule for maximum return. The anticipated production may also be conservative. A statistically "worst" case for the amount of pollutant added was used, thereby maximizing the amount of absorbent that would be required. However, these potential increases in profitability could be offset by other factors.

4.4.6 Sensitivity Analysis

As shown in Subsection 4.4.5, our representative project is financially attractive in terms of DCFROI, net present value, payoff period, equity exposure, and cumulative profit. However, the base case described therein does not explicitly address two major concerns: economic risk and the impact of factors beyond the project's control. These concerns are addressed by sensitivity analyses.

Table 4.4-II summarizes the results of our testing of assumptions and sensitivity analyses. Note that sensitivity is measured in terms of deviation of a single item from the base case, without considering simultaneous deviations in several variables. The project is profitable through the range of conditions tested.

The base case excludes expected price increases and cost escalation. The effect of inflation was not considered, and is not considered in the sensitivity analyses. However, the effect of variations in the tipping fee were analyzed. Environmental concerns and diminishing landfill accessibility in the United States have led to rapid increases in tipping fees for "advanced" waste disposal processes (average tipping fees are now increasing at slightly less than 7% percent per year- see Reference 4.4-2). Therefore, we investigated the financial impact of tipping fees of \$45 and \$60 per ton of municipal waste or industrial waste deposited at the facility. An increase to \$60 would increase the DCFROI from 22.7% to 32.9%, and the cumulative cash flow by more than 50% (from almost \$133 million to over \$202 million as shown in the right-hand column of the table).

TABLE 4.4-II
SENSITIVITY ANALYSIS

RUN	DCFROI (%)	PAYOUT (Years)	MAX EQUITY EXPOSURE (\$ * 1000)	CUM CASH FLOW (\$ * 1000)
Base Case	22.65	6	-26,741	132,913
Sensitivity				
Operating Costs +25%	20.45	7	-26,741	119,188
Operating Costs +50%	18.19	7	-26,741	105,467
Electricity @ \$0.04/kw-hr	12.76	12	-26,741	71,825
Capital Cost +25%	16.79	8	-31,888	114,553
Capital Cost +50%	12.29	13	-37,036	96,148
Tipping Fee @ \$45/ton	27.93	6	-26,741	167,686
Tipping Fee @ \$60/ton	32.91	5	-26,741	202,424
DER = 0 (All Project Financed)	14.16	8	-74,228	161,632

We considered several down-side risks, with results as illustrated under "Sensitivity" in Table 4.4-II. Three common problems are underestimation of capital costs, underestimation of operating costs, and overestimation of production or product sale prices.

The case in which capital costs prove to be 50% higher than estimated does not endanger the viability of the project, although it does reduce the DCFROI from 22.7% to 16.8%. Thus, the risk inherent in underestimating capital cost is moderate.

The project is somewhat less sensitive to underestimation of operating costs. Note that the project still returns 18.2% on equity if direct costs prove to be 50% higher than projected.

Perhaps the least favorable financial output is obtained under conditions in which the project receives only 4.0 cents per kw-hr of electric power sold. Under such a condition, the DCFROI would be 12.8% - still within the range usually specified as viable for this type of project. Thus, the project's economic viability is quite robust, i.e., able to withstand significant down-side risks.

Finally, note that 100-percent project financing (debt:equity ratio of zero) significantly decreases the project's projected DCFROI (to 14.2 percent) but produces a high cumulative cash flow (\$161.6 million) over a 24-year period. This accentuates the attractiveness of the project from a financial standpoint.

We emphasize our belief that the overriding advantage of our concept is its environmental benefits. However, we believe it is significant that these benefits can be obtained while maintaining economic viability.

5.0 FINDINGS AND CONCLUSIONS

Our conclusions do not differ in substance from those drawn by Hazen Research Inc. (see Appendix B); however, we have made some clarifications.

Our basic conclusions are:

A. The process works as expected. Oil shale absorbs acid gas pollutants and produces an ash which could be, at the least, disposed of in a normal landfill in an environmentally benign manner.

B. Based upon test data from the Hazen Research Inc. 6-inch circulating fluidized bed test unit, we find that:

1. An absorbent/pollutant (A/P) ratio of 3.0 is sufficient to meet Best Available Control Technology (BACT) requirements for acid gas pollutants,
2. The chlorine is essentially eliminated from the offgas,
3. At an A/P ratio of 4.5, over 98.5% of the acid gas pollutants are removed,
4. Temperature has a major effect on the viability of the process, and needs to be maintained between 1436° F and 1508° F (780 and 820° C),
5. Operating temperatures below 1400° F could significantly impact the process; however, temperatures in that range were not tested,
6. Temperatures of about 1450° F appear most desirable if the ash reaches sufficiently cementitious properties,
7. Data from the 15-inch tests indicates that the reduced residence time and/or reduced temperature in the freeboard resulted in reduced acid gas absorption.

C. Considering the results of the 15-inch Hazen unit tests in conjunction with those from the 6-inch test unit, we find that:

8. The 15-inch data without oil shale shows an excess sulfur output compared with the input; the excess probably came from contamination in the 15-inch system. If we compute the effects of contamination, and if we assume that the contamination is constant, we note that the effect of the excess output moves the 15-inch unit performance closer to that of the 6-inch unit, especially at high A/P ratios. This is shown in Table 4.3-I and Figure 4.3-1.

9. Using a 3.5:1 A/P ratio, we find that the amount of oil shale required for 95%+ removal of the acid gas is 15.8 pounds of oil shale per 100 pounds of RDF for the "as received" RDF without Ca and Mg. For the "severe" RDF case, 21.04 pounds of oil shale per 100 pounds of RDF

will be required.

10. At higher A/P ratios, it appears that changes in residence time may be overcome, with significant process ramifications.

11. The chemical reaction model fits the data with a .9+ probability. Therefore, the model is judged to be applicable to RDF/oil shale combustion as well as coal/oil shale mixtures.

12. The temperature effect is more pronounced at lower A/P ratios than at the higher ratios.

13. Oil shale ash fusion is not a problem at the operating temperatures tested.

D. Using data from all available tests and from the literature, we find that:

14. The process can easily meet all applicable acid gas emission regulations and all applicable regulations pertaining to solid waste disposal in landfills.

15. A representative facility employing this process appears to be economically viable over wide regions of the United States, with a discounted cash flow return on investment (DCFROI) of 22.65% and payout within six years from design commencement (less than four years of operation).

6.0 RECOMMENDATIONS

The following recommendations follow from our findings and conclusions:

1. Run tests on the 6-inch CFBC at reduced temperatures to determine reactor performance,
2. Determine the effect of residence time,
3. Modify the 15-inch reactor to duplicate temperature and residence times determined from the 6-inch reactor,
4. Produce sufficient ash for compression tests to determine whether construction material production from the system is viable,
5. Complete preliminary engineering on a full-scale model,
6. Determine the costs of a full-scale facility and verify the economics.

APPENDIX A

TEST PLAN

APPENDIX A

MSW/OIL SHALE TEST PLAN

The test plan is designed to demonstrate that oil shale co-combusted with municipal solid waste (MSW) can reduce gaseous pollutants (SO_2 , CO) to acceptable levels (90%+ reduction) and produce a cementitious ash which will, at a minimum, be acceptable in normal land fills. The small-scale combustion testing will be accomplished in a 6" circulating fluid bed combustor (CFBC) at Hazen Research Laboratories. This work will be patterned after the study we conducted in 1988 when coal and oil shale were co-combusted in a program sponsored by the Electric Power Research Institute.

The specific purposes of the test program will be to:

1. Determine the required ratio of oil shale to MSW by determining the ratio of absorbent to pollutant (A/P)
2. Determine the effect of temperature and residence time in the reactor.
3. Determine whether the kinetic model developed for coal/oil shale mixture is applicable.

Equipment: 6" circulating Fluid Bed Combustor with monitor for off-gas for SO_2 , NO_x , Cl^- , O_2 , CO .

Feeder for MSW and oil shale (separate)

Test 1. Analyze: MSW for Sulfur Chlorine
Oil shale for Ca, Mg, S, Cl, Grade

Establish: MSW feed rate nominal 12 #/hr (same as coal)
Oil shale rate to establish 2x
stoichiometric/pollutant A/P ratio

Temperature 1550 °F, residence time 2 seconds (gas)

NOTE: pollutant concentrations in exhaust after
reaching equilibrium.

Test 2. Same as above but go to 3x (A/P=1) ratio.

Test 3. If 2x satisfactory, go to 1.5x (A/P=1) ratio to establish minimum value. If 2x unsatisfactory, go to 4x A/P ratio.

Test 4. Chose the preferred A/P ratio: increase residence time to 4 seconds; obtain steady state.

Test 5. Same as 4 but reduce the A/P ratio by 25%.

Test 6. Same as 4 but increase the A/P ratio by 25%.

Hazen Research, Inc.
4601 Indiana St. • Golden, Colo. 80403
Tel: (303) 279-4501 • Telex 45-860
FAX: (303) 278-1528

Prepared for

Synfuels Engineering and Development, Inc.
4927 Piceance Creek Road
Rifle, Colorado 81650

**PILOT-SCALE TESTING OF INVENTION NO. 612:
COMBUSTION OF MUNICIPAL SOLID
WASTE WITH OIL SHALE IN A
CIRCULATING FLUIDIZED BED**

December 7, 1995

Copy No. 2
HRI Project 8493

**PILOT-SCALE TESTING OF INVENTION NO. 612:
COMBUSTION OF MUNICIPAL SOLID WASTE
WITH OIL SHALE IN A CIRCULATING FLUIDIZED BED**

Prepared by:

Kari D. Fox
Project Engineer

Approved by:

Harry Mudgett
Project Manager

Lawrence D. May
Senior Project Engineer

Rodney C. Hodgson
Vice President

TABLE OF CONTENTS

	Page
INTRODUCTION AND SUMMARY	1
PROCESS MATERIALS DESCRIPTION	4
TEST EQUIPMENT	11
SIX-INCH-DIAMETER CIRCULATING FLUIDIZED-BED SYSTEM	11
FIFTEEN-INCH-DIAMETER FLUIDIZED-BED SYSTEM	12
TESTING PROGRAM AND SUMMARY	24
TESTS PERFORMED FEBRUARY 6, 1995	25
TESTS PERFORMED MARCH 21, 1995	27
TESTS PERFORMED MARCH 23, 1995	27
ANALYTICAL TESTS	29
CONCLUSIONS	38
RECOMMENDATIONS	40
APPENDIX	
Logbook Entries	
Operational Data Sheets	
SO ₂ Mass Calculations	
Sample Calculations	

INTRODUCTION AND SUMMARY

Synfuels Engineering Development, Inc. (SED) secured a grant for testing of Invention No. 612, "Combustion of Municipal Solid Wastes with Oil Shale in a Circulating Fluidized Bed" through the Department of Energy Energy Related Inventions Program (DOE ERIP). In December 1994, SED authorized Hazen Research, Inc. to conduct a series of pilot-scale combustion tests in support of the grant project. A test plan was developed by SED and Hazen to demonstrate that co-combustion of municipal solid waste (MSW) and oil shale can reduce emissions of gaseous pollutants to acceptable levels.

