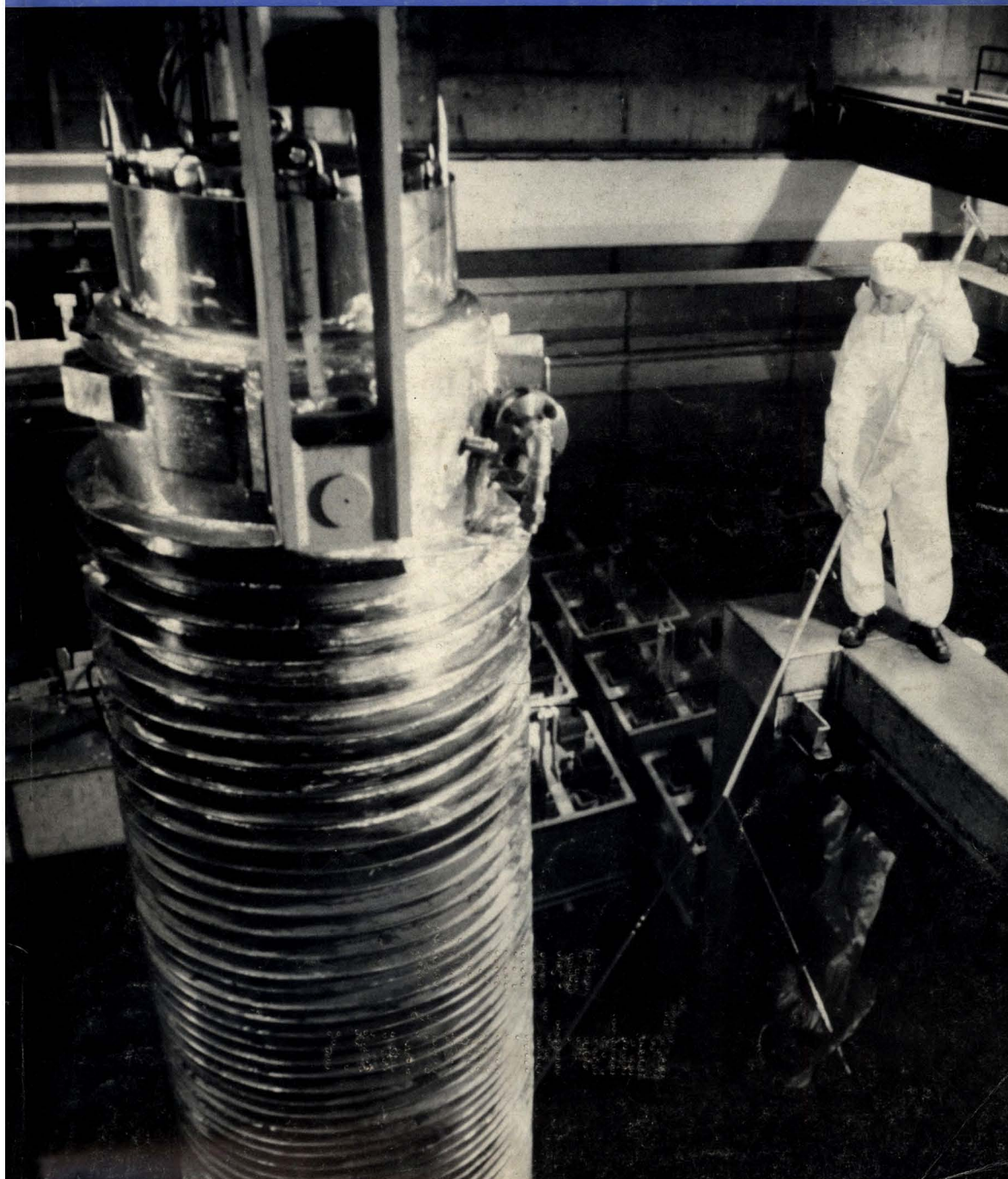


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Cover: Storage pool for radioactive material in Dresden, Illinois power plant.

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Energy Shorts

DOE Selects Coal Combustor Projects

Six new coal burning concepts—tailored to meet the needs of small factories, businesses, and residential complexes—will be developed under cost-shared contracts to be negotiated by the U.S. Department of Energy.

Each of the experimental coal combustors, selected by the DOE in two, separate competitive procurements, will be based on the “fluidized bed” coal burning concept.

