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Research and Development

COMBUSTION OF OIL SHALE IN FLUIDIZED-BED COMBUSTORS AN OVERVIEW

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COMBUSTION OF OIL SHALE IN
FLUIDIZED-BED COMBUSTORS -
AN OVERVIEW

Final Report

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Prepared by

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DISCLAIMER

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INTRODUCTION

The potential for using raw oil shale, spent shale, or coal-oil shale mixtures as a direct fuel, fuel supplement, or SO_2 sorbent in atmospheric fluidized-bed combustors (AFBC) has been examined. Such uses of the AFBC could include the following:

- direct combustion of raw (low-grade) oil shales (<20 gal oil per ton shale),

- direct combustion of retorted or spent shale,

- direct combustion of coal-oil shale mixtures wherein the oil shale is utilized as an SO_2 absorption media,

- shale retorting processes.

Primary emphasis is placed upon those areas pertaining to direct combustion; retorting processes are addressed only insofar as they impact the fuel quality of the spent shale and the solid waste from combustion of spent shale in fluidized-bed combustors.

The major information sources have been several literature references and communications with research organizations and other companies involved in oil shale-related activities. Those groups that have been contacted during the course of this program and their areas of involvement are listed in Table 1. In general, results of investigations being conducted by private firms--either for themselves or clients--are unavailable due to the proprietary nature of such data. On the other hand, work done under government sponsorship is usually published and is readily available.

The current status and technical aspects of utilization of FBC technology for oil shale combustion, perceived benefits and problems, available emission data, and a preliminary economic analysis are presented in Section 2 of this report while a summary and conclusions are provided in Section 3.

BACKGROUND

Deposits of oil shale are found in both the Eastern and Western U.S. as shown in Figure 1.[1] Some of the richest oil shale deposits in the world are found in the Green River formation of Colorado, Utah and Wyoming where beds up to 1500 ft in thickness are found.[2] Total U.S. resources are estimated

TABLE 1. ORGANIZATIONS CONTACTED DURING STUDY RELATIVE
TO OIL SHALE COMBUSTION

	Organization name and location	Individual(s) contacted	Area of investigation[a]
(1)	Argonne National Laboratory Argonne, IL	Irving Johnson	3
(2)	Morgantown Energy Technology Center Morgantown, WV	Joseph S. Mei	1, 3
(3)	Pace Company Denver, CO	Charles Hook Tom Hendrickson	NDI
(4)	Dorr-Oliver, Inc. Stamford, CT	Clarence J. Wall	1
(5)	Lurgi Corp. Eastern Division River Edge, NJ	Ted Polleart	1
(6)	Institute of Gas Technology Chicago, IL	R. David Matthews	NDI
(7)	Babcock and Wilcox Research & Development Division Alliance, OH	J. B. Doyle	1, 3
(8)	Hazen Research, Inc. Golden, CO	Rod Hodgson	1, 2
(9)	U.S. EPA - Region VIII Denver, CO	Mike Hammer	NDI
(10)	Denver Research Institute Chemical Division Denver, CO	Andrew Jovanovich	NDI
(11)	State of Wyoming Air Quality Division Lander, WY	Lee Gribb	NDI
(12)	U.S. EPA Environmental Monitoring Systems Laboratory Las Vegas, NV	Wesley Kinney	NDI
(13)	Bartlesville Energy Research Center Bartlesville, OK	Mr. Linville	NDI
(14)	TOSCO Denver, CO	Dan Fratello Warren Broman	1, 2

(continued)

TABLE 1 (continued)

Organization name and location	Individual(s) contacted	Area of investigation[a]
(15) Foster-Wheeler Development Corp. Livingston, NJ	Joseph F. McMahon	1, 2
(16) Science Applications, Inc. Golden, CO	Dr. Chang Yul Cha	1
(17) Occidental Petroleum Co. Grand Junction, CO	Mr. Davis	NDI
(18) Southern Indiana Shale Oil, Inc. Charlestown, IN	C. R. Thomas	NDI
(19) Union Oil Co. Grand Junction, CO Brea, CA (R&D)	Mr. Heckel Arnold Kelley	NDI NDI
(20) U.S. Geological Survey Grand Junction, CO	Donald Johnson	NDI
(21) Superior Oil Co. Engelwood, CO	Loren Young	NDI
(22) Paraho Corp. Grand Junction, CO	Joseph Glassett	NDI
(23) Dept. of Energy Laramie, WY	Bill Little	NDI
(24) Massachusetts Institute of Technology Cambridge, MA	Prof. Wei Dr. Tong	NDI
(25) Energy Resources Co. Cambridge, MA	Robert Davis	2

[a] 1 = Raw Shale Combustion
2 = Spent Shale Combustion
3 = Coal/Shale Mixtures

NDl = Not directly investigating

Legend:

▨ Tertiary deposits: Green River formation in Colorado, Utah, and Wyoming; Monterey formation in California; middle Tertiary deposits in Montana. Black areas are known high-grade deposits.

▨ Mesozoic deposits: Marine shale in Alaska. Boundary dashed where concealed or where location is unknown.

▤ Permian deposits: Phosphoria formation in Montana.

▧ Devonian and Mississippian deposits. Boundary dashed where concealed or where location unknown. Shading indicates known areas of oil shale deposits; white indicates areas of possible scattered oil shale deposits (extent not defined).

SOURCE: Duncan, D. C., and V. E. Swanson. Organic-Rich Shale of the United States and World Land Areas. USGS Circular 523. Washington DC, U.S. Geological Survey, 1965.

Figure 1. Principal oil shale deposits in the United States.

at over 2 trillion barrels of equivalent oil with about 1.8 trillion barrels in the Green River formation. Not all of this oil is considered to be a recoverable reserve due to current technology and economic conditions, but the U.S. Geological Survey estimates that approximately 600 billion barrels are potentially recoverable from the rich (those containing at least 25 gal of oil per ton of shale) Green River deposits. Of these high grade deposits, approximately 85 percent are located in Colorado's Piceance Creek Basin, 10 percent in Utah's Uinta Basin, and 5 percent in the Green River, Washakie, and San Wash Basins of Wyoming. The estimate of 600 billion barrels represents about a 90-year U.S. oil supply.

The oil shale in the Eastern U.S. is spread over a much larger area than Western shale, but is of much lower quality, as determined by the Fisher Assay test,* mostly being less than 15 gal/ton. These Eastern deposits are estimated at approximately 400 billion barrels of equivalent oil in place, but until recently have been considered nonrecoverable. However, ongoing research at the Institute of Gas Technology in Chicago has involved a retorting method utilizing hydrogen under pressure (hydroretorting) which can potentially increase the shale's yield by a factor of 2.5.[3] This would mean potentially recoverable reserves of about 1 trillion barrels if this technology could be applied successfully to these Eastern deposits.

Optimistic production estimates for an emerging oil shale industry range as high as 3 million bbl/day by the year 2000. A level of 0.5 million bbl/day is probably more realistic. These figures can be compared to the 7 to 8 million bbl/day that the U.S. now imports from foreign countries.

Assuming that an oil shale industry capable of producing 1 million bbl/day eventually becomes established, 20 commercial plants would be required since it is generally assumed that each commercial plant would be capable of producing 50,000 bbl/day. The amount of raw shale that would need to be mined and retorted is equivalent to 1 to 2 tons per barrel of oil, depending upon the quality of the shale. Since only 10 to 15 percent by weight of shale is recoverable as oil, 1 to 2 tons of spent shale is also generated for each barrel produced. For example, a surface retorting operation producing 50,000 bbl/day from high grade shale would use 60,000 ton/day of shale containing 35 gal/ton and would produce 51,000 ton/day of spent shale requiring disposal. It is seen why spent shale disposal is one of the most serious environmental issues facing this industry. Calculations indicate that from 63 to 250 fluid bed combustors would be required for each 50,000 bbl/day plant if all of the spent shale were to be directly combusted (see Appendix A for calculations). Since this is obviously unfeasible, a more realistic approach would be the direct combustion of some fraction of the plant's spent shale for the purpose of recovering energy for use at the facility.

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*In a Fischer assay, small samples of crushed oil shale are heated to 932°F under carefully controlled conditions. The oil yield by this method is the standard measure of oil shale quality.

SECTION 2

TECHNOLOGY OVERVIEW

This section of the report discusses technical aspects associated with direct combustion of oil shale in an atmospheric fluidized bed unit. A discussion is presented first that describes previous usage or experience with FBC technology in the oil shale industry. Next, various problems or concerns with respect to oil shale combustion that are considered to be unique to this application are addressed. Following these brief discussions, past and on-going investigations of the feasibility of direct combustion (raw or virgin shale, spent or retorted shale, and coal-shale mixtures) are summarized. Finally, a preliminary economic analysis of direct combustion of coal-shale mixtures is presented.

PREVIOUS EXPERIENCE WITH FBC TECHNOLOGY

The concept of fluidized bed technology in the oil shale industry is not new, although past uses have primarily related to retorting process applications. During the late 1960s, Sinclair Oil Corporation (now part of Atlantic Richfield Company) investigated FBC retort applications.[4] Brief summaries of several U.S. patents issued since that time include the following:

- U.S. Patent No. 3,484,364, Dec. 16, 1969
  to: Esso Research and Engineering Co.
  Fluidized retorting of oil shale by contact of the fluidized oil shale with a hot free hydrogen-containing gas.[5]

- U.S. Patent No. 3,925,190, Dec. 9, 1975
  to: The Oil Shale Corp. (TOSCO)
  Preheating of oil shale prior to pyrolysis in a series of at least two dilute phase fluidized beds.[6]

- U.S. Patent No. 4,075,081, Feb. 21, 1978
  to: Cities Service Co.
  Fluidized bed hydroretorting of oil shale.[7]

- U.S. Patent No. 4,137,053, Jan. 30, 1979
  to: Chevron Research Co.
  Fluidized bed retorting or gasification of hydrocarbon-containing solids such as oil shale, tar sands, etc.[8]

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- U.S. Patent No. 4,152,245, May 1, 1979
  to: Atlantic Richfield Co.
  Fluidized bed classifier for removing rock solids
  from heat-carrier solids which are cycled in the
  retorting process.[9]

UNIQUE CONCERNS OF OIL SHALE COMBUSTION

Heat loss due to calcium carbonate decomposition, determining the optimum shale size for combustion, and achieving high combustion efficiency are three areas that require specific attention when evaluating oil shale combustion processes.

Carbonate calcination is a potentially troublesome aspect of raw oil shale combustion since oil shale contains appreciable amounts of limestone and because the calcination (forward) reaction,

$$CaCO_3 \rightleftharpoons CaO + CO_2 \text{ (g)}$$
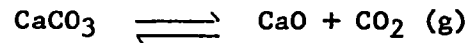
is endothermic, absorbing 43 kcal/g-mole or from 650 to 750 Btu/lb of reactant depending on the temperature. At room temperature, the reverse reaction occurs almost exclusively while the forward reaction becomes more favorable as the temperature is increased. The commercial production of CaO is accomplished at 1650°F[10] and therefore combustion of virgin oil shale solely for heat recovery should be carried out at less than this temperature to avoid this heat consuming reaction. Depending upon the quality of the shale, heat loss through calcination can consume on the order of one-third to two-thirds of the available energy content of the shale.

The size of the as-fired shale is also an important consideration. Specific conclusions drawn by DOE regarding an optimum oil shale size for combustion were as follows:[11]

- Excessive size reduction will cause the release of volatile hydrocarbons trapped inside the shale; this is known to be one of the causes of volatile hydrocarbon loss.

- Fines generated by size reduction operations combined with excessive moisture can cause solid handling problems. This is particularly true for pneumatic air feeding systems.

- Power consumption for oil shale size reduction can be excessive. Inefficient application of power to oil shale size reduction can be economically unattractive, and can lead to operational problems.

A third concern regarding shale combustion is that the achievement of suitable combustion efficiencies may be more difficult than with other fuels. This would be partially due to the fact that small amounts of residual carbon (1 to 3 percent) dispersed throughout a relatively large particle may not burn satisfactorily since mass transfer of oxygen to the carbon that is dispersed in the mineral substrate would be difficult.

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An additional area that has not yet been adequately addressed through on-going research concerns the characterization of solid waste from direct combustion of oil shale in an FBC unit. It would be important to know, for example, whether or not AFBC technology represents a step forward in coping with ultimate disposal of wastes from oil shale extraction processes. Data that are currently available pertain to raw and retorted shale but not to such shale that has been subjected to direct combustion. Table 2 reports results of laboratory-based leachate studies on surface storage of raw and retorted shale conducted at Colorado State University.[12] These data provide some perspective as to what constituents would be present in the feed to an FBC unit burning spent or raw shale. In general, such material contains significant quantities of total dissolved solids, sulfates, and bicarbonates, such inorganic ions as calcium, magnesium, sodium, potassium, and silicon, and lesser amounts of trace metals and organic compounds. Research presently underway at B&W will attempt to assess the leachable character of solid waste from direct combustion of oil shale in their pilot facility.*

Past and on-going investigations have and are attempting to resolve some of the questions regarding calcination heat loss, optimum oil shale size, combustion efficiency, and solid waste characterization. These investigations are discussed in the following paragraphs in order of industry-wide activity level.