The purpose of the program conducted at Hazen was to determine:

- the ratio of oil shale to MSW (absorbent/pollutant) which provides containment of 90% or more of acid gas emissions (SO_2 and HCl). For this study, the absorbent is defined as the total moles of calcium and magnesium and the pollutant as the total moles of sulfur and chlorine in the system.
- the effects of temperature and residence time on acid gas emissions using various absorbent/pollutant ratios.
- if a cementitious ash is produced which will, at a minimum, be acceptable in normal landfills.

Pelletized MSW was obtained from BFI Recycling Systems (BFI) that approximated the composition set forth by SED. Oil shale used during the testing was obtained from Rio Blanco Oil Shale. Sulfur dioxide and HCl gases were metered to the fluid bed to simulate higher percentages of sulfur (from rubber or leather wastes) and chlorine (from textile wastes), which might be encountered in the MSW but which were not present in the MSW evaluated for this program.

Five days of combustion testing were conducted in Hazen's 6-inch-diameter circulating fluid-bed combustor (CFBC) during February and March 1995. Hazen's 15-inch-diameter CFBC was used to perform an additional test on March 23, 1995. The demonstration runs were witnessed by Harry E. McCarthy and Reed Clayson, representatives of SED in the interface with Hazen.

The results from the 6-inch-diameter CFBC tests on March 21, 1995, are summarized below; the operating temperature varied between 1476 and 1503°F. It can be seen that while increasing the ratio of absorbent to pollutant (A/P) at a constant temperature, the containment of pollutants (S + Cl) increased correspondingly. The emissions were measured as SO₂ and HCl and then calculated to sulfur and chlorine equivalents.

**Six-inch-diameter CFBC
March 21, 1995**

Test Run	A/P Ratio	Pollutants					
		Input, lb		Emissions, lb		% Contained	
		S	S + Cl	S	S + Cl	S	S + Cl
1	1.77	0.04973	0.2157	0.01179	0.01869	76.3	94.1
4	2.98	0.05009	0.14897	0.00738	0.00780	85.2	94.7
3	3.73	0.09776	0.23426	0.00883	0.00942	91.0	96.0
2	4.65	0.19072	0.35955	0.00568	0.00646	97.0	98.3

The same trend can be seen in the test run of March 23, 1995, in the 15-inch-diameter CFBC. With an increasing A/P ratio, the amount of pollutant containment is greater. However, it should be noted that the percent containment of pollutants is consistently less in the 15-inch-diameter fluid-bed combustor tests compared to testing in the 6-inch-diameter combustor. The cause of this discrepancy is not known; however, an explanation for this effect could be that there was less retention time at temperature for the gaseous and solid reactants during testing with the 15-inch CFBC. The freeboard zone of the 15-inch CFBC was about 200°F lower than the bed temperature, whereas there was very little temperature differential between the bed and freeboard during testing in the six-inch CFBC. Additionally, if one assumes that SO₂ production from the pollutant is

constant, we can calculate that the production rate for an MSW-only case would be 3.18×10^{-3} moles per hour. This is based on the assumption that the MSW acts the same in the 6-inch and 15-inch combustors. If this production rate is constant, we can further assume that the containment of pollutant is from oil shale scrubbing. Using this, we can compute that the A/P ratio (for the total production of SO_2 only) is 3.3, which would allow sustained absorbing. The following tabulation and graph demonstrate the containment of pollutants related to various A/P ratios during testing in the 15-inch CFBC.

15-inch-diameter CFBC
March 23, 1995

Test Run	A/P Ratio	Pollutants					
		Input, lb		Emissions, lb		% Contained	
		S	S + Cl	S	S + Cl	S	S + Cl
2	3.49	0.23904	0.59723	0.10965	0.11646	54.1	80.5
3	4.00	0.23942	0.52384	0.09163	0.09691	61.7	81.5
4	4.45	0.29628	0.57763	0.05207	0.05487	82.4	90.5
5	4.74	0.34758	0.62931	0.04268	0.04389	87.7	93.0

By increasing the A/P ratio in the combustion process through the addition of oil shale, the amount of pollutant emissions can be greatly reduced.

PROCESS MATERIALS DESCRIPTION

The two process feeds used in this test program were MSW and oil shale. A "typical" composition of MSW, obtained from literature searches by SED, is provided in Table 1. The MSW sample in Table 1 represents an extreme scenario, demonstrating the highest concentrations of sulfur and chlorine that are likely to be present in MSW, based on published literature.

Two shipments of MSW were received at Hazen in conjunction with this project. The first shipment from the City and County of Denver of approximately five tons, received December 7, 1994, was rejected and returned to the landfill on December 12, 1994. The intent had been to manually sort the waste to obtain the desired composition of "typical" MSW. Blending and pelletizing was to follow. The program budget did not anticipate the labor-intensive activities associated with this preparation. A second shipment received on January 1, 1995, from BFI of Eden Prairie, Minnesota, of approximately two tons was accepted and used for the test runs. The actual composition of the MSW pellets as provided by BFI is shown in Table 2.

The MSW was delivered with negligible amounts of metal, glass, and dirt as prescribed in the test plan. The as-received pellets were approximately four inches long by one inch square; these pellets were blended and a representative portion obtained for analysis. Another portion was broken to one-inch top-size cubes using a hammermill. The broken MSW pellets were blended and split to produce a homogeneous mixture for feeding to the 6-inch-diameter fluid-bed combustor. The size reduction was necessary only for work conducted with the 6-inch fluid bed due to the relatively small size of the equipment. Testing conducted with the 15-inch-diameter fluid bed utilized the as-received MSW pellets.

Ultimate, proximate, heating value, and chlorine analyses of the MSW are provided in Table 3. Included are typical analysis results as provided by BFI and analysis results by Hazen on a representative split of the MSW that was tested during this program.

An x-ray fluorescence (XRF) spectrographic scan was performed on a sample of MSW ash that was produced by combustion in a muffle furnace. The XRF provides a semiquantitative scan for 31 metals. Results of the XRF, provided in Table 4, show that the MSW ash is composed mostly of alumina (25.9%) and silica (37.3%). The concentrations of CaO and MgO, which are known to be absorbents of sulfur dioxide, measured 16.6 and 2.16% respectively in the ash (1.6 and 0.2% as-received basis). The relatively high calcium and magnesium concentrations were not expected.

The supplier of the MWS pellets, BFI, routinely uses lime as a binder (approximately 12% by weight). Pellets produced for the combustion testing were made at BFI without lime addition; however, the pelletizing equipment apparently was not thoroughly cleaned prior to the Hazen pellet run. Emissions of SO_2 from combustion of MSW would likely be controlled to some extent by the presence of calcium and magnesium in the MSW ash.

Four drums of oil shale were received at Hazen from SED on October 20, 1994, totaling approximately 1200 pounds (lb). The oil shale was prepared by crushing to minus 6 mesh (3.35 millimeters) and blending. Next, the sample was crushed to minus 20 mesh (0.85 millimeters), blended, and split. Representative portions were taken for analysis while the remaining splits were used for the test runs. Ultimate, proximate, heating value, and chlorine analyses results of the oil shale (performed by Hazen) are compiled in Table 3. The particle size distribution of the prepared oil shale is given in Table 4. The mean particle size was approximately 48 mesh (0.3 millimeters).

Included in Table 5 are the results of selected chemical analyses performed on the oil shale. The silica assay of the shale was 40.9%. Calcium and magnesium, reported as oxides, assayed 9.2% and 5.6%, respectively.

Absorbent was defined as the total calcium and magnesium present in the MSW and oil shale (moles per hour) and the pollutant was defined as the total sulfur and chlorine (moles per hour) present in the MSW and oil shale. Since the pelletized MSW used for these combustion tests did not contain the desired quantity of pollutants, cylinders of SO_2 and HCl were used to spike the process gas to reach the anticipated concentration of sulfur and chlorine in a "severe" waste composition (0.21% sulfur and 0.70% chlorine). These values were obtained from literature research by SED. The A/P ratio was defined as the ratio of pound moles of calcium and magnesium to pound moles of sulfur and chlorine on a one-hour basis for the given flow rates of each run. A sample calculation for a typical A/P ratio is given in the Appendix.

**Table 1. Typical Waste Composition
and Predicted Ultimate Analysis of High-concentration MSW**

Item	Preseparation wt%	Postseparation wt%	Ultimate Analysis of High-concentration MSW	
			Sulfur, wt%	Chlorine, wt%
Paper and Paperboard	37.45	43.58	0.140	0.115
Glass	6.75	0.39	0.000	0.000
Metals				
Ferrous	6.29	0.12	0.000	0.000
Aluminum	1.38	0.03	0.000	0.000
Other Nonferrous	0.61	0.59	0.000	0.000
Plastics				
Rigid	5.02	5.84	0.600	2.250
Film	3.26	3.79	0.083	0.528
Rubber and Leather	2.35	2.73	1.500	0.000
Textiles	2.86	3.33	0.050	3.480
Wood	6.29	7.31	0.250	0.170
Food Wastes	6.74	7.85	0.150	0.320
Yard Trimmings	17.88	20.81	0.220	0.320
Misc. Inorganic Wastes	1.48	1.72	0.090	0.410
Other	1.64	1.91	0.090	0.410
Total	100.00	100.00	3.173	8.003
Total Weight % in Typical High-concentration MSW			0.221	0.436

Table 2. Pellet Composition
Provided by BFI Recycling Systems

Material	Weight Percent
Paper	64.40
Plastic	4.90
Cardboard	3.76
Dirt	2.44
Wood	1.20
Textiles	1.02
Organic Material	0.76
Styrofoam	0.62
Aluminum Foil	0.06
Foam Rubber	0.04
Glass	0.04
Metal	0.04
Rubber	0.04
Moisture	20.50
Total	99.82