The fluidized bed technique is being used commercially today in large boilers, but additional development is needed before it is feasible for the small boiler market.

To assist industry in this development, the Energy Department's Morgantown, W. Va. Energy Technology Center has selected two firms to develop concepts suitable for boilers that generate between 75,000 and 150,000 pounds per hour of steam—the size typical of a small, industrial operation.

The center also selected four projects tailored to even smaller sizes—down to 1,000 pounds of steam per hour—typical of the small boilers in shopping centers, apartment buildings, and other commercial establishments.

The two large-scale projects will be conducted by York-Shipley and Riley-Stoker. Total costs to the government of the two projects is expected to be \$2.7 million, while the private firms will contribute about \$2.6 million. If the projects proceed as planned, the two firms will design and ultimately fabricate

small test combustors that could then be scaled up for commercial use.

York-Shipley will combine the characteristics of a more conventional fluidized bed combustor with those of a cyclone furnace. Rather than feeding air through the bottom of the combustor, as many fluidized beds do, some of the air is injected through jets in the walls. This gives a swirling, vortex motion to the burning coal, increasing combustion intensity.

Riley-Stoker will develop a similar, circulating fluidized bed concept. In that firm's concept, coal and limestone particles are transported by the gas stream to the top of the combustor, then separated from the air flow and returned to the bottom of the combustor. Larger and heavier limestone particles also are added to the lower region of the boiler to “hold” some of the coal and limestone before they are caught up in the air flow. This gives the combustor both a “bubbling” and “circulating” fluidized bed motion.

Riley-Stoker also is one of the four firms selected to develop concepts for the smaller boiler market. The company was chosen along with Hydrocarbon Research, Inc., Energy and Environmental Research, Inc., and Management and Technology Consultants, Inc. Together, the four will share \$6.2 million in federal funds while contributing a total of approximately \$3.4 million.

As with its larger concept, Riley-

Stoker hopes to achieve high combustion efficiency, economical limestone use, and a high degree of nitrogen and sulfur emission control. The system is projected to be capable of handling a wide range of coal sizes, including coal directly from the mine, as well as liquid and slurried fuels. It also is projected to be capable of responding rapidly to changes in steam demand.

Hydrocarbon Research, Inc. proposes a concept called a “dual-sided, multi-riser” fluidized bed. This combustor will have several baffle-like devices inside the combustor—called “risers” and “downcomers”—to channel the air and solids up and down. This configuration makes the combustor very compact and economical, while providing good combustion efficiency, safe and reliable operation, and good environmental performance.

Energy and Environmental Research will develop an air-cooled, fluidized bed combustor to supply warm air, hot water or process steam to a commercial or light industrial establishment. Since air is used to convey the heat from the boiler, no in-bed steam tubes are present, thereby eliminating tube erosion and corrosion.

Management and Technology Consultants will evaluate a concept which sends pulsed air through the limestone particles and burning coal. The pulses enhance the combustion process and boost the amount of heat released and transferred to the boiler tubes. The result is a smaller combustor size.

Letter

On "Process Screening and Selection for Refinery Acid Gas Removal Processing"

To the editor:

In their article [*Energy Progress*, 6, 239 (1986)] Gupta, Stanbridge, Wendt and Brocoff state that iron sponge is "unsuitable for refinery applications." We find this opinion unsupportable since iron sponge is presently being used to sweeten both refinery gases and light hydrocarbon liquids in numerous locations world-wide.

The authors remark that iron sponge is unsuitable "due to safety (fire hazard) and environmental (odor and disposal problems) considerations." Also, the authors suggest that the process is suitable only for removal of "ppm quantities of H₂S (hydrogen sulfide)."

Iron sponge is a completely safe, non-hazardous chemical. In certain circumstances, *spent* iron sponge can display the hazardous characteristic of "ignitability." However, when properly handled prior to disposal, the spent material will quickly revert to non-hazardous iron oxide and elemental sulfur on the wood substrate. This handling primarily involves keeping the spent material wet and spread into a thin layer for a period of several days. This procedure is accepted by Federal and State hazardous waste agencies.

Properly designed and operated iron sponge systems remove hydrogen sulfide and other sulfur compounds, such as mercaptans, to below the ppm level. Iron sponge systems are completely closed and do *not* produce any odor problems. The *spent* material may pick up the odor of hydrocarbons only, since all sulfur bearing compounds are reacted.