RAW SHALE COMBUSTION

Direct combustion of oil shale is a possible alternative or adjunct to oil recovery via retorting. There are only three known commercial applications in the world at the present time. The largest is in the Soviet Union (Estonia), where a rich oil shale having a heat content of 2700 Btu/lb is used directly for fuel in a 3200 MW power plant. A second unit is located in Southern Germany. In the late 1950's, Lurgi's roasting technology was applied industrially in the Lurgi-Rohrbach process for burning oil shale in the city of Dotternhausen.[13] In this application, the economics of the process are favorably influenced by the fact that the cinders are suited for the production of special grade cement. This plant has a capacity of 794 ton/day and has an electrical output of 6 MW. A third installation exists in China at the Mouming Petroleum Company located in Kwang-tung. The boiler there is rated at 32,000 lb/hr steam and burns oil shale fines with a heating value of 1,860 Btu/lb. This unit, commissioned in 1965, was the first demonstration fluidized-bed combustion boiler in China. It has a circular bed of 7.4 ft diameter and a superficial air velocity of 8.8 ft/sec. Bed temperature is 1470°F.[14] This unit is presently being shut down for repair.†

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*Personal communication between Doug Roeck, GCA and John Doyle, Babcock and Wilcox, October 28, 1981.

†Personal communication between Doug Roeck, GCA and Jerry Shang, U.S. Department of Energy, Morgantown Energy Technology Center, November 4, 1981.

TABLE 2. SUMMARY OF THE RANGE OF OBSERVED CONCENTRATIONS
OF ELEMENTS IN RAW AND SPENT SHALE LEACHATE[a]

Parameter	Units	Colony raw shale	Union natural retorted shale
Al	mg/l	<0.05-0.75	<0.05-0.1
As	mg/l	<0.005	<0.005
B	mg/l	<0.025-2.75	0.165-0.39
Ba	mg/l	0.07-0.48	0.028-0.35
Be	mg/l	<0.025	<0.025
Ca	mg/l	40-1550	10-49
Cl	mg/l	1.1-22	1.1-15
CO_3	mg/l	0.03-1.6	<1.0-22
Cr	mg/l	<0.025-0.04	<0.025-0.71
Cu	mg/l	<0.025-0.41	<0.025-0.33
EC	μmhos/cm	240-5400	300-1300
F	mg/l	4.0-7.2	5.4-6.6
Fe	mg/l	<0.03-0.89	<0.05
HCO_3	mg/l	50-558	168-585
Hg	mg/l	<0.0005	<0.0001
K	mg/l	1.7-59	9.2-74
Li	mg/l	0.02-0.151	0.14-0.51
Mg	mg/l	5.5-140	30-108
Mn	mg/l	0.074-2.74	<0.05
Mo	mg/l	0.09-0.65	0.065-0.45
Na	mg/l	5.8-145	12-75
Ni	mg/l	<0.05-0.10	<0.05
NO_3	mg/l	0.9-25	0.5-8
Pb	mg/l	<0.05-0.64	0.05-0.16
pH	-	7.06-8.18	7.2-8.81
Se	mg/l	<0.01	<0.005
Si	mg/l	2.12-10.58	8.8-19.1
Sn	mg/l	0.12-0.67	-
SO_4	mg/l	28-5150	9-128
TDS	mg/l	110-7160	460-1200
Zn	mg/l	<0.02-0.68	0.02-0.15

[a]Source: Reference 12.

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In the United States, there are no known commercial applications, although research is being conducted by several companies and government agencies. The U.S. Department of Energy (DOE) has conducted tests on oil shale from Colorado, Israel, and Morocco at their Morgantown Energy Technology Center (METC). Babcock and Wilcox (B&W), Research and Development Division, has done combustion tests on oil shale from Colorado in their Alliance, Ohio experimental facility. Other companies involved in raw shale combustion investigations, some of which cannot release data because of confidentiality, include Dorr-Oliver, Inc., Lurgi Corporation-Eastern Division, Hazen Research, Inc., Foster-Wheeler Development Corp., and Science Applications, Inc. (see Table 1). Published results from DOE, B&W, and Foster-Wheeler are described in the following paragraphs.

Department of Energy Experiments[11]

In September 1978, oil shale from Colorado was successfully burned in the METC's 18-inch diameter fluidized bed combustor. In July 1979, similar testing was done on Israeli oil shale in both the 6-inch and 18-inch FBC units. During 1980, DOE conducted additional tests on Moroccan oil shale in the 6-inch unit.

The 6-inch laboratory-scale FBC is a refractory-lined cylindrical vessel with a conical air distributor plate that has an open area equal to 2.8 percent of the total distributor area. The bed region is 16 inches high and the expanded freeboard section is 8 inches in diameter and 64 inches high. Fluidizing air from the plenum passes through 64, 1/8-inch diameter nozzles arranged in three concentric circles. Fuel from the hopper is fed directly into the bed by a 1-inch, variable-speed metering screw which is cooled by a water jacket.

The 18-inch FBC unit is 45 inches high with an expanded freeboard section 24 inches in diameter and 27 inches high. The combustor is equipped with horizontal water or steam cooled heat exchangers submerged in the bed, comprised of six hairpin configuration, 1/2-inch, 316 stainless steel tubes, any one of which can be placed in service. The freeboard area is equipped with a single-pass heat exchanger used to provide heat transfer data and to control exhaust gas temperature. Fuel is metered from a hopper which is suspended from a weigh cell by a 1.1875-inch variable speed screw. Feed rates of 280 lb/hr were used with the oil shales. Bed temperatures are maintained by varying the bed level or fuel feed rate and through manipulation of the bed heat exchangers.

Two different processes were explored in conducting the preliminary oil shale combustion studies:

● The oil shales were combusted intensively for maximum combustion efficiency with the objective of producing process heat for power generation. (Test results reported subsequently were generated while operating in this mode).

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- The oil shales were heated in a fluidized bed where partial combustion was allowed to occur for the purpose of driving off the hydrocarbons for fuels recovery. Since the current METC facilities do not provide for recovery of hydrocarbon vapors, these vapors were burned in the over-bed space.

Colorado Oil Shale--[11]

The Colorado oil shale combusted at METC came from the Piceance Creek Basin section of the Green River formation and yielded a Fisher Assay from 10 to 25 gal/ton. The purpose of the combustion studies was to explore the feasibility of direct combustion of Colorado oil shale in an AFBC and to determine combustion characteristics by studying combustion efficiency, sulfur retention, pollutant emissions, and extent of carbonate calcination. (The Colorado shale contained about 18 percent by weight limestone).

The combustion of the Colorado shale led to the following conclusions:

- NO_x emissions were mainly a function of air-to-oil shale ratio; when this ratio exceeded 2.25 lb air/lb oil shale, NO_x emissions exceeded 0.6 lb/10^6 Btu. (The percent excess air during these tests ranged from 33 to 66).

- SO_x emissions were not affected by the air-to-oil shale ratio and were mostly less than 0.2 lb/10^6 Btu.

- Percent sulfur retention was found to be a function of bed temperature and the percent of limestone calcined. High sulfur retention (90 to 95 percent) was achieved with about 60 percent of the limestone calcined at a bed temperature of 1300° to 1350°F. At a bed temperature of 1500°F, with 90 percent of the limestone calcined, sulfur retention was below 90 percent.

- Combustion efficiency for the Colorado shale was found to be a function of both air-to-oil shale ratio and bed temperature. Greater than 99 percent efficiency was attained at an air-to-oil shale ratio of 2.0 to 2.5 and a bed temperature range of 1300° to 1500°F.

Israeli Oil Shale--[11,15]

DOE has also test burned Israeli oil shale in the 6- and 18-inch combustors. This shale came from the Efe deposit discovered in 1962 during a survey of phosphorite in that area. The ultimate and ash analyses for this shale are given in Table 3. The volatile matter and calcium carbonate contents are 38 and 47 percent, respectively. Based on the ash analysis, this oil shale has an inherent calcium to sulfur molar ratio of about 7.3.

The operating conditions explored during combustion of the Israel oil shale in the 6- and 18-inch diameter units are shown in Table 4. The lowest operating temperature occurred when burning shale with 18 percent moisture without preheated air. The highest bed temperature, 1490°F, was attained when the oil shale contained only 5 percent moisture and was burned in the

TABLE 3. CHEMICAL ANALYSIS OF ISRAELI OIL SHALE[a]

Component (%)	Shale fed to 6 in. FBC	Shale fed to 18 in. FBC
Ultimate Analysis:		
Moisture	7.93	8.15
Ash	55.28	56.93
Sulfur	2.43	2.35
Hydrogen	1.45	2.05
Nitrogen	0.44	0.29
Total carbon	14.69[b]	14.94[c]
Oxygen	17.78	15.29
Carbonate as CO_2	20.80	22.36
Heating value, Btu/lb	1,902	1,893
Volatile matter	38.0	
Calcium carbonate	47.0	
Ash Analysis:		
Silicon, SiO_2	20.85	23.17
Aluminum, Al_2O_3	8.27	9.24
Iron, Fe_2O_3	3.66	4.48
Calcium, CaO	55.04	51.85
Magnesium, MgO	0.68	1.17
Sodium, Na_2O	2.46	0.50
Potassium, K_2O	0.52	0.62
Phosphorus, P_2O_5	0.37	2.42
Titanium, TiO_2	-	0.35
Sulfur, SO_3	6.56	5.99

[a]Source: Reference 15.

[b]Carbon (CO_2) = 5.62%, Organic carbon = 9.07%.

[c]Carbon (CO_2) = 6.10%, Organic carbon = 8.84%

6-inch unit with air preheated to 600°F. Carbon combustion efficiencies of 98 percent or better were attained in the 6-inch combustion unit without recycling of fines while efficiencies of 96 percent or better were achieved in the 18-inch unit with recycling. The higher combustion efficiencies for the laboratory-scale unit were attributed to that unit's higher freeboard section which provided for a longer gas residence time. This was confirmed by the much lower carbon monoxide and hydrocarbon concentrations detected in the flue gases. Heat losses due to calcium carbonate decomposition were minimized by maintaining low bed temperatures. Calcium carbonate decomposition was found to range from a low of 28 percent at 1180°F to a high of 86 percent at 1490°F.

TABLE 4. OPERATING CONDITIONS FOR FLUIDIZED-BED COMBUSTION OF ISRAELI OIL SHALE[a]

Parameters	Combustion unit	
	6-inch diameter	18-inch diameter
Moisture content, %	2.0 - 18.5	8.0 - 16.0
Feed size, in. or mesh	1/8" x 0 & 14 mesh	1/4" x 0
Fuel feed rate, lb/hr	18.5 - 52	180 - 300
Excess air, %	12 - 50	28 - 72
Preheated air temp. °F	70 - 675	70
Bed temperature, °F	1180 - 1490	1250 - 1400
Bed depth, in.	12	36
Superficial velocity, ft/sec	2.5 - 6.0	2.8 - 4.5

[a]Source: Reference 15.

Because of the high ash, moisture, and calcium carbonate contents of the Israeli oil shale, large quantities were required in order to maintain the desired bed temperature. Such handling of the feed material and the spent bed would be one of the specific operating problems that would need to be considered in designing any commercial unit.

Some of the specific conclusions drawn from these combustion tests were as follows:

- Devolatilization of the Israeli oil shale was completed at a temperature of 1076°F.

- Emissions of NOx and SO_2 from the 18-inch unit were low (0.4 lb $NO_x/10^6$ Btu and 0.7 lb $SO_2/10^6$ Btu).

- Overall heat transfer coefficients of 53.4 and 7.5 $Btu/ft^2-hr-°F$ were obtained for in-bed and freeboard heat exchange tubes in the 18-inch combustor.

- A hybrid combustion system would be the best means for the direct combustion of Israeli oil shale: devolatilization would take place in the main section of the fluidized bed reactor operated at substoichiometric conditions and at a bed temperature less than 1300°F; subsequent combustion of volatiles would be carried out in the freeboard section. Heat in the spent bed material could be partially recovered through a moving-bed cooler for preheating the fluidizing air.

Moroccan Oil Shale[11]--

Limited tests were conducted on Moroccan oil shale in DOE's 6-inch combustion unit. This shale had a heating value of 2400 to 2700 Btu/lb and a sulfur content of ~2.3 percent. The ultimate analysis showed about 62 percent ash, from 4 to 7 percent moisture, and about 16 percent total carbon. Experimental results from burning this shale are shown in Table 5.

B&W Experiments[16]

The Research and Development Division of Babcock and Wilcox in Alliance, Ohio has performed various tests on both raw shale and coal-shale mixtures (these latter tests are discussed further on in this section). The pilot facility used is referred to as a 1 ft x 1 ft fluid bed combustion unit and is 13-1/2 inches square by 16 ft high. The unit is equipped with an in-bed heat exchanger and a high efficiency cyclone separator. A gas burner is used to preheat the combustion zone prior to the introduction of solid fuels. Fuel and reagent feed rates are independently controlled by variable speed screws which discharge into a common chute through which material falls by gravity to a rotary valve. At the rotary valve discharge point, material is pneumatically picked up by an air eductor and sent to the furnace at sufficient pressure to overcome furnace resistance.