**Table 3. Ultimate, Proximate, Heating Value, and Chlorine Analyses of
MSW and Oil Shale**

	MSW per BFI		MSW per Hazen		Oil Shale per Hazen	
	As Received	Dry Basis	As Received	Dry Basis	As Received	Dry Basis
Ultimate, %						
Moisture	12.08	0.00	8.20	0.00	0.62	0.00
Carbon	14.43	16.41	41.90	45.64	12.14	12.22
Hydrogen	1.56	1.78	5.35	5.83	1.34	1.35
Nitrogen	0.11	0.12	0.43	0.47	0.29	0.29
Sulfur	0.06	0.07	0.17	0.19	1.35	1.36
Ash	17.68	20.11	9.86	10.74	75.28	75.75
Oxygen ^{1/}	54.08	61.51	34.09	37.13	8.98	9.03
Total	100.00	100.00	100.00	100.00	100.00	100.00
Proximate, %						
Moisture	12.08	0.00	8.20	0.00	0.62	0.00
Ash	17.68	20.11	9.86	10.74	75.28	75.75
Volatile	59.68	67.88	n/a	n/a	n/a	n/a
Fixed Carbon	10.56	12.01	n/a	n/a	n/a	n/a
Total	100.00	100.00	18.06	10.74	75.90	75.75
Heating Value						
Btu/lb	5660	6438	7445	8110	1630	1640
Chlorine, %			0.50	0.54	0.01	0.01

^{1/} Oxygen determined by difference

n/a Not assayed

Table 4. Oil Shale Particle Size Distribution

Tyler Sieve Size	Direct Weight	Direct Weight %	Cumulative Retained	Weight % Passing
28	81.6	16.6	16.6	83.4
35	102.7	20.9	37.5	62.5
48	70.8	14.4	52.0	48.0
65	44.8	9.1	61.1	38.9
100	26.7	5.5	66.6	33.4
150	31.5	6.4	73.0	27.0
200	18.6	3.8	76.8	23.2
270	13.6	2.8	79.6	20.4
325	5.1	1.0	80.6	19.4
400	4.7	1.0	81.6	18.4
Pan	90.4	18.4	100.0	0.0
	490.5	100.0		

**Table 5. Chemical Characterization Analysis
of MSW and Oil Shale**

Constituent	MSW Ash ^{1/}	As-received Oil Shale ^{2/}
Analysis, %		
Aluminum (Al ₂ O ₃)	25.90	n/a
Barium (BaO)	0.11	n/a
Calcium (CaO)	16.60	9.2
Chlorine	2.72	0.01
Iron (Fe ₂ O ₃)	3.38	n/a
Magnesium (MgO)	2.16	5.59
Manganese (MnO)	0.11	n/a
Phosphorus (P ₂ O ₅)	0.93	n/a
Potassium (K ₂ O)	1.63	2.31
Silicon (SiO ₂)	37.30	40.9
Sodium (Na ₂ O)	3.12	2.72
Sulfur (SO ₃)	0.86	3.38
Titanium (TiO ₂)	5.45	n/a
Total, %	100.27	
Analysis, ppm		
Arsenic	40	
Chromium	335	
Cobalt	21	
Copper	1927	
Lead	743	
Molybdenum	39	
Niobium	19	
Nickel	60	
Rubidium	43	
Strontium	1700	
Tin	148	
Thorium	49	
Tungsten	55	
Uranium	31	
Vanadium	123	
Yttrium	20	
Zinc	1440	
Zirconium	205	
Total, ppm	6998	
Total, %	0.70	

^{1/} Results were determined by XRF spectrographic scan.

^{2/} Results were determined by atomic absorption.

n/a Not assayed.

TEST EQUIPMENT

SIX-INCH-DIAMETER CIRCULATING FLUIDIZED-BED SYSTEM

Hazen's externally heated, 6-inch-diameter CFBC test facility was used during the first phase of this testing program. This facility included a stainless steel fluid-bed vessel six inches in diameter, external cyclone, baghouse with fabric filters, and an offgas scrubber with recirculating caustic (NaOH) solution to neutralize any remaining acid gases before exhausting to the atmosphere. A schematic of the six-inch facility is provided as Figure 1. Detailed drawings of the reactor, windbox, and cyclone are shown in Figures 2, 3, and 4, respectively.

The six-inch fluidized-bed system was equipped with numerous instruments to permit control of the combustion operation and to obtain operating information:

- Ten Type K thermocouples were located at various heights within the fluid-bed zone to measure the temperature of the combustor.
- A multi-point recorder with digital display was used to continuously log system temperatures.
- Magnehelic gauges measured pressures throughout the system.
- Magnehelic gauges were used to indicate differential pressure (d.p.) across the fluidized bed, cyclone, and baghouse. A differential pressure recorder was used to monitor the bed d.p. signal.
- Flowmeters measured auxiliary air, propane, sulfur dioxide, and HCl additions to the unit.
- An orifice meter was used to measure the flow of primary fluidizing air.
- A computer data acquisition system was used to continuously log operating information, selected system temperatures, and offgas composition.

The fluidized-bed vessel was encased in an insulating chimney and was externally heated with a propane-fired combustion chamber. Air and propane mixtures were ignited in the combustion chamber and entered the insulated chimney one foot above the bottom of the vessel. Additional

propane injection nozzles were located at three elevations in the chimney to maintain operating temperatures during testing. Hot combustion gases rise up the chimney and discharge from a side port near the top of the six-inch vessel.

The total reaction length of the 6-inch vessel was 21 feet measured from the air distribution plate to the cyclone inlet. A fluidizing-gas distribution plate, containing 21 orifices, was attached to a removable windbox. Each orifice was 0.141 inch in diameter, resulting in a total open area of 0.326 square inch. This windbox assembly was inserted into the bottom of the vessel and held in place with flanges. A 1.5-inch-diameter bed discharge port, positioned in the center of the windbox, extended 12 inches into the combustor and was used periodically to collect bed ash from the system. Cyclone ash was recirculated to the 6-inch vessel through a J-leg and 45-degree-angle port that entered the reaction vessel at the distribution plate (see Nozzle 1 on Figure 2). Fine particle ash (fly ash) that escaped the cyclone was collected in a baghouse.

Prepelletized MSW feed was metered from a hopper with a variable-speed belt conveyor and entered the combustor through a 6-inch-diameter rotary valve on the vessel lid. A feed transfer pipe (a 4-inch-diameter stainless steel tube) was positioned through the reactor lid and extended into the 6-inch-diameter vessel about 3.5 feet below the cyclone inlet. This tube was installed to deliver feed into the combustion zone and to reduce the potential for short circuiting the MSW directly to the cyclone.

Oil shale was introduced to the CFBC using a variable speed two-inch-diameter screw feeder and lock-hopper system. The metered feed discharged from the screw and passed through a stainless steel tube that extended vertically into the fluid bed through the windbox and distribution plate. A portion of the fluidizing air was used to assist in the transport of the feed through the tube and into the fluid-bed zone.

HCl and SO₂ gases were added during selected tests to simulate emissions from combustion of MSW wastes that have higher concentrations of chlorine and sulfur. These gaseous additions (supplied from compressed gas cylinders) were injected into the vessel through the windbox.

FIFTEEN-INCH-DIAMETER FLUIDIZED-BED SYSTEM

The second phase of this program was conducted using a circulating fluidized-bed combustor 15 inches in diameter. The system included the combustor, an external cyclone and baghouse, and an exhaust gas scrubber. An induced-draft blower was installed downstream of the scrubber to exhaust

gases from the system and to control pressure within the combustor. The fluid-bed vessel was refractory-lined with 4½ inches of high-temperature, abrasion-resistant castable backed by 4½ inches of insulating castable to minimize heat losses. The 15-inch-diameter section extended to a height of five feet followed by a three-foot length of expanded freeboard, 21 inches in diameter. Applicable drawings of this system, including an overall layout, reactor detail, cyclone, and air distribution plate details, are provided as Figures 5, 6, 7, and 8, respectively.

The 15-inch fluidized-bed system was equipped with numerous instruments to permit control of the combustion operation and to obtain operating information:

- Twenty-one Type K thermocouples were used to measure temperatures in the preheating windbox, combustor, and exhaust gas system. Eleven thermocouples were located at various heights within the 15-inch-diameter section.
- A multi-point recorder with digital display was used to continuously log system temperatures.
- Magnehelic gauges measured pressures throughout the system.
- Magnehelic gauges were used to indicate d.p. across the fluidized bed, cyclone, and baghouse. A differential-pressure recorder was used to monitor the bed d.p. signal.
- Flowmeters measured auxiliary air, propane, sulfur dioxide, and HCl additions to the unit.
- An orifice meter was used to measure the flow of primary fluidizing air.
- A computerized data acquisition system was used to continuously log operating information, selected system temperatures, and offgas composition.

The combustor was directly heated by a propane-fired windbox attached to the bottom of the 15-inch-diameter section. Fluidization air was metered into the windbox and entered the combustor through a distribution plate equipped with 14 T-bar tuyeres. The tuyeres assure equal distribution of the fluidization gas across the cross-sectional area of the fluidized bed.

MSW was metered to the 15-inch fluid bed with a variable-speed belt conveyor. The metered feed discharged from the conveyor and passed through a rotary valve which provided a gas seal to reduce the possibility of unmetered air entering the combustor or process gases escaping the combustor. Following the rotary valve, the MSW entered a screw 4 inches in diameter which

conveyed the material into the fluid bed at a point 22 inches above the distribution plate. Air was metered into the screw in order to reduce the possibility of having moisture from the wet MSW condensing within the screw.

Oil shale was fed to the combustor using the two-inch-diameter variable-speed screw and lock-hopper system that was used during testing in the six-inch system. The metered oil shale was pneumatically conveyed and entered the reactor horizontally through a side port located approximately six inches above the distribution plate.

Hot gases from the fluid bed passed to an eight-inch-diameter external cyclone. Dust particles that disengaged from the gas stream entered a two-inch-diameter screw which fed the cyclone dust back into the fluid-bed combustor. Gases and fine-particle dust which escaped the cyclone passed to a baghouse for final dust cleaning. A wet-scrubbing circuit of caustic solution was used to neutralize any remaining acid gases before exhausting the process gas to the atmosphere.

HCl and SO₂ gases were added during selected tests to simulate emissions from combustion of MSW wastes that have higher concentrations of chlorine and sulfur. These gaseous additions (supplied from compressed gas cylinders) were metered with rotameters and injected into the vessel through the windbox.

During testing (in both the 6-inch and the 15-inch fluid-bed systems), the process exhaust gas was sampled continuously at the baghouse inlet and analyzed for oxygen (O₂), carbon dioxide (CO₂), carbon monoxide (CO), nitrogen oxides (NO_x), and sulfur dioxide (SO₂) using continuous emissions monitors (CEM). The gas sample passed through a heated two-micron filter to remove any residual entrained dust. The sample was then cooled rapidly to condense and remove water from the analyzed gas. Specific analyzers used during this combustion program are listed as follows:

O ₂ , %	Infrared Industries Model 2200
CO ₂ , %	Infrared Industries Model 703
SO ₂ , ppm	Thermo Electron Model 40
CO, ppm	Beckman Industries, IR Model 864
NO _x , ppm	Beckman Industries, Model 951A

In addition to CEM measurements, the process exhaust gas was measured periodically for HCl content using a Sensidyne™ sampling apparatus.