The oil shale used in the B&W test program was obtained from the Occidental Petroleum mine located in the Debeque, Colorado area. This shale had a heating value of about 1000 Btu/lb (dry basis), a Fisher-Assay of 10 gal/ton, an ash content of about 71 percent and a sulfur content of 0.2 percent. Planned and actual test conditions are provided in Table 6 while experimental results are shown in Table 7.

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TABLE 5.   TEST RESULTS DURING COMBUSTION OF MOROCCAN OIL SHALE[a]

Operating Conditions

Bed temperature, °F	1250 to 1600
Excess air, % vol.	Stoichiometric to 85
Superficial velocity, ft/sec	2.0 to 4.0
Bed Ht. (static), inches	4 to 6
Fuel size, mesh	12 x 0
Ca/S molar ratio	~4 to 5

Results

	Run No.		
	2	3	4
Combustion efficiency,[b] %	99.24	98.79	97.91
$CaCO_3$ calcination, %	77.62	75.50	79.71
Heat loss through calcination,[c] %	3.45	3.52	4.81
Flue Gas Analysis:			
CO_2, % vol.	12.5	14.5	14.0
CO, % vol.	0.05	0.93	0.14
O_2, % vol.	7.0	0.0	6.25
SO_2, ppm	100	100	100
SO_2, $lb/10^6$ Btu	0.27	0.33	0.29
NO/NO_x, ppm	800	900	700
NO_x, $lb/10^6$ Btu	1.09	0.92	1.0
THC, ppm	100	4000	100
Bed temperature, °F	1400	1400	1450
Fluidization velocity, ft/sec	2.84	1.97	3.96
Bed depth (static), inches	6	6	4
Excess air, %	47.73	0.0	40.0

[a]Source: Reference 11.

[b]Based on solid analyses.

[c](Heat loss/total heat input) x 100.

TABLE 6. B&W TEST CONDITIONS - 100 PERCENT OIL SHALE[a]

	Test No.					
	4		5		6	
	Planned	Actual	Planned	Actual	Planned	Actual
Oil shale size	1/4 x 0	1/4 x 0	1/4 x 0	1/4 x 0	1/4 x 0	1/4 x 0
Superficial velocity, ft/sec	7 to 8	7.5	7 to 8	7.4	7 to 8	7.2
Bed depth, in.	30 to 36	33	30 to 36	39	30 to 36	33
Flue gas O_2, %	2.8 to 3.5	2.8	2.8 to 3.5	3.0	2.8 to 3.5	1.4
Windbox air temp., °F	400	432	400	443	400	424
Transport air, lb/hr	–	108	–	108	–	106

[a]Source: Reference 16.

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TABLE 7. B&W TEST RESULTS

Test No.	4	5	6
Date	11/3/78	11/3/78	11/3/78
Oil shale, lb/hr	360	355	358
Heat input, 10^6 Btu/hr	0.339	0.335	0.338
Bed temp., °F	1473	1450	1398
Ca/S molar ratio	◄———————	∿47	———————►
SO_2 out, lb/10^6 Btu	0.05	0.05	0.05
SO_2 reduction, %	99	99	99
NO_x out, lb/10^6 Btu	3.0	3.0	2.92

[a]Source: Reference 16.

The goal of the test program--to successfully burn oil shale as a fuel without the need for auxiliary fuel-- was achieved. However, due to the low heat content of the shale, a normal heat input to the unit (0.7 x 10^6 Btu/hr) could not be attained. The B&W tests showed that oil shale could be burned as a primary fuel in a fluid bed combustor without the need for auxiliary fuels to maintain combustion. The major drawback to this mode of operation was the high NO_x emissions (3.0 lb/10^6 Btu). In reviewing the chemical analysis of the fuels, B&W noted that the total theoretical fuel-bound nitrogen contained in the oil shale was approximately double that of the coal tested on a lb/10^6 Btu basis. The percentage of the total fuel-bound nitrogen which appeared in the outlet gas averaged about 42 percent compared to an average of 13 percent for the coal-oil shale mixtures. B&W concluded that the conversion of fuel-bound nitrogen to NO_x for oil shale was similar to that of a typical oil.

Other Investigations

Foster-Wheeler Development Corporation has also performed combustion tests on oil shale in a fluidized bed.[17] Their fluid bed combustor consists of a refractory-lined steel cylinder 7 ft in height, 1.5 ft internal diameter at the bottom, and 2.0 ft internal diameter in the freeboard section. The unit is equipped with two compound-flow multiple-pass tube bundles, one in the bed zone, the other in the convection pass. Products of combustion leave the furnace through a water-cooled stack and then pass through two stages of cyclone separators. Following the cyclones, gases can either be piped through a dilution flow mixer and then into a baghouse filter or directly out of the building. Bed material and fuel are metered from bulk hoppers to a common feed air lock by variable-speed screw feeders. After mixing in the air lock, the bed/fuel mixture is pneumatically conveyed to a vertical feed pipe which discharges material about 3 ft above the bed. While most of Foster-Wheeler's tests were on coke breeze and various grades of coal, one test was conducted with oil shale from Utah. This oil shale contained 73 percent ash, 1.15 percent sulfur, and 22.5 percent carbon. Test results were as follows:

superficial velocity, 8.6 ft/sec; average bed temperature, 1550°F; 14 percent excess air; combustion efficiency (no recycle), 99 percent; and NO_x emissions, 430 ppm (v). No specific conclusions were drawn concerning this one oil shale test run.

Other companies are known to have conducted proprietary tests on the combustion of raw oil shale in fluidized-bed combustors. These include: Dorr-Oliver, Inc. (Stamford, Conn.); Hazen Research, Inc. (Golden, Colo); Lurgi Corp.-Eastern Division (River Edge, N.J.); and Science Applications, Inc. (Golden, Colo.).

SPENT SHALE COMBUSTION

Spent shale refers to oil shale that has been retorted by any one of various processes for recovery of oil. The properties of the retorted or spent shale depend on the raw shale characteristics and on the type of retorting procedure used. Spent shale from an indirectly heated retort will contain an organic carbonaceous residue which remains after pyrolysis of the kerogen—the organic component of shale rock. For example, the indirectly heated TOSCO II retort produces a spent shale resembling black talcum powder. The chemical composition of this spent shale is shown in Table 8.

TABLE 8. CHEMICAL COMPOSITION OF TOSCO II
SPENT SHALE[a]

Component	Weight %
SO_3	2.63
Total carbon	9.83
Inorganic carbon	4.41
Organic carbon	5.41
SiO_2	33.07
Al_2O_3	9.14
Fe_2O_3	3.24
CaO	17.56
MgO	7.74
Na_2O	0.77
K_2O	1.39

[a]Source: Reference 18.

Although more of the carbon residue is consumed in the combustion process, spent shale from directly heated retorts still contains a small percentage of carbon. The directly heated Union "A" retort produces a gray decarbonized spent shale resembling coal ash and clinkers. The composition of the spent shale from Union Oil Co.'s "A", "B" (indirectly heated), and SGR (indirectly heated) retorts is given in Table 9. From a process in which none of the carbon residue is burned in the retorting process, the organic carbon may represent about 140 Btu/lb for every 1 percent of residual organic carbon (based upon a heat of combustion for carbon of 14,087 Btu/lb). While reported values of residual carbon in spent shale from the various retorting processes range from 0.3 (Union SGR) to 5.4 percent (TOSCO II), thus indicating potential heat contents of as much as 750 Btu/lb, actual values may realistically be lower since modifications to indirectly heated retorting processes will attempt to utilize much of this available heat energy. For example, TOSCO has patented processes to burn the retorted shale as fuel for their ball heater while the indirect Lurgi-Ruhrgas process also incorporates integral use of available carbon residue.[19]

TABLE 9. COMPOSITION OF UNION SPENT OIL SHALES[a]

Component	Union A[b]	Union B[c] (in weight %)	Union SGR[c]
SiO_2	35.3	31.5	39.2
CaO	27.2	19.6	27.3
MgO	9	5.7	8.2
Al_2O_3	8.5	6.9	8.9
Fe_2O_3	7.3	2.8	3.8
Na_2O	5.5	2.2	3.7
K_2O	2.8	1.6	2.7
SO_3	0.1	1.9	1.4
P_2O_5	2.2	0.4	0.5
Mineral CO_2	1.6	22.9	3.1
Organic C	0.5	4.3	0.3
Inorganic C	-	-	-
Texture	graded gravel to silty gravel	silty gravel	silty gravel
pH	12.5 - 13.0	8.7	12.5

[a]Source: Reference 19.

[b]Directly heated.

[c]Indirectly heated.

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Four companies are known to be investigating combustion of spent shale; however, no test data are available due to confidentiality. Hazen Research, Inc. (Golden, Colo.) has conducted tests for a fluidized bed combustor manufacturer.* Foster-Wheeler Development Corp. (Livingston, N.J.) has performed spent shale combustion tests and state that they hope eventually to be able to offer industrial size units up to 500,000 to 600,000 lb/hr steam for the burning of shale.† TOSCO (The Oil Shale Corp.--Denver, Colo.) is also investigating direct combustion of spent shale.‡ Because their program was proprietary, it could not be determined if direct combustion in a fluidized-bed combustor was the method being evaluated. Energy Resources Co. (Cambridge, Mass.) has previously conducted spent shale tests in their 20-inch I.D. combustor. These tests were done on spent shale from TOSCO's retorting process and the results showed that this shale could be successfully combusted.§ It would seem reasonable that other companies involved in oil shale projects are also investigating the prospects of spent shale combustion.

COMBUSTION OF COAL-SHALE MIXTURES

Various research efforts have been conducted with respect to the burning of oil shale-coal mixtures for the purpose of SO_2 removal from the coal. Because oil shale contains appreciable amounts of calcium, it is felt that the shale can be used as a replacement for limestone in conventional firing of FBC boilers. One difference between the firing of such mixtures and the direct combustion of spent or virgin shale for pure energy recovery is that bed temperatures are higher in the former case so as to promote the calcination reaction of $CaCO_3$ to CaO. This will enhance SO_2 capture because the calcium oxide is the eventual reactant with the sulfur dioxide pollutant. The following paragraphs describe ongoing and past work in this area.

Testing of coal-oil shale mixtures has been performed at the Research and Development Division of Babcock and Wilcox in Alliance, Ohio. Results of this test program were presented at a recent symposium and are summarized here.[16] The FBC unit operated by B&W was started with Ohio No. 6 coal as the fuel and a bed of spent limestone from a previous test. (No fresh limestone was used in the B&W program described herein). The unit went through a normal startup mode with coal as the fuel source until steady-state conditions were achieved (as determined by a leveling out of the SO_2 outlet concentration). Once steady-state was achieved, a transition was made from coal to oil shale by slowly adding shale while backing off on the coal feed. Bed temperature dropped during this transition due to the low heat content of the shale and

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*Personal communication between Doug Roeck, GCA and Rod Hodgson, Hazen Research, Inc. May 19, 1981.

†Personal communication between Doug Roeck, GCA and Joseph F. McMahon, Foster-Wheeler Development Corp. May 22, 1981.

‡Personal communication between Doug Roeck, GCA and Warren Broman, TOSCO. June 26, 1981.

§Personal communication between Doug Roeck, GCA and Robert Davis, Energy Resources Co. June 29, 1981.

because the quantity of transport air had to be increased to compensate for the higher solids loading in the feed system. To maintain bed temperature near the 1500°F level, the effective heat transfer surface in the bed was reduced.

During the transition from coal to oil shale it was noted that the outlet SO_2 concentration dropped significantly as soon as the oil shale had been added. A series of tests were performed to investigate this phenomenon and are shown in Table 10. Tests 3 and 7-12 were designed to study the effect of the shale on SO_2 reduction. Also shown in this table are two earlier tests (13 and 14) that had been conducted with high sulfur coal and a high grade limestone as the reagent. As can be noted from the results, the oil shale was far superior to the limestone in regards to reacting with sulfur dioxide on a molar basis. The NO_x emissions were equivalent to that of the coal-limestone mixture tests conducted previously. Total carbon conversion of the coal and oil shale was at least equivalent to that of the coal alone. A plot of SO_2 reduction versus Ca/S molar ratio for the coal-oil shale and the coal-limestone tests is shown in Figure 2. Although the total quantity of oil shale required for a given SO_2 reduction is about double that of the limestone requirement, this is partially offset by the fact that the oil shale is a contributor to the system heat input. As a result of these studies, B&W concluded that a coal-oil shale mixture is an attractive fuel for use in a fluid bed combustor; the heat content of the oil shale can be efficiently converted to energy and high sulfur coal can be burned with oil shale to effectively reduce SO_2 emissions.

Other testing of coal-oil shale mixtures has been performed by the Department of Energy at the Morgantown Energy Technology Center (METC) and the Argonne National Laboratory (ANL).

Testing at METC, done in early 1979, involved the burning of Colorado oil shale containing 18 percent limestone and Pittsburgh seam bituminous coal having a sulfur content of 4.5 percent.[11] These tests showed 95+ percent sulfur retention at bed temperatures of 1500° to 1600°F. Observations with an electron microscope revealed that the limestone in the oil shale had "blossomed" into a flower-like fossil structure, thereby making surface area available for SO_2 absorption. The presence of coal was found to reduce the nitrogen oxides emissions to well below 0.6 lb/10[6] Btu, possibly due to reduction of the oxides by the coal char.

Additional testing for DOE has been done at the Argonne National Laboratory (ANL) in Argonne, Illinois.[20-22] These tests were designed to compare virgin and spent Green River oil shale with two limestones (Germany Valley and Greer) and one dolomite (Tymochtee) for SO_2 reactivity and attrition rate. The compositions of these materials are indicated in Table 11. As seen in this table, the virgin oil shale contained 28 percent limestone and the spent shale (prepared by heating virgin shale in air at 750°F for 3 hours) contained 34 percent limestone. Oil shale reactions with SO_2 were studied with a thermogravimetric analyzer (TGA) at temperatures of 1300° to 1920°F. The conversion of CaO to $CaSO_4$ in virgin shale was slightly lower

TABLE 10. TEST RESULTS FROM COAL-OIL SHALE COMBUSTION TESTS CONDUCTED AT BABCOCK & WILCOX, ALLIANCE, OHIO[a]

Test number	1	2	3	7	8	9	10	11	12	13	14
Date	10/27/78	10/27/78	11/01/78	11/30/78	01/11/79	01/11/79	01/31/79	02/02/79	02/06/79	03/07/78	10/05/78
Coal, lb/hr	41	24.5	53	35.6	47.9	50.1	48.6	46.7	51.8	51.0	78.0
Oil shale, lb/hr	130	282	43	23.9	56.6	41.0	37.1	54.1	40.0		
Limestone, lb/hr										26.5	36.0
Heat input, 10^6 Btu/hr	0.626	0.567	0.691	0.468	0.649	0.660	0.647	0.639	0.689	0.612	0.822
Calcium/sulfur	6.9	25.1	1.8	1.6	3.0	2.1	1.7	2.4	1.7	5.6	3.6
lb oil shale/lb coal	3.2	11.5	0.81	0.67	1.11	0.82	0.76	1.16	0.77	0.52	0.46
Bed temp., °F	1605	1565	1554	1596	1530	1571	1552	1558	1589	1525	1555
Bed depth, in.	36	36	36	30	36	36	33	36	36	35	30
Gas velocity, ft/sec	7.5	7.6	7.1	5.9	7.2	7.2	7.2	7.4	7.5	6.6	6.8
SO_2 out, lb/10^6 Btu	0.1	0.1	1.1	1.3	0.8	1.3	1.6	1.0	2.1	b	1.3
SO_2 reduction, %	98	99	82	78	86	78	74	84	66	82	80
NO_x out, lb/10^6 Btu	-	-	0.38	0.49	0.44	0.38	0.39	0.38	0.35	-	0.4
Carbon conversion, %	-	-	93.0	94.5	93.7	89.6	88.7	88.5	99.1	84.0	93.0

[a]Source: Reference 16.

[b]SO_2 outlet concentration was measured but not recorded.

NOTES:

(1) Tests 1-3 and 7-12--with Ohio No. 6 coal,
     12,000 Btu/lb, 3.6% S and Colorado oil shale,
     900 to 1000 Btu/lb, 0.2% S.

(2) Test 13--with Pittsburgh No. 8 coal, 12,800 Btu/lb,
     3.0% S and Lowellville limestone, 92% $CaCO_3$

(3) Test 14--with Illinois No. 10 coal, 10,500 Btu/lb,
     4.5% S and Lowellville limestone, 92% $CaCO_3$.

22
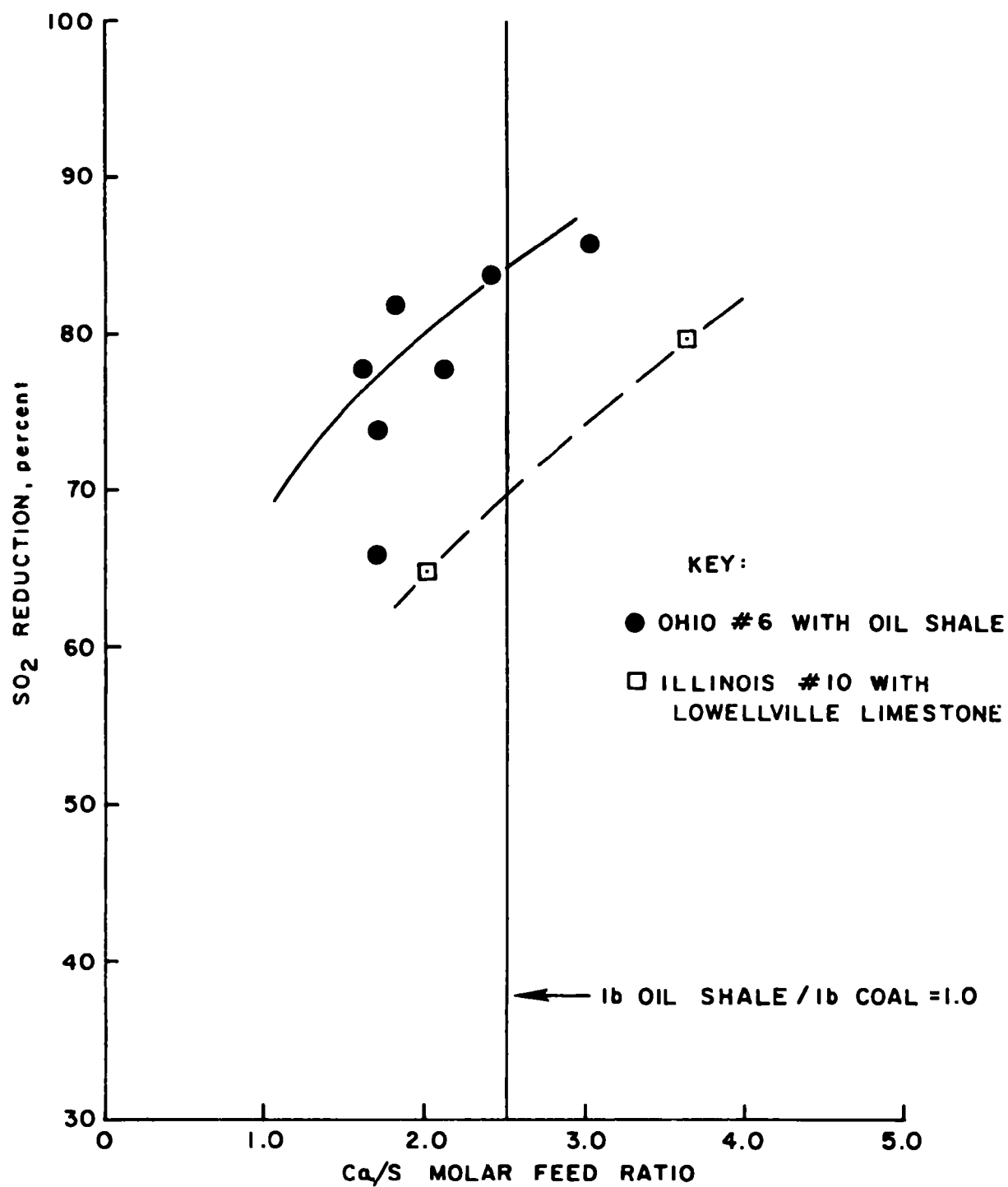
Figure 2. Sulfur dioxide reduction for coal-limestone and coal-oil shale mixtures.

Source: Reference 16.

(73 percent) than for the spent shale (86 percent). A plot of this conversion rate for spent shale is shown in Figure 3.

TABLE 11. CONCENTRATIONS (in wt %) OF MAJOR CONSTITUENTS OF CALCAREOUS MATERIALS[a]

	$CaCO_3$	$MgCO_3$	Fe_2O_3	Al_2O_3	SiO_2	Na_2O	K_2O	S
Virgin Oil Shale	27.7	13.8	2.60	4.14	30.1	2.19	1.21	0.89
Spent Oil Shale	33.5	16.7	3.14	5.0	36.4	2.64	1.46	0.66
Tymochtee Dolomite	51.8	43.3	0.41	1.46	3.61	0.07	–	–
Greer Limestone	80.4	3.50	1.24	3.18	10.34	2.23	–	–
Germany Valley Limestone	97.75	0.6	0.1	1.8	0.2	0.25	–	–

[a]Source: Reference 21.

The inverse variation of reaction rate with temperature shown in Figure 3 was attributed to other competing reactions at the higher temperatures. However, the shale sulfated between 700° and 800°C (1300° and 1500°F) resulted in the formation of a complex, highly water-soluble salt that might pose a waste disposal problem. Therefore, the experimental kinetic data at 900°C (1650°F) were later used to estimate the quantity of oil shale necessary to meet an SO_2 emission standard of 1.2 lb/10^6 Btu. The comparison of spent shale, dolomite and the two limestones at this temperature is shown in Figure 4. It is seen that although during the first hour of reaction the percent of Ca changed to $CaSO_4$ was higher for the oil shale than for the highly reactive Tymochtee dolomite, by the end of the 3 hours, 98 percent of the CaO in the dolomite had been utilized to capture SO_2 compared with 84 percent for the spent shale. The reaction rates for the two limestones were much lower than those for the shale and dolomite.

The SO_2 reactivity data were converted to SO_2 retention versus Ca/S ratio for prediction of the quantity of calcareous material necessary per unit of coal to meet the SO_2 emission standard. Figure 5 shows these results at the different temperatures studied.

The results of these analyses in terms of quantities of materials required are given in Table 12. The major conclusions of the ANL studies were as follows:

24
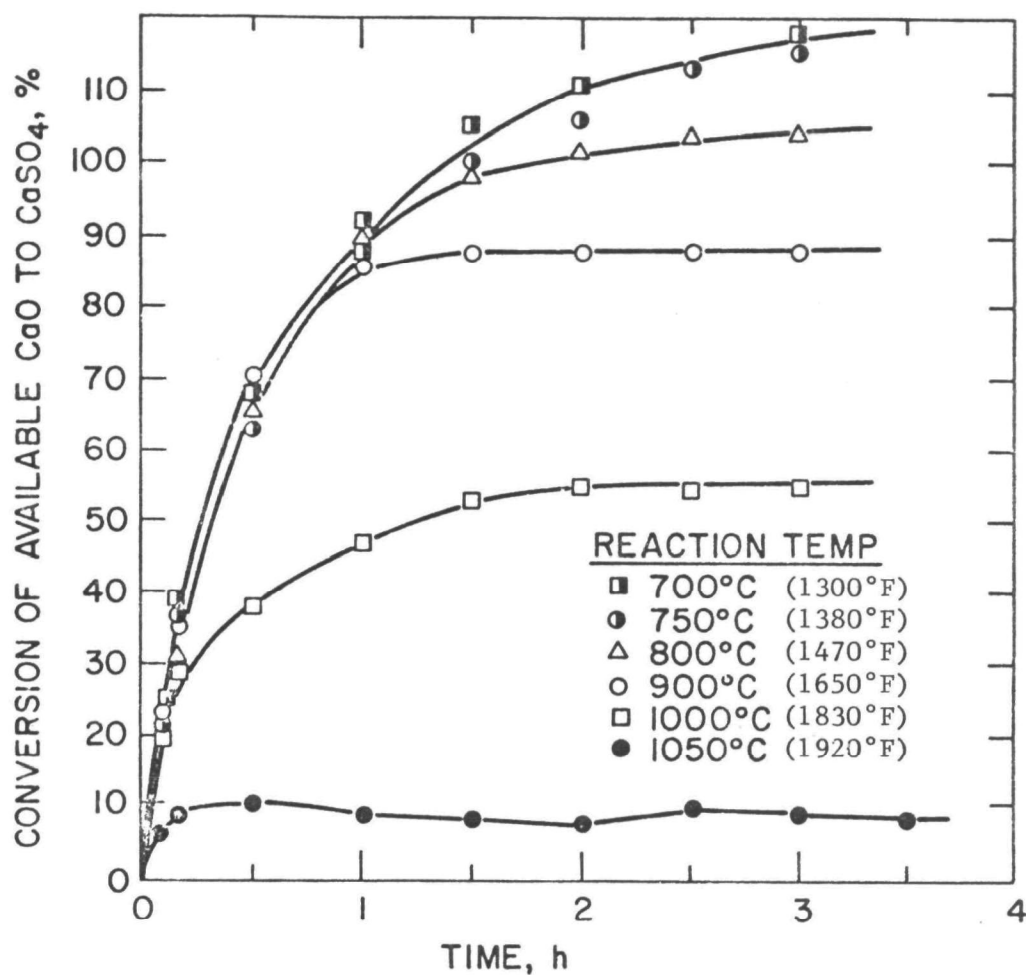
Figure 3. Conversion (measured with a TGA) of Precalcined CaO to CaSO$_4$ in Spent Green River Oil Shale at 700° to 1050°C (1300° to 1920°F). Reaction conditions: precalcined, 20% CO$_2$-bal N$_2$ (-50 +70 mesh); sulfation gas, 0.3% SO$_2$-5% O$_2$-20% CO$_2$ in N$_2$.

Source: Reference 20.