Process conditions, selected temperatures, and exhaust gas composition were recorded continuously using a Molytek data acquisition system and computer. Operational data were recorded at one-hour intervals on logsheets, and comments concerning the operation of the system were recorded in a project journal.

FIGURE 1. 6" CIRCULATING FLUID-BED REACTOR EQUIPMENT SCHEMATIC

FIGURE 2. 6-INCH REACTOR DETAIL

FIGURE 3. 6-INCH FLUID-BED REACTOR WINDBOX

FIGURE 4. 6-INCH-DIAMETER CYCLONE DETAIL

Figure 5. 15-Inch-diameter Fluidized-bed Reactor

Hazen Research, Inc.

FIGURE 6. 15 INCH FLUIDIZED-BED REACTOR DETAIL

PLAN VIEW

ELEVATION

FIGURE 7. 8-INCH-DIAMETER CYCLONE DETAIL

FIGURE 8. AIR DISTRIBUTION PLATE

TESTING PROGRAM AND SUMMARY

This program was planned under the direction of SED, corresponding to a research grant from DOE ERIP for the testing of Invention No. 612, "Combustion of Municipal Solid Wastes with Oil Shale in a Circulating Fluidized Bed." This program used fluidized-bed combustors 6 inches and 15 inches in diameter. Equipment shakedown occurred on January 30 and 31, and February 1, 1995. Test runs using the 6-inch CFBC were performed on February 6 and March 21, 1995, and using the 15-inch CFBC on March 23.

The following abbreviations are used throughout this report.

scfm gas flow, standard cubic feet per minute

SSV superficial space velocity in feet per second (ft/sec), calculated at operating temperature and pressure assuming an empty vessel. The following formula was used to calculate the space velocity.

$$SSV = \frac{\text{scfm}}{\text{Area}} \times \frac{T_{\text{bed}}}{T_{\text{std}}} \times \frac{P_{\text{std}}}{P_{\text{fb}}} \times \frac{1}{60}$$

where T_{bed} = operating temperature ($^{\circ}\text{R}$)

T_{std} = standard temperature (70°F , 530°R)

P_{std} = standard pressure, 1.0 atmosphere (14.7 psi)

P_{fb} = freeboard pressure, psia (gauge + barometric)

Area = cross-sectional area of fluid-bed combustor, in square feet (ft^2)

Barometric pressure in Denver, Colorado is 0.83 atmospheres

For each 6-inch-diameter combustor test run, the CFBC was preheated to approximately 1500°F . Gas flows were set at approximately 28 scfm. Material for a starting bed was added and ash recirculation was established. MSW was fed at approximately 12 to 14 pounds per hour (lb/hr) and oil shale was fed at varying rates, typically between 1.5 and 6 lb/hr. SO_2 and HCl gases were metered into the combustor to simulate high-pollutant waste. The draft control was adjusted to maintain a zero pressure point in the combustor freeboard. Temperature adjustments were made using the external chimney heater.

The purpose of this test program was to determine the effect of varying the oil shale feed rate on the acid gas emissions (SO_2 and HCl). The amounts of sulfur and chlorine in the MSW were of primary concern as pollutant; calcium and magnesium are known to be absorbers of these pollutants in combustion processes. When comparisons between different runs were made, an A/P ratio was used. The absorbent is defined as the amount of calcium and magnesium in moles per hour in the system, and pollutant as the total amount of sulfur and chlorine (moles per hour). The A/P ratio is defined as the respective ratio of the two values. Calcium and magnesium were found in both the oil shale and the MSW. Sulfur and chlorine existed in the oil shale, MSW, and the metered gases (SO_2 and HCl respectively).

The shakedown of the equipment on January 30 and 31 demonstrated the importance of having sufficient ash for recirculation and providing a good heat sink to diminish temperature fluctuations. Sand was added to the ash in the bed to increase the recirculating load. Insufficient ash within the system is often encountered when conducting short runs on a pilot-scale basis. Other problems discovered during the equipment shakedown testing included bridging of the MSW within the entry pipe and periodic burning of MSW on the feeder valve due to its close proximity to the hot furnace. On a larger scale, bridging should not cause significant problems because the entry pipe would be much larger.

Tests conducted on February 1 demonstrated good recirculation of ash and stable feed rates of MSW and oil shale. The average temperature within the combustor ranged between 1500°F (815°C) and 1600°F (871°C). At the higher temperatures, the SO_2 measured in the process exhaust gas fluctuated; the stability of the results was questioned, and the results are therefore not included in detail in this report.

Tests originally scheduled for February 2 were postponed until February 6 due to high wind conditions which resulted in the inability to maintain temperatures within the 6-inch CFBC.

TESTS PERFORMED FEBRUARY 6, 1995

The objective of the tests performed in the 6-inch CFBC on February 6 was to determine the temperature dependence of the process at various A/P ratios. Operating temperature and the A/P ratio were varied simultaneously to see which had the greater effect on the emissions of SO_2 . It was predicted by a previous EPRI-sponsored project in 1988 that high A/P ratios would decrease the acid gas emissions (SO_2 and HCl) while higher temperature would tend to depress absorption of acid gases.

Table 6 summarizes the operating conditions for each of the five tests conducted on February 6. Emissions of SO_2 versus temperature and various A/P ratios for the five tests are graphically depicted in Figure 9.

From the test data it can be seen that even though the A/P ratio is increasing, the SO_2 emissions generally rise with increasing temperature. The temperature of the test run is noted above the corresponding data point on the graph. The trend shows the increase in temperature above a certain point affects the SO_2 emissions, which increase correspondingly with temperature. It is therefore concluded that the temperature of the combustion process has a greater influence on reduction of acid gas emissions than the A/P ratio. From this day of testing, the "best" temperature for the remaining runs was determined to be between 1450 and 1510°F (787 to 821°C).

Tests 1 and 4 confirmed the importance of temperature within the combustor on the resulting control of acid gas emissions. For Test 1, using an A/P ratio of 4.62, the SO_2 concentration in the exhaust gas was measured at 12 parts per million (ppm), while Test 4, having essentially the same the A/P ratio (4.56) resulted in a measurement of 141 ppm SO_2 , an order of magnitude difference. The distinction between the two tests is found in the operating temperature within the fluid-bed vessel during each experiment; Test 1 was conducted at 1438°F (781°C) while Test 4 was conducted at 1504°F (818°C) apparently causing the substantially higher SO_2 measurement. This data would then suggest that the "best" temperature might lie at the lower end of the 1438 to 1504°F (781 to 818°C) range and could feasibly be below 1438°F (781°C). However, further tests would need to be performed to determine a more precise optimal temperature for pollutant absorption.

Additionally, a comparison between Test 1 and Test 2 shows that at a nearly constant temperature (1438°F vs. 1450°F, respectively) the A/P ratio affects the SO_2 emissions. Test 1, using an A/P ratio of 4.62, had emissions measured at 12 ppm SO_2 , while Test 2 showed 22 ppm SO_2 in the exhaust gas for an A/P ratio of 3.59. While an increase in the A/P ratio from 3.59 to 4.62 decreased the SO_2 concentration by one-half at essentially the same operating temperature, it should be noted that this change appears to be minimal compared to the magnitude of change seen with an increase in temperature by less than 100°F. It is therefore concluded that, while increasing the A/P ratio can reduce the emission level of SO_2 , changes in temperature show a more substantial effect.

From this day's testing, it was determined that SO_2 emissions tend to increase with increasing combustion temperature and decrease with increasing A/P ratios. Also, when both the A/P ratio and temperature are varied, temperature has the dominant effect on the SO_2 emissions.

TESTS PERFORMED MARCH 21, 1995

On March 21, 1995, four tests were performed in the 6-inch CFBC to determine the best ratio of A/P to be used for the absorption of acid gas contaminants during the combustion of MSW with oil shale.

The tests were performed on essentially a constant temperature basis, ranging from 1476 to 1503°F (802 to 817°C). Gas compositions, flow rates, HCl and SO₂ additions, etc. (see Table 7) were maintained approximately the same as the operating conditions of the tests performed on February 6. The A/P ratio was varied from test to test over a range of 1.77 to 4.10. The graphical depiction of the A/P ratio versus the concentration of SO₂ in the process exhaust gas is shown in Figure 10.

Figure 10 shows the relationship between the A/P ratio and the SO₂ emissions, at essentially constant temperature and MSW feed rate. The graph clearly shows a decrease in the SO₂ emissions with the increasing A/P ratio. Very little change is seen in the SO₂ concentration until the A/P ratio exceeds 3.0, and a substantial change is seen with an A/P ratio greater than 3.5. Tests performed on March 21 successfully demonstrated a correlation between the A/P ratio and the emissions of SO₂.

During this day's testing, it was determined that at a constant temperature, increasing A/P ratios produced lower SO₂ emissions. Also, the feed rate of oil shale, as a form of process control, could be varied to adjust the A/P ratio in order to meet desired emission levels.

TESTS PERFORMED MARCH 23, 1995

Five tests were performed using the 15-inch CFBC on March 23, 1995, to utilize the larger-scale CFBC for confirmation tests. Flow rates of gases, solids feed (MSW and oil shale), and HCl and SO₂ additions were scaled up proportionally to the six-inch CFBC testing. The A/P ratio was varied, and the temperature was maintained essentially constant in a range of 1476 to 1519°F (803 to 826°C). A summary of the operating conditions is provided in Table 8.

The SSV in the bed zone was approximately 5.3 ft/sec in the 15-inch CFBC, while it was nearly twice that value using the six-inch combustor (8.4 ft/sec and 10.2 ft/sec on February 6 and March 21, 1995, respectively). The 15-inch-diameter CFBC is 8 feet tall, including a 3-foot length of expanded freeboard (equivalent length is about 11 feet) and the 6-inch combustor is 21 feet tall. A comparison of the gas residence times is given in the following tabulation, indicating that even

though the superficial space velocities varied between the two combustor systems, similar residence times were achieved.

CFBC	SSV ft/sec	Residence Time, sec
6"	8.4	2.5
6"	10.2	2.1
15"	5.3	2.1

As the A/P ratio was increased (see Figure 11), the SO₂ emissions were reduced, which is consistent with the trend observed during testing on March 21 using the 6-inch CFBC. However, when comparing the SO₂ emissions of March 21 (6-inch CFBC) with those of March 23 (15-inch CFBC), there is a distinctly apparent difference in the concentrations of the SO₂ for similar A/P ratios. The SO₂ concentrations measured during the 6-inch CFBC testing were much lower than those measured in the 15-inch CFBC for similar A/P ratios and operating temperature. The following tabulation provides a comparison of absolute SO₂ values measured during the combustion testing with the two test systems.