25
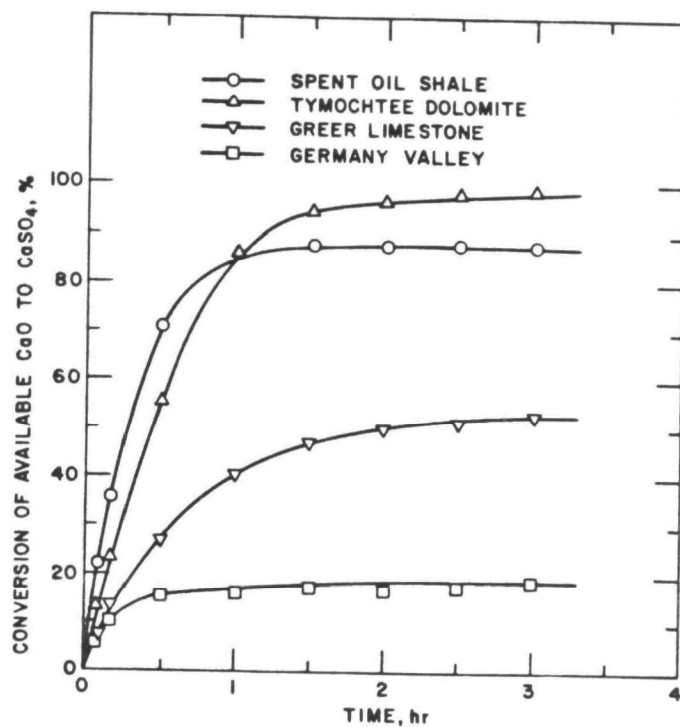
Figure 4. Conversion (measured with a TGA) of CaO to $CaSO_4$ in Spent Oil Shale, Tymochtee Dolomite, Greer Limestone, and Germany Valley Limestone, using 0.3% SO_2-5% O_2 in N_2 at 900°C (1650°F).

Source: Reference 21.
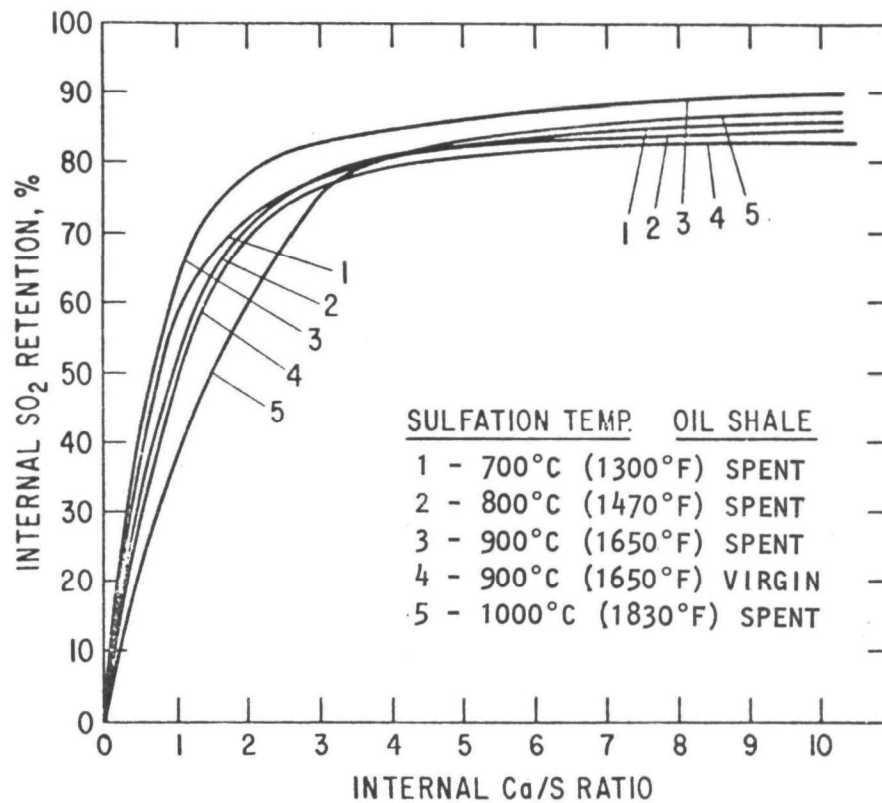
Figure 5. Predicted SO_2 retention vs. Ca/S ratio for Oil Shale at various temperatures for a gas velocity of 12.5 ft/sec and a bed height of 3.5 ft. Based on TGA data.

Source: Reference 20.

TABLE 12. REQUIRED QUANTITIES OF GREEN RIVER OIL SHALE, GERMANY VALLEY LIMESTONE, GREER LIMESTONE, AND TYMOCHTEE DOLOMITE TO MEET AN SO_2 STANDARD OF 1.2 $lb/10^6$ Btu[a]

Calcium-based stone	Ca/S ratio required				Unit wt. stone/unit wt. coal required			
	1.23% S	2.0% S	3.0% S	4.3% S	1.23% S	2.0% S	3.0% S	4.3% S
Spent oil shale[b]	0.5	1.2	1.9	not possible	0.06	0.20	0.58	not possible
Virgin oil shale[c]	0.6	1.1	3.1	4.3	0.10	0.25	1.4	3.3
Germany Valley limestone	1.3	2.5	3.8	7.5	0.05	0.16	0.36	1.0
Greer limestone	1.6	2.8	3.1	3.8	0.08	0.23	0.36	0.6
Tymochtee dolomite	0.4	0.7	1.0	I.5	0.03	0.03	0.18	0.4

[a] Basis: 12,183 Btu/lb coal. FBC operates at 12.5 ft/sec and 3.5 ft bed depth.

[b] Spent shale contains 0.66% S.

[c] Virgin shale contains 0.9% S and has a heating value of 3,020 Btu/lb.

Source: Reference 22.

- more oil shale than either limestone or dolomite would be required for equivalent SO_2 reduction

- the use of oil shale as an SO_2 sorbent may be desirable if the FBC is operated at low gas velocity (<12.5 ft/sec) or with low sulfur coals (<3 percent)

- Green River oil shale should not be used in an FBC that employs a carbon burnup cell (CBC) because the high silica content (~30 percent) of this shale would lead to SO_2 being released at the high temperatures (1800° to 2000°F) encountered in a CBC

- the attrition rate of virgin oil shale was found to be similar to that of the dolomite and limestone tested.

It was noted by the ANL investigators that their results apply specifically to atmospheric FBCs operated at a high superficial gas velocity. Decreasing the velocity and/or increasing the combustion chamber pressure may decrease shale requirements considerably. In addition, only Green River oil shale was used in the test program and since SO_2 reactivity and attrition rates of limestones vary widely, a large variation would also be expected for shales of differing compositions.

PRELIMINARY ECONOMIC ANALYSIS

Of the three options considered in the previous discussions (combustion of raw and spent shale, and coal-shale mixtures), the most viable alternative would seem to be the use of oil shale as a sorbent with coal. Therefore, an economic analysis has been performed so that operating costs for such coal-oil shale firing can be compared to a conventional FBC system using limestone sorbent. The base case for this analysis is a DOE study[23] which assessed the costs for a 250,000 lb/hr (steam) FBC unit. The coal used for this study was a mid-western bituminous coal with a high heating value of 10,430 Btu/lb, a sulfur content of 3.5 percent, and an ash content of 9.2 percent. Based upon a coal firing rate of 34,012 lb/hr, the heat input is 355 x 10^{6} Btu/hr. The system is designed to meet an SO_2 emission standard of 1.2 lb/10[6] Btu input plus 85 percent reduction and a particulate emission standard of 0.03 lb/10[6] Btu input plus 99 percent reduction and 20 percent opacity.

The costs determined by DOE for the FBC system are shown in Table 13. For the purposes of determining costs for comparison with the $4.28/1000 lb steam direct operating cost figure shown in Table 13, three sets of oil shale-coal test data summarized in this report were chosen for economic analysis (one developed by B&W and two determined by thermogravimetric analysis at Argonne National Laboratory). These three sets of data were discussed previously in conjunction with Table 10 (Test No. 11) and Table 12. (These data are also described and expounded upon more fully in Section 3, Table 16b-- B&W Test No. 11 and ANL Test No's. 17 and 20).

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TABLE 13.   DIRECT OPERATING COSTS FOR A CONVENTIONAL
FLUIDIZED BED COMBUSTION SYSTEM[a]

	Unit cost[b]	Load	Cost ($/1000 lb steam)
Coal	$35/ton	34,012 lb/hr	$2.38
Limestone	$25/ton[c]	12,256 lb/hr	0.61
Electricity	$0.0325/kwh	2800 kw	0.36
Oil	$0.50/gal	35 gal/hr	0.07
Labor	$16/mhr	11 men	0.70
Solid waste removal	$6/ton	13,484 lb/hr	0.16
		Total	$4.28

[a]The costs shown reflect first year operating costs and an annual load factor of 68 percent.

[b]Unit costs are site specific for Fort Wayne, Indiana area.

[c]Limestone is delivered precrushed to Fluid Bed facility.

Source: Reference 23.

The test data chosen for analysis are referred to as Case No's. 1, 2 and 3 in the remainder of the discussion. To facilitate comparison with the DOE cost data, oil shale and coal feed rates as described by the B&W and ANL investigators were scaled up to the 250,000 lb/hr (steam) FBC size by equating to the heat input rate of 355 x 10^{6} Btu/hr. All calculations for this assessment are included in Appendix B. Direct operating costs for the oil shale-coal systems have been determined only for coal, shale, and solid waste removal. Limestone costs are replaced with costs for shale while electricity, oil, and labor costs are assumed to remain unchanged. The results of this preliminary economic assessment are provided in Table 14. Sorbent costs (especially for shale) are potentially subject to wide variation and therefore total direct operating costs are computed for shale costs ranging from 5 to 30 dollars per ton. The costs at the low end of this range are reasonable based upon information supplied by two industry sources wherein the shale would be mined close to the surface of the ground and made available to a plant at or very near the mine mouth.* Unit costs above the $5/ton figure for shale are basically

*Personal communication between Doug Roeck, GCA and John B. Doyle, Babcock and Wilcox, and between Doug Roeck, GCA and Dr. Vyas, Davy McKee Company, October 28, 1981.

TABLE 14. COMPARISON OF DIRECT OPERATING COSTS FOR A 250,000 lb/hr FBC BOILER BURNING COAL WITH LIMESTONE AND OIL SHALE SORBENTS

	Coal			Sorbent			Ca/S molar ratio	SO$_2$ reduction (%)	Solid waste rate (lb/hr)	Cost data ($/1000 lb steam)							
										Coal	Sorbent					Solid waste removal	Total
Test description	Sulfur (%)	Feed rate (lb/hr)	HHV (Btu/lb)	Type	Feed rate (lb/hr)	HHV (Btu/lb)				at ($35/ton)	at ($5/ton)	at ($10/ton)	at ($20/ton)	at ($25/ton)	at ($30/ton)	at ($6/ton)	direct operating cost[a]
Base case[b]	3 5	34,012	10,430	Limestone	12,256	–	3.0	85	13,484	2 38	–	0 25				0.16	3.92
													0.49			0.16	4.16
														0 61		0.16	4.23
															0 74	0.16	4.41
Case No. 1 (B&W Test No. 11)	3.6	26,377	12,300	Virgin shale	30,557	1,000	2 4	84	31,257	1.85	0.31					0.38	3.67
												0.61				0.38	3 97
													1.22			0.38	4.58
														–	1.83	0.38	5.39
Case No. 2[c] (ANL Test No. 17)	3.0	21,632	12,183	Virgin shale	30,285	3,020	3 1	87 5	23,168	1.51	0.30					0.28	3.22
												0.61				0.28	3 53
													1.21			0.28	4.13
														–	1.82	0 28	4.74
Case No 3[c] (ANL Test No. 20)	2.0	28,950	12,183	Spent shale	5,790	400	1 2	95 9	8,546	2 03	0.06					0.10	3.32
												0 12				0.10	3.35
													0.23			0.10	3 49
														–	0.35	0.10	3.61

[a]Not shown in this table are other components of the total direct operating cost electricity, oil, and labor

[b]Source. Reference 23.

[c]Tests done at Argonne National Laboratory (ANL) were by thermogravimetric analysis

indicative of costs that would be incurred for purchasing shale at increasingly further distances from the shale source. Whereas the DOE base case utilized only one cost for limestone ($25/ton), additional unit costs have been used here for sake of comparison. Figure 6 depicts the total direct operating cost for each of the cases as a function of the sorbent cost. In all, 9 of 12 data points from the three example case curves are indicated to be lower than the base case cost of $4.28/1000 lb steam. It should be reiterated, however, that test data for Case No's. 2 and 3 (Argonne National Laboratory Test No's. 17 and 20) were developed with thermogravimetric analyzer data which may not be as realistic as pilot plant tests results such as those determined in B&W Test No. 11 (Case No. 1).

A comparison of the annualized operating costs for each of the example cases with the coal-limestone base case is shown in Table 15. For this comparison, the indirect operating cost (which is a function of the system's capital costs) is assumed to remain unchanged for all examples. In actuality, the indirect cost may be higher (for Case No's. 1 and 2) since total system feed rates (coal plus shale) are higher than coal plus limestone for the base case (see Table 14) thus requiring larger sized equipment and hence, increased capital expenditure. However, any such increases in capital invest-ment are not expected to significantly affect this analysis. Compared to the base case, annualized operating costs range from -15 to +13 percent for the systems using oil shale as an SO_2 sorbent material.