6-inch CFBC 3/21/95		15-inch CFBC 3/23/95	
A/P	ppm SO ₂	A/P	ppm SO ₂
1.77	40	1.72	280
3.73	32.5	4.00	175
4.65	15.6	4.74	78

The difference appears to arise in the difference in combustor design. The 6-inch-diameter combustor does not have an expanded-diameter freeboard, but is the same diameter throughout its length and, during testing, had a relatively constant temperature profile throughout the length of the vessel. This was due to the fact that the 6-inch CFBC was externally heated, and heat could be applied as required. The 15-inch-diameter combustor has an expanded freeboard where the upper 3-foot section of the combustor is 21 inches in diameter. When combusting a fuel that does not display significant freeboard burning tendencies, the freeboard zone of the vessel might have a temperature significantly less than the temperature in the bed zone. That is apparently the situation

that occurred during this program; therefore, the temperature in the freeboard zone of the 15-inch CFBC may not have been high enough to allow the oil shale to adequately react with the pollutant emissions. The following tabulation provides a comparison between the temperatures measured in the bed and freeboard zones of the 15-inch CFBC during the testing conducted on March 23, 1995.

Test	Temperature, °F	
	Fluid-bed Vessel	Freeboard
1	1508	1386
2	1519	1317
3	1479	1342
4	1477	1357
5	1506	1393

During this day's tests, the effects of scaleup and a CFBC with expanded freeboard were determined. The SO₂ emissions were still affected by temperature and A/P ratio in the same manner as during previous tests. The reduction in the total acid gas production (SO₂ and HCl) was greater than 92%. However, the SO₂ emissions from the 15-inch CFBC were somewhat higher than the emissions from the 6-inch CFBC. The higher concentration of SO₂ in the 15-inch combustor was caused by differences in temperature and residence time as compared with the 6-inch combustor. The lower temperature within the freeboard and the reduced residence time at temperature negatively affects the amount of pollutant absorption by the oil shale.

ANALYTICAL TESTS

Stabilization tests and Toxicity Characteristic Leaching Procedure (TCLP) analysis were conducted to determine if cementitious ash product could be disposed of in a sanitary landfill. Ash produced from the fluid-bed combustion of MSW and oil shale was collected as baghouse fines (fly ash) and bed discharge or bottom ash. The stabilization tests utilized ash collected from Test 3 of the six-inch fluid-bed operation on March 21, 1995. The total ash collected consisted of 85% (by weight) baghouse fines and 15% bed discharge. Two stabilization test mixtures were prepared with this ash; the first used Type II Portland cement as the binding agent. The second used the Portland cement and granulated blast furnace slag. The formulations for the stabilization testing are given in Table 9.

The ash samples and reagents were mixed with water to form a thick paste. Each prepared mixture was put in a mold and allowed to cure at ambient conditions for seven days. Following the cure period, the molds were crushed to 100% passing $\frac{3}{8}$ inch and subjected to the TCLP. The results of the leachate analyses for the RCRA metals are given in Table 10. Both samples passed the TCLP.

The leachate from the mixture containing only Type II Portland cement contained 0.49 mg/l chromium, whereas the leachate from the sample stabilized with blast furnace slag contained only 0.08 mg/l chromium. This demonstrated that the reducing agents in the blast furnace slag (sulfides) were successful in fixating this RCRA metal such that it was less susceptible to leach from the solids.

Table 6. Summary of Average Operating Conditions
Fluidized-bed Combustion Testing of MSW/Oil Shale in the Six-inch CFBC

	Test 1	Test 2	Test 3	Test 4	Test 5
Test Date	2/6/95	2/6/95	2/6/95	2/6/95	2/6/95
Test Duration, hours	0.8	0.5	0.5	1.8	0.2
Temperatures, °F					
Fluid-bed Vessel	1438	1449	1465	1504	1578
Cyclone Outlet	1164	1202	1206	1186	1206
Cyclone Leg	1173	1189	1207	1162	1164
J-leg	1119	1042	993	898	883
Baghouse Inlet	433	469	484	484	430
Baghouse Outlet	313	338	358	363	363
Process Air Flows, scfm					
Primary Fluidizing Air	25.1	25.1	25.1	25.1	25.1
Oil-shale-feed Transport Air	2.4	2.4	2.4	2.4	2.4
Bed-pressure Tap Air	0.3	0.3	0.3	0.3	0.3
HCl Addition, cc/min	189	189	189	189	189
SO ₂ Addition, cc/min	51	51	51	51	51
Total Gas Flow, scfm	27.8	27.8	27.8	27.8	27.8
Gas Velocity in Bed, ft/sec	8.4	8.4	8.4	8.4	8.4
Gas Retention Time, sec	2.5	2.5	2.5	2.5	2.5
Pressures					
Windbox, psig	4.0	4.0	4.0	4.0	4.0
Bed Differential, "H ₂ O	11.0	10.0	10.0	8.0	10.0
Freeboard Pressure, "H ₂ O	1.0	0.0	0.0	0.5	2.0
Cyclone Differential, "H ₂ O	3.0	3.0	3.2	3.2	3.4
Baghouse Differential, "H ₂ O	27.0	26.0	25.0	29.0	28.0
Feed and Product Data, lb/hr					
MSW Feed Rate	12.5	14.0	14.3	14.3	14.3
Oil Shale Feed Rate					
Target	6.0	3.0	3.0	6.0	8.0
Actual	6.0	2.8	3.2	6.1	8.6
Bed-ash Recovery	n/a	n/a	n/a	n/a	n/a
Baghouse Ash	1.7	1.7	1.1	1.5	1.7
Exhaust Gas Composition (dry-volume basis)					
O ₂ , %	11.6	14.1	13.5	12.2	9.9
CO ₂ , %	8.4	5.9	6.4	7.7	10.2
CO, ppm	1311.0	1089.0	1113.0	1125.0	1852.0
SO ₂ , ppm	12.0	22.0	87.0	141.0	418.0
NO _x , ppm	242.0	128.0	163.0	358.0	418.0
HCl, ppm	<5	<5	<5	<5	<5
Absorbent, lb/hr	0.76	0.46	0.5	0.79	1.04
Pollutant, lb/hr	0.21	0.18	0.18	0.22	0.26
Molar A/P Ratio, $\frac{\text{moles (Ca + Mg)}}{\text{moles (SO}_2 + \text{Cl}_2)}$	4.62	3.59	3.74	4.56	5.00

Figure 9. Tests 1 to 5
Six-inch-diameter CFBC
February 6, 1995

Table 7. Summary of Average Operating Conditions
Fluidized-bed Combustion Testing of MSW/Oil Shale in the Six-inch CFBC

	Test 1	Test 2	Test 3	Test 4
Test Date	3/21/95	3/21/95	3/21/95	3/21/95
Test Duration, hours	1.8	1.7	1.4	1.0
Temperatures, °F				
Fluid-bed Vessel	1486	1486	1503	1476
Cyclone Outlet	1026	1119	1144	1144
Cyclone Leg	880	1009	1044	1047
J-leg	777	862	858	855
Baghouse Inlet	372	435	473	507
Baghouse Outlet	293	327	358	405
Process Air Flows, scfm				
Primary Fluidizing Air	24.5	23.6	23.9	23.7
Oil-shale-feed Transport Air	2.4	2.4	2.4	2.4
Bed-pressure Tap Air	0.3	0.3	0.3	0.3
HCl Addition, cc/min	190	190	190	190
SO ₂ Addition, cc/min	50	50	50	50
Total Gas Flow, scfm	28.0	27.1	27.1	27.4
Gas Velocity in Bed, ft/sec	10.3	10.0	10.2	10.0
Gas Retention Time, sec	2.0	2.1	2.1	2.1
Pressures				
Windbox, psig	3.0	3.0	3.0	3.0
Bed Differential, "H ₂ O	4.0	4.0	4.0	4.0
Freeboard Pressure, "H ₂ O	0.0	0.0	0.0	0.0
Cyclone Differential, "H ₂ O	2.2	2.4	2.8	2.1
Baghouse Differential, "H ₂ O	11.0	9.5	9.0	10.0
Feed and Product Data, lb/hr				
MSW Feed Rate	11.1	12.4	12.1	12.4
Oil-shale Feed Rate				
Target	0.0	6.0	3.0	1.5
Actual	0.0	6.1	3.0	1.5
Bed-ash Recovery	0.0	0.0	1.4	1.0
Baghouse Ash	0.0	5.4	4.2	3.0
Exhaust Gas Composition (dry volume basis)				
O ₂ , %	14.4	12.2	12.4	11.0
CO ₂ , %	5.8	8.3	7.8	8.2
CO, ppm	1681	1150	630	1244
SO ₂ , ppm	40	16	33	39
NO _x , ppm	369	555	550	516
HCl, ppm	<5	<5	<5	<5
Absorbent, lb/hr	0.14	0.76	0.46	0.31
Pollutant, lb/hr	0.12	0.21	0.17	0.15
Molar A/P Ratio, $\frac{\text{moles (Ca + Mg)}}{\text{moles (SO}_2 + \text{Cl}_2)}$	1.77	4.69	3.73	2.98

Figure 10. Tests 1 to 4
Six-inch-diameter CFBC
March 21, 1995

Table 8. Summary of Average Operating Conditions
Fluidized-Bed Combustion Testing of MSW/Oil Shale in the Fifteen-inch CFBC

	Test 1	Test 2	Test 3	Test 4	Test 5
Test Date	3/23/95	3/23/95	3/23/95	3/23/95	3/23/95
Test Duration, hours	0.7	1.3	1.0	1.0	1.0
Temperatures, °F					
Windbox	340	306	280	259	237
Fluid-bed Vessel	1508	1519	1479	1477	1506
Freeboard	1386	1317	1342	1357	1393
Cyclone Outlet	977	930	939	932	973
Baghouse Inlet	531	514	500	473	478
Baghouse Outlet	406	419	406	390	392
Process Air Flows, scfm					
Primary Fluidizing Air	82.1	83.0	82.9	82.9	83.0
MSW-feed Transport Air	2.7	2.7	2.7	2.7	2.7
Oil-shale-feed Transport Air	1.5	1.5	1.4	1.5	1.5
HCl Addition, cc/min	633	633	633	633	633
SO ₂ Addition, cc/min	168	168	168	168	168
Total Gas Flow, scfm	88.6	89.5	89.3	89.4	89.5
Gas Velocity in Bed, ft/sec	5.3	5.3	5.0	5.3	5.0
Gas Retention Time, sec	2.1	2.1	2.1	2.1	2.1
Pressures					
Windbox, psig	0.7	0.7	0.7	0.7	0.7
Bed Differential, "H ₂ O	3.5	3.2	3.5	4.0	4.0
Freeboard Pressure, "H ₂ O	0.0	0.0	0.0	0.0	0.0
Cyclone Differential, "H ₂ O	3.0	2.5	2.2	2.2	2.0
Baghouse Differential, "H ₂ O	0.1	0.2	0.2	0.2	0.3
Exhaust Draft, "H ₂ O	-6.0	-7.5	-7.8	-8.0	-9.2
Feed and Product Data, lb/hr					
MSW Feed Rate					
Target	32.0	32.0	32.0	32.0	32.0
Actual	34.2	30.5	32.2	31.5	31.5
Oil-shale Feed Rate					
Target	0.0	8.0	12.0	16.0	20.0
Actual	0.0	7.6	11.5	15.8	19.6
Bed-ash Recovery	0.0	0.0	0.0	0.0	0.0
Baghouse Ash	0.0	0.2	0.3	1.5	6.1
Exhaust Gas Composition (dry-volume basis)					
O ₂ , %	11.0	11.6	11.3	11.3	11.4
CO ₂ , %	8.0	8.2	8.7	9.6	8.4
CO, ppm	102	114	100	124	88
SO ₂ , ppm	280	175	175	98	78
NO _x , ppm	460	495	495	520	560
HCl, ppm	10	10	10	<5	<5
Absorbent, lb/hr	0.44	1.15	1.56	1.98	2.36
Pollutant, lb/hr	0.38	0.46	0.52	0.58	0.63
Molar AP Ratio, $\frac{\text{moles (Ca + Mg)}}{\text{moles (SO}_2 + \text{Cl}_2)}$	1.72	3.49	4.00	4.49	4.74

Figure 11. Tests 1 to 5
Fifteen-inch-diameter CFBC
March 23, 1995

Table 9. Fluid-bed Ash Stabilization Testing Formulations

Sample No.	Fluid-bed Ash, g		Reagent Additives, g		
	Bed	Baghouse	Type II Cement	Blast Furnace Slag	Water Added
1	9.0	141.0	50	0	87
2	8.4	131.6	6	54	78

Table 10. Toxicity Characteristic Leaching Procedure (TCLP)
Leachate Analysis Results

RCRA Metal	Sample 1 mg/l	Sample 2 mg/l	EPA Regulatory Limit mg/l
Arsenic	<0.01	<0.01	5.0
Barium	0.2	0.2	100
Cadmium	0.03	0.04	1.0
Chromium	0.49	0.08	5.0
Lead	0.4	0.4	5.0
Mercury	<0.0001	<0.0001	0.2
Selenium	<0.01	<0.01	1.0
Silver	0.03	0.03	5.0

CONCLUSIONS

This test work demonstrated that co-combustion of MSW and oil shale under some conditions can reduce emissions of gaseous pollutants (SO_2 and HCl) by greater than 92%. Ash stability and TCLP tests were conducted to determine if the process would also produce a cementitious ash product which could be disposed in a sanitary landfill.

During this program it was determined:

- Reduced SO_2 and acid gas emissions occurred while increasing the A/P ratio during the combustion process.
- Combustion temperatures greater than about 1510°F (821°C) tended to elevate the SO_2 emissions.
- If both the A/P ratio and the combustion temperature were varied, temperature generally had the predominant control on absorption of acid gas emissions.
- Due to the lower temperature in the freeboard, the 15-inch-diameter CFBC system showed higher SO_2 emissions.
- Residence time also appeared to affect the amount of pollutant absorption by the oil shale.
- The "best" temperature range for this process appeared to be between 1436 and 1508°F (780 to 820°C), with strong indication that the lower end of that range is more desirable.
- An operating A/P ratio should be chosen that is sufficient to stay within desired emission limits.
- A/P ratio of less than 3 showed only marginal decrease in SO_2 .
- A/P ratio between 3 and 4 showed significant decrease in SO_2 emissions.
- An A/P ratio greater than 4 showed substantial decrease in the level of SO_2 emissions.

If one assumes that SO₂ production from the contaminant is constant, we can calculate that the production rate in the MSW-only case is 3.18×10^{-3} moles per hour. This is based on the assumption that the MSW acts the same in the 6-inch and 15-inch combustors. If this production rate is constant, we can further assume that the reduction of emissions is from oil shale scrubbing. Using this, we can compute that the A/P ratio for the total production of SO₂ only is 3.3, which would allow sustained absorbing. The following tabulation shows the reduction in emissions compared to the A/P ratio while operating each of the test systems.

CFBC	Test Run	A/P Ratio	% SO ₂ Reduction	% Acid Gas Reduction
6"	1	1.77	76.3	94.1
6"	4	2.98	85.2	94.7
6"	3	3.73	91.0	96.0
6"	2	4.65	97.0	98.3
15"	2	3.49	54.1	80.5
15"	3	4.00	61.7	81.5
15"	4	4.45	82.4	90.5
15"	5	4.74	87.8	92.9

RECOMMENDATIONS

- Additional test runs should be considered in order to further optimize the combustion temperature.
- More studies would need to be performed to determine optimal residence time.
- The feed rate of oil shale could be used as a form of process control. Since the composition of wastes will vary, the A/P ratio may have to be adjusted. In the operation of a system for co-combustion of MSW and oil shale, if the monitored SO₂ emissions were too high, the temperature of the system should be checked. If the temperature is outside of the optimal range (either too high or too low) then it should be corrected first. If the level of SO₂ still does not return to specifications, then the oil-shale feed rate (essentially the A/P ratio) should be increased accordingly. Or conversely, in order to conserve resources, if the SO₂ level is below necessary limits, the oil shale rate could be decreased.

Hazen Research, Inc.
4601 Indiana St. • Golden, Colo. 80403
Tel: (303) 279-4501 • Telex 45-860
FAX: (303) 278-1528

Prepared for

Synfuels Engineering and Development, Inc.
4927 Piceance Creek Road
Rifle, Colorado 81650

ANNEX A

**PILOT-SCALE TESTING OF INVENTION NO. 612:
COMBUSTION OF MUNICIPAL SOLID
WASTE WITH OIL SHALE IN A
CIRCULATING FLUIDIZED BED**

March 19, 1996

Table 1

**Toxicity Characteristic Leaching Procedure
Leachate Analysis Results of Fluid Bed Ash
for
Synfuels Engineering and Development, Inc.**

Mold prepared using bed and baghouse ash products collected during 6-inch fluid bed testing of MSW/oil shale on March 21, 1995 (Test #3). Mold consisted of 15% bed and 85% baghouse ash (by weight).

<u>RCRA Metal</u>	<u>Leachate mg/liter</u>	<u>EPA Regulatory Limit, mg/liter</u>
Arsenic	<0.03	5.0
Barium	<0.5	100
Cadmium	0.03	1.0
Chromium	0.70	5.0
Lead	<0.5	5.0
Mercury	<0.001	0.2
Selenium	<0.03	1.0
Silver	0.05	5.0

*Advanced
Combustion
Engineering
Research
Center*

FLUIDIZED-BED COMBUSTION OF MUNICIPAL SOLID WASTES WITH OIL SHALE

Combustion Tests and Analysis of Results

Craig N. Eatough, Ph.D.

Description of Tests

Combustion of pelletized municipal solid waste (MSW) was performed in fluidized-beds with ground western oil shale as a fuel feed. The ash portion of the oil shale has the properties required for acid gas capture within the fluid bed. The resulting spent bed material may be highly cementitious. The captured pollutant species in the bed ash may be non-leachable when included in the cement.

The objectives of the combustion tests were to 1) establish combustion conditions that enhance acid gas (particularly SO_2 and HCl) retention in the bed, and 2) investigate the effect of MSW to oil shale ratio on the ability of the bed material to absorb acid gases and retain them in the bed.

Sulfur Capture

Sulfur compounds found in oil shale and MSW can be classified as either organic or pyritic. During combustion, essentially all of the organically bound sulfur is released. The release of pyritic sulfur from the inert portion of the shale is dependent on various parameters including temperature, reaction time, and the presence of catalytic material in the ash. Meyers (1977) indicated that in 800°C air about 90% of pyritic sulfur is oxidized. Under excess oxygen combustion conditions, sulfur will be oxidized and released primarily as SO_2 .

The practice of adding a basic solid, such as calcium oxide, to the bed material in fluidized bed combustors (FBC) is used to absorb and retain sulfur in the bed. CaO reacts with SO₂ to form calcium sulfate, effectively removing SO₂ from the combustion gases. Calcium oxide is formed by the *in situ* calcination of limestone as follows:

and desulfurization proceeds in oxidizing atmospheres according to either of the two following routes:

or in reducing atmospheres according to:

Route (3) is significant only if SO₃ producing catalysts, such as heavy metal salts, are present in the bed (Yates 1983).

Limestone and dolomite are frequently used as sources of calcium for sulfur capture in fluid-bed combustion. The effectiveness of sorbents for sulfur capture is generally described by the Ca/S mole ratio and varies according to their physical structure and chemical make-up. Borgwardt and Harvey (1972) tested 11 specimens representing a broad spectrum of limestones and dolomites. They found that sulfation reaction rates increased with an increase in surface area of the calcine, corresponding to smaller pores, but that reaction capacity increased with the pore size of the calcines. They also found dolomites to be more reactive than limestones. Lyngfelt and Leckner (1989a) suggest that there should be an optimum calcine pore-size distribution. This distribution includes large pores to facilitate transport far into the lime particle, and smaller pores giving sufficient area of reaction. The pore-size distribution depends on both the limestone and the calcination conditions. Higher calcination temperatures for limestones produce bigger pores and hence less surface area (Mulligan et al. 1989).

Dolomites undergo two stages of calcination:

The half-calcination reaction for dolomites occurs rapidly at temperatures above 600°C and results in the development of considerable porosity (Yang et al. 1975, Roberts et al. 1983). The calcination of calcium carbonate is thought to occur on the surface of the much larger MgO particles exposing a large surface area of CaO. The high reaction capacity of dolomites (100% CaO conversion to CaSO₄) is accounted for by the fact that the ratio of pore volume to CaO content is about twice that of limestone, providing adequate space to accommodate the entire reaction product (Borgwardt and Harvey 1972). Limestones containing significant amounts of silicon have also been shown to have higher rates of sulfation due to the same process. Attempts to sulfate MgCO₃ and MgO have been unsuccessful, even at 969°C, indicating that the high sulfur-retention capacities of dolomites are not due to magnesium sulfate formation but to the development of a highly porous structure with a large CaO surface area (Yang et al. 1975, Mulligan et al. 1989).