Taking the scope and nature of the budgetary costs developed in the DOE base case study into consideration as well as the assumption that indirect costs would be the same for the coal-shale systems, the resulting annualized cost estimates for both limestone and shale are probably accurate to $\pm$ 30 percent and thus should be judged to be very similar.
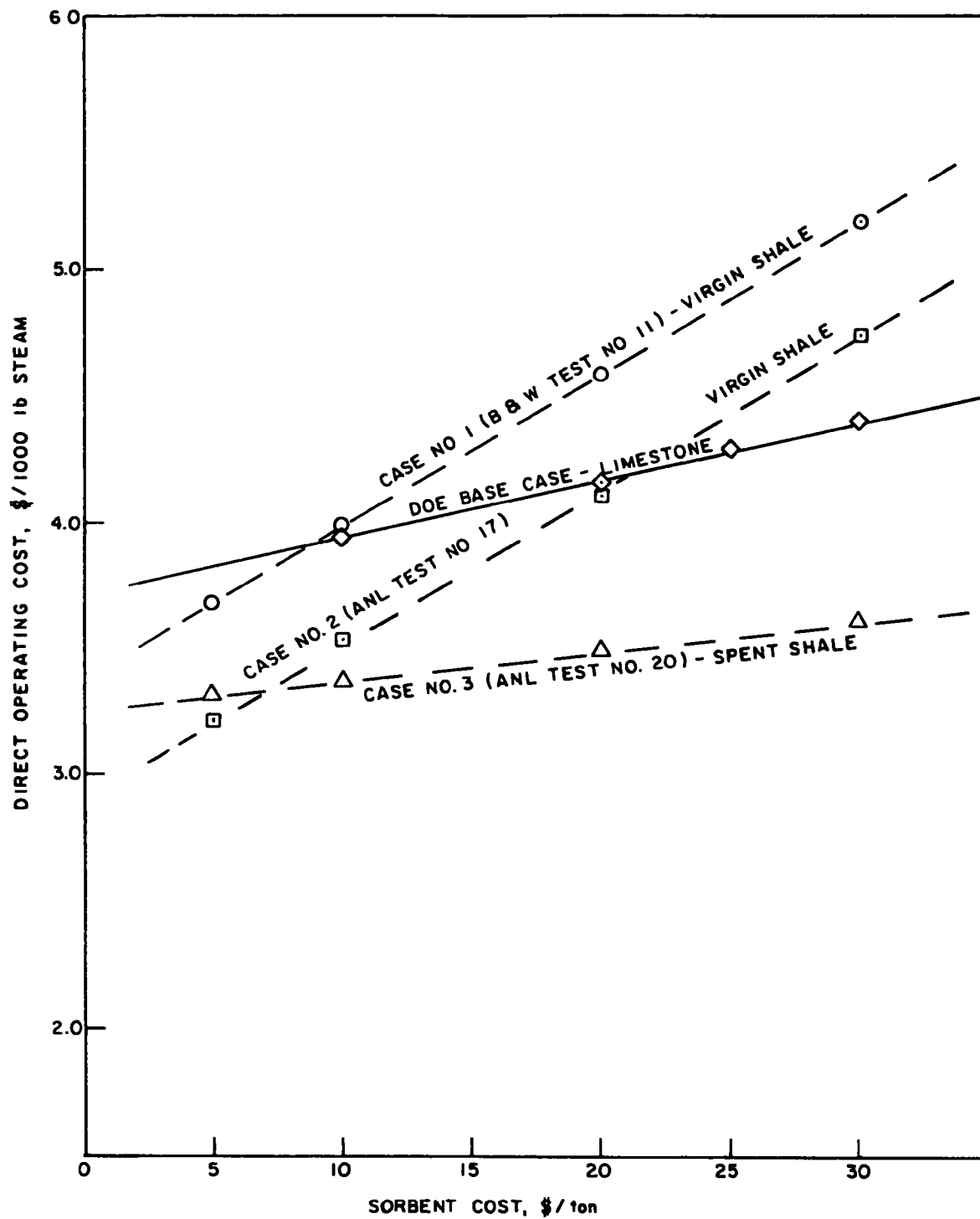
Figure 6. Direct operating cost as a function of sorbent cost for various coal-sorbent systems.

TABLE 15. ANNUALIZED COST COMPARISON BETWEEN COAL-OIL SHALE SYSTEMS AND DOE BASE CASE

| Cost parameter | Test description | Coal-limestone base case | Cost, $/1000 lb steam (percent change over base case) | | | | | | | | | | | |
| | | | Case No. 1 with shale at $/ton | | | | Case No. 2 with shale at $/ton | | | | Case No. 3 with shale at $/ton | | | |
			5	10	20	30	5	10	20	30	5	10	20	30
Direct operating cost		4.28	3.67 (−14.3)	3.97 (−7.2)	4.58 (+7.0)	5.19 (+21.3)	3.22 (−24.8)	3.53 (−17.5)	4.13 (−3.5)	4.74 (+10.7)	3.32 (−22.4)	3.38 (−21.0)	3.49 (−18.5)	3.61 (−15.7)
Indirect operating cost		2.93	2.93	2.93	2.93	2.93	2.93	2.93	2.93	2.93	2.93	2.93	2.93	2.93
Annualized operating cost		7.21	6.60 (−8.5)	6.90 (−4.3)	7.51 (+4.2)	8.12 (+12.6)	6.15 (−14.7)	6.46 (−10.4)	7.06 (−2.1)	7.67 (+6.4)	6.25 (−13.3)	6.31 (−12.5)	6.42 (−11.0)	6.54 (−9.3)

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SECTION 3

SUMMARY AND CONCLUSIONS

The use of fluidized bed combustors in the United States for direct combustion of oil shale is primarily in the research phase at this point in time. Commercial applications of direct combustion of oil shale are limited at this time, and exist only in China, West Germany, and the Soviet Union. To date there are no commercial applications of this technology in the United States.

As discussed in Section 2 of this report, most published research efforts regarding direct combustion of oil share are or have been performed by fluidized bed combustion equipment manufacturers and the U.S. Department of Energy. Results to date have been encouraging in terms of the technical feasibility of combusting both raw and spent shale.

Combustion of raw or virgin shale has been successfully demonstrated by DOE's Morgantown Energy Technology Center and by Babcock and Wilcox Company. DOE successfully burned oil shales from Colorado, Israel, and Morocco, all with combustion efficiencies over 90 percent, while B&W performed favorable tests on a different Colorado shale. Emissions of nitrogen oxides are a potential concern when combusting raw oil shale and during the B&W test program averaged about 3.0 lb/10^6 Btu. Although all of these tests demonstrated the combustibility of virgin oil shale, it is not expected that this will see much widespread application. At present, oil shale research and development efforts are more directed toward retorting technology. One exception could be Eastern U.S. shale which has a low hydrogen to carbon ratio and is more difficult to liquefy than Western shale. However, the HYTORT process being developed at the Institute of Gas Technology (Phillips Petroleum Co. has since purchased the rights to this process*) is aimed specifically at retorting Eastern shale.

Spent shale combustion has also been demonstrated in fluidized bed units with encouraging results. Because spent shale has been subjected to retorting, the type of retort process employed will determine whether or not the shale can be submitted to further useful combustion. Some retorts

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*Personal communication between Doug Roeck, GCA and C. R. Thomas, Southern Indiana Shale Oil, Inc. May 27, 1981.

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(directly heated) utilize all or part of the available organic carbon in the shale as part of the overall process and therefore discharge a product containing very little residual carbon, about 2 to 3 percent or less. Such retorting proceses may produce an off-gas having a heat content of about 100 Btu/scf. Examples of such processes include the Paraho, Union "A", Lurgi-Ruhrgas, and the Superior retorts.

Indirectly heated retorts, on the other hand, produce a spent shale that may contain 4 to 5 percent residual carbon and an off-gas containing 800 Btu/scf. The Paraho (can be operated in both modes), Petrosix, Union "B", TOSCO II, and Gas Combustion processes are all examples of indirectly heated retorts. Because of these aforementioned spent shale characteristics, only processed shale from indirect mode retorts could be considered for fluidized bed combustion. Although several companies are reportedly investigating direct combustion of spent shale in an AFBC, no data are yet publicly available.

Testing of coal-shale mixtures has been performed by several groups since it is felt that oil shale may meet the same three criteria as limestone for use as an SO_2 sorbent in an FBC unit:

1. high calcium carbonate content and hence high SO_2 reactivity

2. low cost

3. wide availability

Testing conducted by the Argonne National Laboratory, the Morgantown Energy Technology Center, and Babcock and Wilcox Co. have all shown favorable SO_2 retention as compared to limestone and dolomite absorbents. Those tests conducted at ANL showed that spent shale had better reaction kinetics than virgin shale. Although more shale would be required than either limestone or dolomite on a unit weight basis for equivalent SO_2 reduction, this can be partially made up for by the heat contribution of the shale. The high calcium content shales that would be desirable for burning with coal are generally found in the Western U.S., whereas high sulfur coals are generally found in the East. A transportation cost penalty would therefore be incurred if such shales were to be used in conjunction with these coals. However, since New Source Performance Standards for combustion units require some degree of sulfur reduction for all coals regardless of sulfur content, the Western shales could be used with the lower sulfur Western coals to achieve the desired removal efficiency. In fact, the ANL tests showed that spent shale requirements were about equivalent to both limestones tested when used with 1.23 and 2.0 percent sulfur coals. Another encouraging aspect of the coal-shale tests (conducted by B&W) was that NO_x emissions were consistently below 0.6 lb/10[6] Btu.

A summary of all available combustion tests performed with raw shale and coal-shale mixtures is provided in Table 16. Most of the data are from the coal-shale tests conducted at Babcock and Wilcox and the Argonne National Laboratory. To provide a means of comparing the tests wherein raw shale, spent shale, limestone, and dolomite are used as SO_2 sorbents, calculations

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TABLE 16a. SUMMARY OF RESULTS FROM DIRECT COMBUSTION OF VIRGIN OIL SHALE

Parameters	Test number				
	1	2	3	4	5
Testing done by[a]	METC	METC	METC	B&W	FW
Shale feed (lb/hr)	18.5-52	180-300	-	358	-
Type	Israel	Israel	Morocco	Colorado	Utah
Sulfur %	2.4	2.4	2.3	0.2	1.15
Ash %	56	56	62	71	73
HHV (Btu/lb)	1,900	1,900	2,400-2,700	1,000	-
Sorbent feed Ca/S mole ratio	7.3	7.3	4.6	∿47	-
Emissions					
SO_2 (lb/10^6 Btu)	0.7	0.7	0.27-0.33	0.05	-
SO_2 reduction %	∿97	∿97	-	99	-
NO_x (lb/10^6 Btu)	0.4	0.4	0.92-1.09	∿3.0	-
NO_x, ppm(v)	-	-	700-900	-	430

[a]METC = Morgantown Energy Technology Center
B&W = Babcock & Wilcox
FW   = Foster-Wheeler

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TABLE 16b. SUMMARY OF RESULTS FROM DIRECT COMBUSTION OF OIL SHALE-COAL MIXTURES

Parameters	Coal-shale mixture tests using virgin shale as sorbent (except tests 13 & 14)													Coal-shale mixture tests using spent shale as sorbent		
	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21
Testing done by[a]	←————————————————————— B&W —————————————————————→									←——————————— ANL ———————————→						
Shale feed																
(lb/hr)	43	23.9	56.6	41.0	37.1	54.1	40.0	-	-	10	25	140	330	6	20	58
Type	←———————— Colorado ————————→							-	-	←————— Colorado (Green River) ————→						
Sulfur %	0.2	0.2	0.2	0.2	0.2	0.2	0.2	-	-	0.9	0.9	0.9	0.9	0.66	0.66	0.66
Ash %	71	71	71	71	71	71	71	-	-	-	-	-	-	-	-	-
HHV (Btu/lb)	1,000	1,000	1,000	1,000	1,000	1,000	1,000	-	-	3,020	3,020	3,020	3,020	400	400	400
Coal feed (lb/hr)	53	35.6	47.9	50.1	48.6	46.7	51.8	51.0	78.0	100	100	100	100	100	100	100
Type	←———————— Ohio No. 6 ————————→							Pittsburgh No. 8	Illinois No. 10	-	-	-	-	-	-	-
Sulfur %	3.6	3.6	3.6	3.6	3.6	3.6	3.6	3.0	4.5	1.23	2.0	3.0	4.3	1.23	2.0	3.0
Ash %	~8.0	~8.0	~8.0	~8.0	~8.0	~8.0	~8.0	~10.0	~10.0	10	10	10	10	10	10	10
HHV (Btu/lb)	12,300	12,300	12,300	12,300	12,300	12,300	12,300	12,800	10,500	12,183	12,183	12,183	12,183	12,183	12,183	12,183
Sorbent feed																
(lb/hr)	43	23.9	56.6	41.0	37.1	54.1	40.0	26.5	36.0	10	25	140	330	6	20	58
Type	←———————— Virgin shale ————————→							Lowellville limestone		←——— Virgin shale ———→				←— Spent shale —→		
Ca/S mole ratio	1.8	1.6	3.0	2.1	1.7	2.4	1.7	5.6	3.6	0.6	1.1	3.1	4.3	0.5	1.2	1.9
Emissions																
SO$_2$ (lb/10^6 Btu)	1.1	1.3	0.8	1.3	1.6	1.0	2.1	-	1.3	1.2	1.2	1.2	1.2	1.2	1.2	1.2
SO$_2$ reduction %	82	78	86	78	74	84	66	82	80	82	84.9	87.5	89.8	95.7	95.9	96.1
NO$_x$ (lb/10^6 Btu)	0.38	0.49	0.44	0.38	0.39	0.38	0.35	-	0.4	-	-	-	-	-	-	-
Solid waste rate																
(lb/hr)	46.3	26.4	57.0	43.3	40.0	55.3	42.6	23.7	35.1	18.5	30.1	107.1	233.6	16.2	29.5	62.0

[a] B&W = Babcock & Wilcox
ANL = Argonne National Laboratory

Note: ANL tests (15-21) are by thermogravimetric analysis (TGA).