Lin et al. (1993) found that with the addition of limestone, NO_x emissions increase. Ammonia can be an important precursor to NO_x formation in fluidized bed combustion. It is suggested that there is a catalytic effect of CaO on oxidation of NH₃ to NO accounting for higher NO_x emission levels with sorbent addition. Studies show that the presence of CO decreases the observed increased NO formation significantly.

Nordin (1995) performed co-combustion tests with high sulfur fuels and biomass in a 5 kW FBC. He achieved high sulfur retention (up to 90%) in the bed and reported that alkali metals from the biomass acted as an important sulfur capture medium since he identified potassium and sodium sulfate in the ash.

Temperature Dependence of Sulfur Retention

The sulfur capture ability of limestones in fluidized beds has been shown to fall dramatically at temperatures above about 850°C (Yates 1983, Lyngfelt and Leckner 1989b,

Mulligan et al. 1989, Hansen et al. 1993). Although the reason is not fully understood, several explanations have been proposed based on both physical and chemical theories.

Physical Theories: The surface area of calcines decreases with increasing calcination temperature. At higher temperatures the structure of the calcined limestone is impaired due to sintering and annealing which decreases the fine surface structure of the calcine and therefore CaO area available for sulfation. Also, as temperature is increased, the rates of calcination and sulfation increase. Stantan (1983) proposed that simultaneous rapid calcination and sulphation lead to premature pore-blocking preventing utilization of the interior of the sorbent particles. He also suggested that higher heating rates cause particle decrepitation resulting in smaller particles and increasing elutriation.

Chemical Theories: Chemical theories generally follow the idea that sulfation reaction (2) requires oxygen to progress. Once formed, CaSO_4 is very stable in oxidizing atmospheres even at FBC combustion temperatures. At higher temperatures, volatiles combustion rates increase reducing local oxygen concentrations in the bed. As oxygen concentration decreases, reaction (2) decreases. It has also been shown that reductive decomposition of calcium sulfate occurs when appreciable amounts of CO or H are present according to:

or

where reaction (7) would result in a release of SO_2 . In tests performed in a 16-MW fluid-bed boiler it was determined that at temperatures below about 830°C the effect of reaction (7) was small. As temperature increased, the effect of reductive decomposition of CaSO_4 became more apparant and above 880°C a net release of SO_2 was noted even with an overall stoichiometric ratio of 1.4 (Lyngfelt and Leckner, 1989b). At 930°C the amount of SO_2 being emitted was more than double the amount of sulfur added in the form of fuel sulfur.

Gas composition measurements in fluidized beds show that, even under overall fuel lean combustion, the dense bed contains appreciable amounts of CO (Hansen et al., 1993) allowing

reaction (7) to proceed. As the temperature increases, reductive decomposition becomes increasingly more dominant because; 1) the rate of volatiles combustion increases, thus depleting the oxygen supply in the bed; 2) the rate of char gasification increases in the bed increasing the CO/H₂ concentration; 3) the kinetic reaction rate of reaction (7) increases; and 4) the equilibrium SO₂ concentration under reducing conditions increases with temperature.

Chlorine Capture

As with sulfur capture, attempts have been made to capture halogen species with active limestone and dolomite sorbents. The halogenation reaction is:

where X = F, Cl, Br, or I. The halogen capture ability of MgO is generally not included in studies presumably due to its ineffectiveness. Liang et al. (1991) attempted chlorine capture with limestone in a CFBC with a bed temperature of 850°C and a Ca/S molar ratio of 2.7. At the conditions tested they achieved poor chlorine capture (about 5%) which they attributed to the low melting point of CaCl₂ (782°C). They explained high capture rates reported by other investigators by the fact that wet or dry post combustion scrubbers were universally used.

Anthony et al. (1993) also report poor chlorine capture by limestone in a fluid-bed operated at 900°C. They also found that chlorine addition leads to an increase in NO_x. However, they concluded that the liquid calcium halide phases can modify the partially sulfated sorbent particles, leading to increased SO₂ capture.

Bradshaw et al. (1989) reported good chlorine capture (about 80%) with limestone and dolomite sorbents. However, they operated their bed at 540°C, below the melting point of CaCl₂. A kinetic analysis of their data showed that chlorine was captured at a faster rate than sulfur. It was suggested that HCl capture is controlled by external diffusion while SO₂ removal rate was controlled by intraparticle diffusion or chemical reaction and was, therefore, slower.

Several investigators have found that as chlorine is increased, CO concentrations increase (Bradshaw et al. 1989, Bulewicz et al. 1989, and Anthony et al. 1993). This is due to the strong

tendency of halogens to be free radical scavengers. Free radicals are necessary to carry out chain-propagating reactions and therefore their reduction results in an inability to convert CO to CO₂. Bulewicz et al. (1989) found that the increase in CO was more prominent for low volatile fuels. Anthony et al. (1993) found an order of magnitude increase in CO with a halogen mole fraction of only 10^{-3} in the fluidizing air. The effect appears to be smaller for larger-scale reactors or in the presence of an active sorbent such as CaO.

FBC Combustion Tests

Test Equipment

Hazen's 6-inch diameter CFBC bench scale unit consists of a stainless steel vessel equipped with a cyclone, bag house, and wet scrubbing system. A schematic diagram of the pilot unit is shown in Figure 1. The stainless steel reaction vessel extends to a height of 23 feet and has a cross-sectional area of 0.22 sq ft. This vessel is encased in a refractory-lined chimney so that processing temperatures can be maintained by heating the outer shell of the vessel with a propane-fired heater.

During combustor operation, a controlled amount of primary air was injected at the bottom of the reaction vessel, where the primary air was combined with feeder air, which transported a controlled amount of fuel solids. After entering the preheated vessel, the fuel combusted and the fine ash products were carried by the process gases through a hot cyclone. A majority of the ash separated from the gases and accumulated in a J-leg device for recycling to the reaction vessel. The portion of ash not separated from the process gas by the cyclone was transported to the bag house and collected as fly ash. Upon leaving the bag house, the exhaust gas was conveyed to a venturi scrubber before exhausting to the atmosphere.

Additional solid product was collected as bottom ash. This material represented the large ash particles which were not transportable by the process gases within the combustion vessel.

Flue gases from the CFBC were monitored continuously for O₂, CO₂, CO, SO₂, NO_x, and hydrocarbons. The sampling port for the analyses was located immediately upstream of the bag house.

The CFBC system was equipped with thermocouples to measure temperatures continuously throughout the length of the combustion unit and the exhaust gas system. Pressures at various locations in the system were monitored. Primary and auxiliary air flows were measured with rotometers. All data pertaining to the operation were recorded on data sheets and by logbook entries.

Pilot-scale combustion tests were performed in Hazen's 15-inch diameter, refractory lined, fluidized bed reactor. It is equipped with an emissions treatment system similar to that of the 6-inch facility consisting of a hot cyclone, bag house, and wet scrubbing system. Process monitoring and data acquisition systems were also similar to that on the 6-inch unit. A schematic diagram of the 15-inch unit is shown in Figure 2.

Both the 6-inch and 15-inch CFBC's were operated to establish a retention time in the combustion zone of about 2 seconds. This necessitated a superficial velocity of about 10 fps in the 6-inch bed and about 5.5 fps and 2.8 fps in the bed zone and freeboard of the 15-inch unit respectively.

Combustion Test Results

The analysis of the oil shale and municipal solid waste used in these combustion tests is given in Table 1.

Table 1. Analysis of MSW and Oil Shale Used in Combustion Tests (wt. %, as received basis).

	M	A	C	H	N	O ¹	S	Cl	CaO ²	MgO ²	HHV Btu/lb
MSW	8.20	9.86	41.90	5.35	0.43	34.09	0.17	0.50	1.64	0.21	7445
OS	0.62	75.28	12.14	1.34	0.29	8.98	1.35	0.01	9.20	5.57	1630

¹ By difference.

² Percent of ash.

Figure 1. 6-inch circulating fluid-bed combustor.

Figure 2. 15-inch fluidized bed reactor.

An initial series of tests to determine the effect of bed temperature on sulfur retention were performed in Hazen's 6-inch FBC. Two subsequent test series, one performed in the 6-inch fluid bed, and one in the 15-inch FBC investigated the effect of varying the ratio of MSW to oil shale feed on pollutant gas capture. The test matrix for the three series of combustion tests is shown in Table 2.

Table 2. Summary of FBC Test Conditions.

	Test 1	Test 2	Test 3	Test 4	Test 5
Series I (6-inch bed)					
Bed Temperature (°C)	781	787	798	818	859
MSW Feed Rate (lb/hr)	12.5	14.0	14.3	14.3	14.3
Oil Shale Feed Rate (lb/hr)	6.0	2.8	3.2	6.1	8.6
Fluidizing Air Flow (scfm)	25.1	25.1	25.1	25.1	25.1
MSW Transport Air (scfm)	0.0	0.0	0.0	0.0	0.0
Shale Transport Air (scfm)	2.4	2.4	2.4	2.4	2.4
HCl Addition (cc/min)	189.0	189.0	189.0	189.0	189.0
SO ₂ Addition (cc/min)	51.0	51.0	51.0	51.0	51.0
Gas Velocity in Bed (ft/sec)	8.4	8.4	8.4	8.4	8.4
Series II (6-inch bed)					
Bed Temperature (°C)	808	808	817	802	
MSW Feed Rate (lb/hr)	11.1	12.4	12.1	12.4	
Oil Shale Feed Rate (lb/hr)	0.0	6.1	3.0	1.5	
Fluidizing Air Flow (scfm)	24.5	23.6	23.9	23.7	
MSW Transport Air (scfm)	0.0	0.0	0.0	0.0	
Shale Transport Air (scfm)	2.4	2.4	2.4	2.4	
HCl Addition (cc/min)	190.0	190.0	190.0	190.0	
SO ₂ Addition (cc/min)	50.0	50.0	50.0	50.0	
Gas Velocity in Bed (ft/sec)	10.3	10.0	10.2	10.0	
Series III (15-inch bed)					
Bed Temperature (°C)	820	826	804	803	819
MSW Feed Rate (lb/hr)	34.2	30.5	32.2	31.5	31.5
Oil Shale Feed Rate (lb/hr)	0.0	7.6	11.5	15.8	19.6
Fluidizing Air Flow (scfm)	82.1	83.0	82.9	82.9	83.0
MSW Transport Air (scfm)	2.7	2.7	2.7	2.7	2.7
Shale Transport Air (scfm)	1.5	1.5	1.4	1.5	1.5
HCl Addition (cc/min)	633.0	633.0	633.0	633.0	633.0
SO ₂ Addition (cc/min)	168.0	168.0	168.0	168.0	168.0
Gas Velocity in Bed (ft/sec)	5.3	5.3	5.0	5.3	5.0

Effect of CFBC Configuration

The difference in size and geometry provided unique operating characteristics for the two furnaces. Since test conditions were established to maintain gas residence times in the combustors of both furnaces of approximately 2 seconds, the different reactor heights dictated that gas velocities, and therefore degree of bed agitation, be different. The superficial gas velocity for the 6 inch bed was about 10 fps while that for the 15 inch bed was about 5.5 fps decreasing to about 2.8 fps in the 21 inch diameter freeboard region. These differences create distinct mixing intensities which effect combustion conditions. They also influence the rate of elutriation of bed material which may effect sorbent effectiveness. The differences in combustion conditions are evidenced by the combustion efficiency of each furnace shown in Figure 3.