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TABLE 16c. SUMMARY OF RESULTS OF TGA STUDIES FOR DIRECT COMBUSTION OF COAL-SORBENT MIXTURES

Parameters	Coal – limestone tests								Coal – dolomite tests			
	22	23	24	25	26	27	28	29	30	31	32	33
Testing done by	←				Argonne National Laboratory							→
Coal feed (lb/hr)	100	100	100	100	100	100	100	100	100	100	100	100
Sulfur %	1.23	2.0	3.0	4.3	1.23	2.0	3.0	4.3	1.23	2.0	3.0	4.3
Ash %	10	10	10	10	10	10	10	10	10	10	10	10
HHV (Btu/lb)	12,183	12,183	12,183	12,183	12,183	12,183	12,183	12,183	12,183	12,183	12,183	12,183
Sorbent feed (lb/hr)	5	16	36	100	8	23	36	60	3	3	18	40
Type	←Germany Valley limestone→				← Greer limestone →				← Tymochtee dolomite →			
Ca/S mole ratio	1.3	2.5	3.8	7.5	1.6	2.8	3.1	3.8	0.4	0.7	1.0	1.5
Emissions												
SO_2 (lb/10^6 Btu)	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2
SO_2 reduction %	40	63.4	75.6	83	40	63.4	75.6	83	40	63.4	75.6	83
Solid waste rate (lb/hr)	15.5	26.9	46.5	104.4	16.4	28.0	38.6	56.6	12.9	12.8	25.6	41.3

have been made to determine total solid waste rates for each test. The solid
waste rate is shown in the last column of the table and includes the calcium
sulfate reaction product as well as any unused sorbent and uncombustible
material (ash) from both the coal and the sorbent. Although solid waste rates
are consistently higher when shale is used as compared to conventional
sorbents, they are not so excessive (especially in the case of the lower sulfur
coals evaluated by ANL) as to be ruled out for use as a sorbent. The basis
for the solid waste rate calculations is provided in Appendix C.

In view of the experimental work done to date on oil shale combustion,
it seems that the concept is technically feasible but as yet untried on
a commercial basis. Because of the tremendous quantities of spent shale
that will be generated once the industry gets underway, it is expected that
companies involved in major oil shale projects will make additional efforts
to extract all available energy from the shale (assuming that such energy
is usable at the facility) prior to ultimate disposal. Although the FBC
represents a method for such heat recovery, it is not the only alternative
and the peculiarities of each retorting facility will determine the best means
for any direct combustion of the shale. As the large oil shale plants get
closer to reality, the companies involved will undoubtedly become more
interested in evaluating potential combustion techniques such as FBC.

The concept of using oil shale as an SO_2 sorbent with coal appears
promising but will depend on the costs of conventional sorbents as well as
other costs related to transporting the raw or spent shale to an industrial
or utility coal-burning facility. Alternatively, a major oil shale plant
using in-situ retorting technology could generate its own power in a coal-
burning power plant using the raw shale removed from the underground mine
as the SO_2 sorbent. It would also seem feasible that oil shale could be
co-fired with high sulfur refinery residuals, Canadian tar sands, or Mexican
sour crudes in an atmospheric fluidized bed combustor.

Since all of the oil shale combustion tests to date have been conducted
in fluidized bed reactors of 18 inches diameter or less, further testing in
larger scale equipment would seem warranted before firm conclusions can be
drawn concerning the prospects of scaling up to commercial application.

The preliminary economic analysis performed in Section 2 would seem to
indicate that costs for operating an FBC unit using oil shale as an SO_2 sorbent
are economically competitive with similarly-sized FBC's using conventional
limestone sorbent. Direct operating costs were shown to range from -25 to +21
percent of the same cost for the coal-limestone base case while total annualized
operating costs ranged from -15 to +13 percent compared to the base case. These
costs are highly dependent on the purchase price of the shale which in turn
would be a function of the distance of the FBC facility from the shale source.
Given that the test data used for this cost comparison were developed from
FBC units of 18-inches diameter or less, such an assessment should be
repeated using test data resulting from larger-scale equipment if and when
such data become available.

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APPENDIX A

CALCULATIONS REGARDING THE MAGNITUDE
OF AN OIL SHALE INDUSTRY'S PRODUCTION
AND DISPOSAL REQUIREMENTS

The following calculations are presented to provide some perspective of the magnitude of shale mining and disposal requirements for an established oil shale industry in the latter part of this century.

Assumptions

- By the 1990-2000 time frame, it is generally assumed that a domestically established oil shale industry will range in size from 400,000 to 1.0×10^6 bbl/day.

- Each individual plant will be sized at 50,000 bbl/day. Thus, from 8 to 20 commercial plants would be required.

- Calculations are performed with Fischer assay values of 35, 25, and 15 gal/ton. For these qualities of shale, the weight percent of oil is about 15, 10, and 6 percent, respectively.

Calculations - (per plant)

(1) Ratio of spent shale to oil--(ton spent shale per bbl oil)

@ 35 gal/ton:

$$\frac{50,000 \text{ bbl}}{\text{day}} \left| \frac{42 \text{ gal}}{\text{bbl}} \right| \frac{\text{ton shale}}{35 \text{ gal}} = 60,000 \text{ ton shale/day}$$

60,000 ton shale/day = 51,000 ton spent shale/day
9,000 ton oil/day

$$\frac{51,000 \text{ ton spent shale}}{50,000 \text{ bbl oil}} \cong 1:1$$

@ 25 gal/ton:

$$\frac{50,000 \text{ bbl}}{\text{day}} \left| \frac{42 \text{ gal}}{\text{bbl}} \right| \frac{\text{ton shale}}{25 \text{ gal}} = 84,000 \text{ ton shale/day}$$

84,000 ton shale/day = 75,600 ton spent shale/day
8,400 ton oil/day

$$\frac{75,600 \text{ ton spent shale}}{50,000 \text{ bbl oil}} \cong 1.5:1$$

@ 15 gal/ton:

$$\frac{50,000 \text{ bbl}}{\text{day}} \left| \frac{42 \text{ gal}}{\text{bbl}} \right| \frac{\text{ton shale}}{15 \text{ gal}} = 140,000 \text{ ton shale/day}$$

140,000 ton shale/day = 131,600 ton spent shale/day
8,400 ton oil/day

$$\frac{131,600 \text{ ton spent shale}}{50,000 \text{ bbl oil}} = 2.6:1$$

(2) Spent shale disposal requirement--

Assume that 25 gal/ton is typical and that the total industry produces 1.0×10^6 bbl/day.

$$\frac{1.0 \times 10^6 \text{ bbl oil}}{\text{day}} \left| \frac{1.5 \text{ ton spent shale}}{\text{bbl oil}} \right| \frac{}{20 \text{ plants}} = 75,000 \text{ ton shale/day per plant}$$

@ 24 hr/day operation:

$$\frac{75,000 \text{ ton}}{\text{day}} \left| \frac{\text{day}}{24 \text{ hr}} \right| \frac{2000 \text{ lb}}{\text{ton}} = 6.25 \times 10^6 \text{ lb spent shale/hr}$$

@ 8 hr/day operation:

$$\frac{75,000 \text{ ton}}{\text{day}} \left| \frac{\text{day}}{8 \text{ hr}} \right| \frac{2000 \text{ lb}}{\text{ton}} = 18.75 \times 10^6 \text{ lb spent shale/hr}$$

(3) Use of FBCs for spent shale combustion--

The largest FBC unit (industrial) built to date fires 338 ton/day or 84,500 lb/hr.

Case A (worst case) -

FBC operates 8 hr/day @ 75,000 lb/hr:

$$\frac{18.75 \times 10^6 \text{ lb/hr}}{75,000 \text{ lb/hr}} = 250 \text{ FBC units per plant}$$

Case B (best case) -

FBC operates 24 hr/day @ 100,000 lb/hr:

$$\frac{6.25 \times 10^6 \text{ lb/hr}}{100,000 \text{ lb/hr}} = 63 \text{ FBC units per plant}$$

Either case indicates that it is most unlikely that all spent shale could be disposed of in this manner. It is more likely that some portion of the shale could be directly combusted for recovery and use of energy in various areas of the plant. The remainder of the shale could be disposed of by more conventional methods such as returning to the mine or landfilling.

APPENDIX B

SUPPORT CALCULATIONS FOR
PRELIMINARY ECONOMIC ANALYSIS
PRESENTED IN SECTION 2

A. **DOE Base Case--**

- 250,000 lb/hr (steam)
- 34,012 lb/hr (coal)
- 10,430 Btu/lb (coal)
- 12,256 lb/hr (limestone)
- 13,484 lb/hr (solid waste)

$$\text{Heat input} = \frac{34{,}012 \text{ lb}}{\text{hr}} \left| \frac{10{,}430 \text{ Btu}}{\text{lb}} \right. = 355 \times 10^6 \text{ Btu/hr}$$

B. **Case No. 1 (B&W Test No. 11)--**

- 54.1 lb/hr (shale)
- 1000 Btu/lb (shale)
- 46.7 lb/hr (coal)
- 12,300 Btu/lb (coal)
- 55.3 lb/hr (solid waste)

(1) Determine coal and shale feed rates scaled to 250,000 lb/hr (steam) boiler size--

x = shale feed rate, lb/hr

y = coal feed rate, lb/hr

$$\frac{x}{y} = \frac{54.1}{46.7} \qquad \text{(a)}$$

- Assume same heat input as base case:

$$1000 \, x + 12{,}300 \, y = 355 \times 10^6 \qquad \text{(b)}$$

Simultaneous solution of Equations (a) and (b) yields:

$$x = 30{,}557 \text{ lb/hr (shale)}$$
$$y = 26{,}377 \text{ lb/hr (coal)}$$

(2) Determine total solid waste rate--

- Assume that the ratio of solid waste to coal and sorbent feed rates is the same for the scaled-up system, s = new solid waste rate

$$\frac{55.3}{54.1 + 46.7} = \frac{S}{30{,}557 + 26{,}377}$$

$$S = 31{,}257 \text{ lb/hr (solid waste)}$$

(3) Determine impact on direct operating cost--

   a. Coal--

$$\frac{26{,}377 \text{ lb}}{\text{hr}} \left| \frac{\text{ton}}{2000 \text{ lb}} \right| \frac{\$35}{\text{ton}} \left| \frac{\text{hr}}{250 \times 10^3 \text{ lb steam}} \right. = \$1.85/10^3 \text{ lb steam}$$

   b. Sorbent--

$$\frac{30{,}557 \text{ lb}}{\text{hr}} \left| \frac{\text{ton}}{2000 \text{ lb}} \right| \frac{\$5}{\text{ton}} \left| \frac{\text{hr}}{250 \times 10^3 \text{ lb steam}} \right. = \$0.31/10^3 \text{ lb steam}$$

   shale @ \$10/ton   = \$0.61/10^3 lb steam

   shale @ \$20/ton   = \$1.22/10^3 lb steam

   shale @ \$30/ton   = \$1.83/10^3 lb steam

   c. Solid Waste Removal--

$$\frac{31{,}257 \text{ lb}}{\text{hr}} \left| \frac{\text{ton}}{2000 \text{ lb}} \right| \frac{\$6}{\text{ton}} \left| \frac{\text{hr}}{250 \times 10^3 \text{ lb steam}} \right. = \$0.38/10^3 \text{ lb steam}$$

C. Case No. 2 (ANL Test No. 17)--

   • 140 lb/hr (shale)
   • 3,020 Btu/lb (shale)
   • 100 lb/hr (coal)
   • 12,183 Btu/lb (coal)
   • 107.1 lb/hr (solid waste)