Figure 3. Relationship between O₂ and CO for the two fluid-beds used in this study. (6" bed; SSV=10.3 fps, MSW=11.1lb/hr. 15" bed: SSV= 5.5 fps, MSW=34.2 lb/hr.)

Figure 3 shows that even at high excess oxygen levels, a significant amount of unburned CO is present particularly in the 6 inch bed. Although, due to its higher gas velocity, the 6-inch bed has a higher potential mixing rate, its high surface area to volume ratio may allow for premature combustion product quenching resulting in high concentrations of CO. Figure 3 demonstrates that a significant concentration of CO is likely available in either bed for reductive decomposition of calcium sulfate to occur.

Effect of Bed Temperature

During the initial combustion tests performed in the 6-inch CFBC, bed temperature was varied by changing the oil shale feed rate (see Table 1). Test results illustrated in Figure 4 show a significant decrease in sulfur capture in the bed (ca. 98 to 60 percent) as temperature increases from about 780-860°C.

Figure 4. Effect of bed temperature on sulfur capture efficiency of sorbent, Ca/Pollutant molar ratio of 1.7-2.5, stoichiometric ratio 1.4-1.6.

Effect of Oil Shale Addition on NO_x

Previous fluidized-bed combustion studies by Lin et al. (1993) found that with the addition of limestone, NO_x emissions increased. This trend was also noted in these combustion tests as is indicated by Figure 5. Presumably, CaO catalyzes the oxidation of ammonia to NO accounting for this increase. Although it has been reported that the presence of CO decreases the CaO catalytic effect significantly, in these tests the NO increase was more pronounced for the 6-inch FBC which produced higher effluent CO concentrations. It may be that the higher mixing rates experienced in the 6-inch bed plays an important role in this effect. Higher mixing rates improve solid-gas contacting effectiveness which is necessary for catalytic reactions to occur. Also, although the 6-inch bed produced more effluent CO which has been shown to lower the increase in NO_x levels, no information is known about local bed CO concentrations for either the 6-inch or 15-inch tests.

Figure 5. Effect of sorbent feed rate on effluent NO_x levels.

Effect of Ca/Pollution Ratio

MSW feed rates were maintained at about 12 lb/hr in the 6-inch bed and 32 lb/hr in the 15-inch bed while the oil shale feed rate was varied. Bed temperatures were maintained at about 800°C for each bed. During combustion tests no measureable HCl was detected in the effluent gases. Maximum theoretical gas sulfur concentrations were calculated for each test condition assuming all sulfur forms in the feed streams were oxidized to SO₂. Figure 6 shows how the percent of total SO₂ that is captured or otherwise retained in the bed is effected by the calcium to pollutant (SO₂ + HCl) ratio. Sulfur capture efficiency tends to peak at a Ca/Pollutant of about 2.75 for both the 6-inch and 15-inch beds.

Figure 6 indicates that the 6-inch bed is more effective at retaining sulfur than the 15-inch bed. There are several factors which may account for this difference, including: 1) due to the high surface area to volume ratio of the smaller fluid-bed, heat loss is greater resulting in an overall cooler bed preventing release of a significant portion of the pyritic sulfur and severely limiting the amount of reductive decomposition of captured sulfur; and 2) The increased mixing in the 6-inch bed, evidenced by the higher superficial velocity, allows better solid-gas contact for sulfur absorption. The maximum sulfur retention efficiencies measured from these tests compare with the very best efficiencies reported for limestones and dolomites which range from 20 to 90% at temperatures between 800-850°C (Mulligan et al., 1989). The intimate contact between the sulfur released from the oil shale and the portion of the shale acting as the sorbent may account for the unusually high sulfur capture ability of the shale.

Figure 6. Effect of calcium to acid gas ($\text{SO}_2 + \text{HCl}$) ratio on the effectiveness of the bed to retain sulfur.

No HCl was detected in the effluent gases during combustion tests. Previous studies report that chlorine capture by limestone is very poor at temperatures above 850°C (Liang et al. 1991 and Anthony et al. 1993), and good capture (80%) at temperatures around 550°C (Bradshaw et al. 1989). The high chlorine capture noted in these tests, where bed temperatures were maintained around 800°C , may have occurred in the freeboard region or in the cyclone which operated between 500 and 600°C . Also, there may be other active chlorine capture media that can retain chlorine at the bed temperature, such as alkali metals, which may contribute to the high chlorine capture. Potassium and sodium chloride have melting temperatures around 800°C . An alkali analysis of the feed materials was not performed for these tests.

Implications to FBC Design

Net sulfur capture in fluidized beds is a result of the trade-off between sulfur absorption and release within the bed. Mechanisms for sulfur capture and release are sensitive to both temperature and local oxygen availability. Results from these tests substantiate those of previous investigations that net sulfur capture is highest at a bed temperature around 800°C. Large-scale studies have shown that the optimum temperature varies somewhat based on combustor design. It appears that a major factor influencing the optimum temperature is the extent of localized reducing zones within the bed (Hansen et al. 1993). Minimizing these zones will be an important parameter to be considered in the final FBC design. The use of a comprehensive, multi-dimensional FBC computer model (several of which are available, Brewster et al. 1993) could provide valuable insight into methods of minimizing reducing zones.

A strong effect of Ca/pollutant mole ratio on sulfur capture was noted in this study. Under the conditions tested a Ca/pollutant mole ratio of about 2.7 was required to achieve maximum sulfur capture. The sorbent feed rate should be determined in light of both its ability to reduce sulfur emissions as well as to increase nitrogen pollutant emissions. Investigation into how this parameter is effected by scale-up could be performed with an FBC model utilizing pollutant chemistry submodels.

Other factors that should be investigated in following-on tests include determining the optimum sorbent particle size, the effect of gas velocity, and the mechanism and location of HCl capture.

References

- Anthony, E.J., Bulewicz, E.M., and Preto, F., "The Effect of Halogens on FBC Systems," 1993 International Conference on Fluidized Bed Combustion, Vol 1, ASME, New York, NY, pp. 41-52, (1993).
- Borgwardt, R.H., and Harvey, R.D., "Properties of Carbonate Rocks Related to SO₂ Reactivity," Environmental Science & Technology, 6, pp. 350-360, (1972).
- Bradshaw, W.M., Krishnan, R.P., Young J.M., and Mohrman, G.B., "Laboratory Testing of a Fluidized-Bed Dry-Scrubbing Process for the Removal of Acidic Gases From a Simulated Incinerator Flue Gas," AIChE Symposium Series 270, AIChE, New York, NY, pp. 94-100, (1989).
- Brewster, B.S., Hill, S.C., Radulovic, P.T., and Smoot, L.D., Comprehensive Modeling, cpt. 8 in Fundamentals of Coal Combustions for Clean and Efficient Use, L.D. Smoot, Ed., Elsevier, New York, NY, (1993).
- Bulewicz, E.M., Janicka, E., and Kandefer, S., "Halogen Inhibition of CO Oxidation During the Combustion of Coal in a Fluidized Bed," 1989 International Conference on Fluidized Bed Combustion, Vol 1, ASME, New York, NY, pp. 163-168, (1989).
- Hansen, P.F.B., Dam-Johansen, K. and Ostergaard, K., "High-Temperature Reaction Between Sulphur Dioxide and Limestone - V. The Effect of Periodically Changing Oxidizing and Reducing Conditions," Chemical Engineering Science, 48, pp. 1325-1341, (1993).
- Liang, D.T., Anthony, E.J., Loewen, B.K., and Yates, D.J., "Halogen Capture by Limestone During Fluidized Bed Combustion," 1991 International Conference on Fluidized Bed Combustion, Vol 2, ASME, New York, NY, pp. 917-922, (1991).
- Lin, W., Johnsson, J.E., Dam-Johansen, K., and van den Bleek, C.M., Interactions Between NO_x Emission and Desulphurization in FBC, 1993 International Conference on Fluidized Bed Combustion, Vol 2, ASME, New York, NY, pp. 1093-1100, (1993).
- Lyngfelt, A. and Leckner, B., "Sulphur Capture in Fluidised-Bed Combustors: Temperature Dependence and Lime Conversion," Journal of the Institute of Energy, pp. 62-72, March (1989a).
- Lyngfelt, A. and Leckner, B., "SO₂ Capture in Fluidised-Bed Boilers: Re-Emission of SO₂ Due to the Reduction of CaSO₄," Chemical Engineering Science, 44, pp. 207-213, (1989b).
- Meyers, Robert A., Coal Desulfurization, Marcel Dekker, Inc., New York, NY, (1977).
- Mulligan, T., Pomeroy, M. and Bannard, J.E., "The Mechanism of the Sulphation of Limestone by Sulphur Dioxide in the Presence of Oxygen," Journal of the Institute of Energy, pp. 40-47, March, (1989).
- Nordin, A., "Optimization of Sulfur Retention in ash when Cocombusting High Sulfur Fuels and Biomass Fuels in a Small Pilot Scale Fluidized Bed," Fuel, 74, pp. 615-622, (1995).

- Roberts, A.G., Pillai, K.K. and Stantan, J.E., Pressurized Fluidized Combustion, from Fluidized Beds - Combustion and Applications, (ed J.R. Howard), Applied Science Publishers, New York, NY, (1983).
- Stantan, J.E., Sulphur Retention in Fluidized Bed Combustion, from Fluidized Beds - Combustion and Applications, (ed J.R. Howard), Applied Science Publishers, New York, NY, (1983).
- Yang, R.T., Cunningham, P.T., Wilson, W.I., and Johnson, S.A., Kinetics of the Reaction of Half-Calcined Dolomite with Sulfur Dioxide, from Sulfur Removal and Recovery from Industrial Processes, (ed J.B. Pfeiffer), Advances in Chemistry Series 139, American Chemical Society, Washington, D.C., (1975).
- Yates, J.G., Fundamentals of Fluidized-bed Chemical Processes, Butterworths, Boston, MA, (1983).