(1) Coal and shale feed rates--

$$\frac{x}{y} = \frac{140}{100} \qquad\qquad (a)$$

$$3{,}020 \, x + 12{,}183 \, y = 355 \times 10^6 \qquad\qquad (b)$$

$$x = 30{,}285 \text{ lb/hr (shale)}$$

$$y = 21{,}632 \text{ lb/hr (coal)}$$

(2) Solid waste rate--

$$\frac{107.1}{100 + 140} = \frac{S}{30{,}285 + 21{,}632}$$

$$S = 23{,}168 \text{ lb/hr (solid waste)}$$

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(3) Impact on direct operating cost—

    a.   Coal—

$$\frac{26{,}632 \text{ lb}}{\text{hr}} \left| \frac{\text{ton}}{2000 \text{ lb}} \right| \frac{\$35}{\text{ton}} \left| \frac{\text{hr}}{250 \times 10^3 \text{ lb steam}} \right. = \$1.51/10^3 \text{ lb steam}$$

    b.   Sorbent—

$$\frac{30{,}285 \text{ lb}}{\text{hr}} \left| \frac{\text{ton}}{2000 \text{ lb}} \right| \frac{\$5}{\text{ton}} \left| \frac{\text{hr}}{250 \times 10^3 \text{ lb steam}} \right. = \$0.30/10^3 \text{ lb steam}$$

                       shale @ \$10/ton   = \$0.61/10^3 lb steam

                       shale @ \$20/ton   = \$1.21/10^3 lb steam

                       shale @ \$30/ton   = \$1.82/10^3 lb steam

    c.   Solid waste removal—

$$\frac{23{,}168 \text{ lb}}{\text{hr}} \left| \frac{\text{ton}}{2000 \text{ lb}} \right| \frac{\$6}{\text{ton}} \left| \frac{\text{hr}}{250 \times 10^3 \text{ lb steam}} \right. = \$0.28/10^3 \text{ lb steam}$$

D.   <u>Case No. 3 (ANL Test No. 20)</u>—

- 20 lb/hr (shale)
- 400 Btu/lb (shale)
- 100 lb/hr (coal)
- 12,183 Btu/lb (coal)
- 29.5 lb/hr (solid waste)

    (1) Coal and shale feed rates—

$$\frac{x}{y} = \frac{20}{100} \qquad\qquad\qquad\qquad\text{(a)}$$

$$400\,x + 12{,}183\,y = 355 \times 10^6 \qquad\qquad\text{(b)}$$

$$x = 5{,}790 \text{ lb/hr (shale)}$$

$$y = 28{,}950 \text{ lb/hr (coal)}$$

    (2) Solid waste rate—

$$\frac{29.5}{100 + 20} = \frac{S}{28{,}950 + 5{,}790}$$

$$S = 8{,}546 \text{ lb/hr (solid waste)}$$

(3) Impact on direct operating cost--

    a. Coal--

$$\frac{28{,}950 \text{ lb}}{\text{hr}} \left| \frac{\text{ton}}{2000 \text{ lb}} \right| \frac{\$35}{\text{ton}} \left| \frac{\text{hr}}{250 \times 10^3 \text{ lb steam}} \right. = \$2.03/10^3 \text{ lb steam}$$

    b. Sorbent--

$$\frac{5{,}790 \text{ lb}}{\text{hr}} \left| \frac{\text{ton}}{2000 \text{ lb}} \right| \frac{\$5}{\text{ton}} \left| \frac{\text{hr}}{250 \times 10^3 \text{ lb steam}} \right. = \$0.06/10^3 \text{ lb steam}$$

                        shale @ \$10/ton   = \$0.12/10^3 lb steam

                        shale @ \$20/ton   = \$0.23/10^3 lb steam

                        shale @ \$30/ton   = \$0.35/10^3 lb steam

    c. Solid waste removal--

$$\frac{8{,}546 \text{ lb}}{\text{hr}} \left| \frac{\text{ton}}{2000 \text{ lb}} \right| \frac{\$6}{\text{ton}} \left| \frac{\text{hr}}{250 \times 10^3 \text{ lb steam}} \right. = \$0.10/10^3 \text{ lb steam}$$

APPENDIX C

SUPPORT CALCULATIONS FOR DETERMINING TOTAL SOLID WASTE
RATES FOR COAL-SORBENT TESTS PRESENTED IN TABLES 16b AND 16c

Solid waste rates have been determined for FBC systems burning coal with both oil shale and conventional sorbents. The calculations are based upon available data in each of the referenced reports cited in the text.

Solid waste rate is defined as the total weight of material that is discharged from the FBC system and includes the following: the sorbent/SO_2 reaction product, calcium sulfate ($CaSO_4$); unreacted calcium oxide (CaO); inerts in the sorbent; and the ash contained in the coal. The following equation is used:

$$\text{SWR} = \text{SFR} \left\{ \left[\left(\frac{F_s}{Ca/S} \right) \left(\frac{136}{100} \right) \left(1 - F_{as} \right) \right] + \left[\left(1 - \frac{F_s}{Ca/S} \right) \left(\frac{56}{100} \right) \left(1 - F_{as} \right) \right] + F_{as} \right\}$$

CaSO$_4$:   CaO: sorbent ash:

coal ash:

$$+ \text{CFR } (F_{ac})$$

where: SWR = solid waste rate

  SFR = sorbent feed rate

  F_s = fractional sulfur removal

  Ca/S = calcium to sulfur molar ratio

$$\frac{F_s}{Ca/S} = \begin{array}{l}\text{calcium} \\ \text{utilization factor} \\ \text{(dimensionless)}\end{array}$$

  $\dfrac{136}{100}$ = weight ratio of $\dfrac{CaSO_4}{CaCO_3}$

  F_{as} = fraction of ash in the sorbent, or

  $1 - F_{as}$ = $CaCO_3$ content of sorbent

  $\dfrac{56}{100}$ = weight ratio of $\dfrac{CaO}{CaCO_3}$

  CFR = coal feed rate

  F_{ac} = fraction of ash in the coal

The solid waste rate calculations have been performed for test numbers 6 through 33 as presented in Tables 16b and 16c of the text. For test no's. 13 and 14 (B&W), it was assumed that the coal contained 10 percent ash. For test no's. 15 through 33 (ANL), several assumptions had to be made:

● ash content of coal = 10 percent

● heat content of spent shale = 400 Btu/lb

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- 78 percent of the sulfur in the shale is converted to SO_2[22]

- a basis of 100 lb/hr coal feed is used for all calculations

Wherever possible, limestone, dolomite, and shale analyses have been used. For example, the limestone analysis as indicated in the footnote to Table 10 is used for B&W Test No's. 13 and 14. Also, the shale, dolomite and limestone constituents as provided in Table 11 are used in the calculations for all ANL tests. Since all of the sorbents shown in Table 11 contain magnesium carbonate ($MgCO_3$), corrections have been made for the fact that this compound does undergo calcination (to form MgO) but does not react with sulfur dioxide. (See Example calculation (5)).

Since reference 22 only provided information on Ca/S ratios and unit weights of sorbent per unit weights of coal required, percent sulfur reduction had to be determined. Table C-1 provides this information.

TABLE C-1. UNCONTROLLED SO_2 LEVELS AND REMOVAL EFFICIENCIES REQUIRED FOR ANL TEST DATA

Sulfur content of coal (%)	Uncontrolled SO_2 levels (lb $SO_2/10^6$ Btu)			Efficiency required to meet 1.2 lb $SO_2/10^6$ Btu (%)	
	Coal	Raw shale	Total	Coal only	Shale + coal
1.23	2.0	4.68	6.68	40	82.0
2.0	3.28	4.68	7.96	63.4	84.9
3.0	4.92	4.68	9.6	75.6	87.5
4.3	7.06	4.68	11.74	83.0	89.8
	Coal	Spent shale	Total	Coal only	Shale + coal
1.23	2.0	25.74	27.74	40	95.7
2.0	3.28	25.74	29.02	63.4	95.9
3.0	4.92	25.74	30.66	75.6	96.1
4.3	7.06	25.74	32.8	83.0	96.3

Several sample calculations are provided as follows:

(1) Uncontrolled SO_2 levels in raw and spent shale and coal for ANL tests:

Raw shale--

0.9% S, 3,020 Btu/lb

Basis: 100 lb raw shale

$$\frac{0.9 \text{ lb S}}{0.302 \times 10^6 \text{ Btu}} \left| \frac{2 \text{ lb } SO_2}{\text{lb S}} \right| \frac{0.78}{} = 4.68 \text{ lb } SO_2/10^6 \text{ Btu}$$

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Spent shale--

0.66% S, 400 Btu/lb

Basis: 100 lb spent shale

$$\frac{0.66 \text{ lb S}}{0.04 \times 10^6 \text{ Btu}} \left| \frac{2 \text{ lb SO}_2}{1 \text{ lb S}} \right| 0.78 = 25.74 \text{ lb SO}_2/10^6 \text{ Btu}$$

3.0% S Coal--

3.0% S, 12,183 Btu/lb

Basis: 100 lb coal

$$\frac{3 \text{ lb S}}{1.2183 \times 10^6 \text{ Btu}} \left| \frac{2 \text{ lb S}}{1 \text{ lb SO}_2} \right. = 4.92 \text{ lb SO}_2/10^6 \text{ Btu}$$

(2) Solid waste rate for B&W Test No. 10 (Table 16b)--

$$\text{SWR} = 37.1 \left\{ \left[\left(\frac{0.74}{1.7}\right)\left(\frac{136}{100}\right)\left(1-0.71\right) \right] + \left[\left(1 - \frac{0.74}{1.7}\right)\left(\frac{56}{100}\right)\left(1-0.71\right) \right] + 0.71 \right\} + 48.6 \ (0.08)$$

SWR = 6.37 + 3.40 + 26.34 + 3.89

SWR = 40 lb/hr

(3) Solid waste rate for ANL Test No. 20 (Table 16b)--

- 2.0% S coal requires 95.9 percent reduction to meet 1.2 lb $SO_2/10^6$ Btu (from Table C-1)

- amount of sorbent (spent shale) required = 0.2 lb spent shale/lb coal (from Table 12)

- Basis: 100 lb/hr coal feed

$$\text{SWR} = 20 \left\{ \left[\left(\frac{0.959}{1.2}\right)\left(\frac{136}{100}\right)\left(0.335\right) \right] + \left[\left(1 - \frac{0.959}{1.2}\right)\left(\frac{56}{100}\right)\left(0.335\right) \right] + \left[0.493 + 0.167\left(\frac{40.3}{84.3}\right) \right] \right\}$$
$$+ \ 100 \ (0.10)$$

SWR = 7.28 + 0.75 + 11.46 + 10

SWR = 29.5 lb/hr

(4) Solid waste rate for ANL Test No. 25 (Table 16c)--

- 4.3% S coal requires 83.0 percent reduction to meet 1.2 lb $SO_2/10^6$ Btu (from Table C-1)

- amount of sorbent (Germany Valley limestone) required = 1.0 lb limestone/lb coal (from Table 12)

- Basis: 100 lb/hr coal feed

$$SWR = 100 \left\{ \left[\left(\frac{0.83}{7.5}\right)\left(\frac{136}{100}\right)(0.9775) \right] + \left[\left(1 - \frac{0.83}{7.5}\right)\left(\frac{56}{100}\right)(0.9775) \right] \right.$$

$$\left. + \left[0.0235 + 0.6 \left(\frac{40.3}{84.3}\right) \right] \right\}$$

$$+ 100 \ (0.10)$$

$$SWR = 14.71 + 48.68 + 31.03 + 10$$

$$SWR = 104.4 \ \text{lb/hr}$$

(5) Solid waste rate for ANL Test No. 32 (Table 16c)--

- 3.0% S_6 coal requires 75.6 percent reduction to meet 1.2 lb $SO_2/10^6$ Btu (from Table C-1)

- amount of sorbent (Tymochtee dolomite) required = 0.18 lb dolomite/lb coal (from Table 12)

- Basis: 100 lb/hr coal feed

$$SWR = 18 \left\{ \left[\left(\frac{0.756}{1.0}\right)\left(\frac{136}{100}\right)(0.518) \right] + \left[\left(1 - \frac{0.756}{1.0}\right)\left(\frac{56}{100}\right)(0.518) \right] \right.$$

$$\left. + \left[0.0555 + 0.433 \left(\frac{40.3}{84.3}\right) \right] \right\}$$

$$+ 100 \ (0.10)$$

$$SWR = 9.59 + 1.27 + 4.72 + 10$$

$$SWR = 25.6 \ \text{lb/hr}$$

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16 ABSTRACT The report gives an overview of the combustion of oil shale in fluidized-bed combustors. Oil shale can be combusted directly, or as a supplement to coal fuel, and can function as an SO2 sorbent in atmospheric fluidized-bed combustion (AFBC). Spent shale from retorting processes may provide for SO2 sorption and add some residual fuel value in AFBC. Commercial applications of direct combustion of oil shale in AFBC are limited, and known to exist only in Estonia (Soviet Union), China, and southern Germany. Pilot- and laboratory-scale combustion of shale and coal/shale mixtures in AFBCs in the U.S. has been investigated. Technical concerns include calcination heat loss, optimal shale sizing, and combustion efficiency. Direct combustion of western U.S. shale in a pilot AFBC produced NOx emissions as high as 3 lb/million Btu; although, for coal/shale mixtures, NOx emissions were < 0.6 lb/million Btu. Because of its calcium carbonate ($CaCO_3$) content, shale can act as an effective SO2 sorbent in AFBC, and has produced SO2 reduction efficiencies of > 85% in AFBC tests with coal/shale mixtures. A preliminary economic analysis indicates that substituting oil shale for limestone in an AFBC may have significant cost advantages, especially where shale (or spent shale) is readily available and at low cost.

17	KEY WORDS AND DOCUMENT ANALYSIS	
a DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c COSATI Field/Group
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