The design procedure for iron sponge gas and liquid treating systems is not limited in sulfur handling capacity. Many operators of iron sponge systems have initial conditions of several thousand ppm hydrogen sulfide.

The iron sponge process would receive high ratings for all of the "Process Characteristics" in Table 5, Weighted Process Evaluation. Iron sponge systems enjoy low capital and operating costs, have proven reliable in hundreds of commercial operations, and are extremely easy to operate.

The authors of this article have discounted iron sponge as a viable alternative for sweetening acid gas in other publications. It is our belief that the process has been unjustly removed from the screening and selection procedure. Iron sponge should not be overlooked as a potential sweetening system of choice in any refinery operation.

Brett Jay Davis
Project Engineer
Physichem Technologies, Inc.
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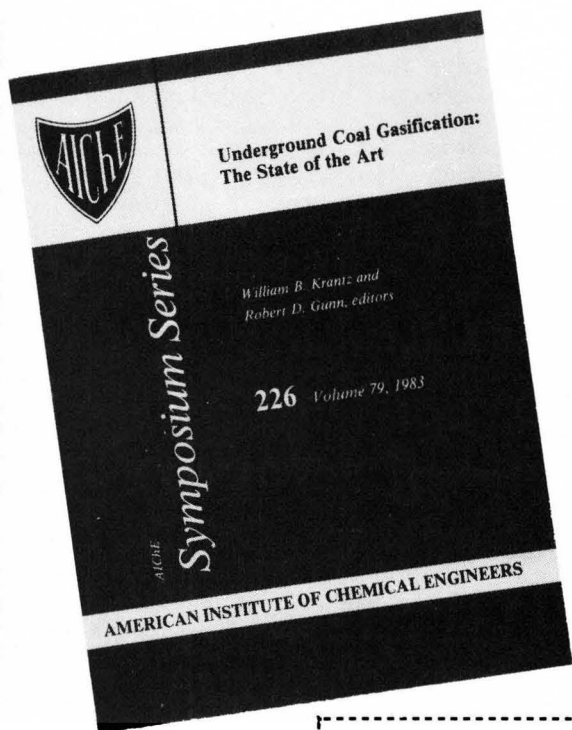
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Underground Coal Gasification: The State of the Art

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- The Hanna Wyoming Underground Coal Gasification Field Test Series
- Review of Underground Coal Gasification Field Experiments at Hoe Creek
- ARCO's Research and Development Efforts in Underground Coal Gasification
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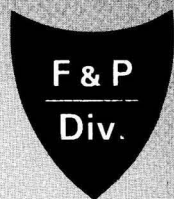
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Future Meetings And Programming

Your help in presenting papers, organizing sessions, or developing new programming areas is needed. If you have a desire to participate in the Fuels and Petrochemicals Division programming activities or wish further information, please contact: **Bill Rodgers, Chairman, F&PD Programming, (615) 574-6819**. The programming committee will meet in Minneapolis on Sunday morning, August 16th, at 9:00 A.M. (room to be announced). Please join us with your ideas and input.

The Fuels and Petrochemicals Division plans the following sessions for the upcoming meetings:

Minneapolis	August, 1987	12 Sessions
New York	November, 1987	6 Sessions
New Orleans	April, 1988	35 Sessions
Denver	August, 1988	12 Sessions

There will be joint programming with the Heat Transfer and Energy Conversion Group and the Health and Safety Division. Be sure to keep posted on the details in the upcoming issues of *Chemical Engineering Progress*.

Survey Results

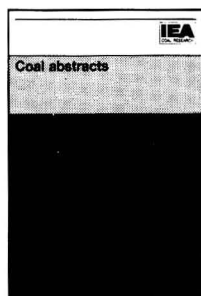
John Heiman recently conducted a survey of a select group of F&PD members. The survey was sent to 350 members and responses were received from 168 people. From the results, the following conclusions were reached.

1. An overwhelming majority of the members are pleased with *Energy Progress* and the newsletter format that we currently have.
2. Programming is obviously the major concern and the most important benefit that our members are receiving.
3. Alternative energy sources are of great concern and interest to our members. The formulation of a national energy policy that makes us independent of foreign oil is important.
4. New technologies and energy conservation and management are important programming areas.
5. There is very little interest in the postprint concept.
6. Over half of those responding have attended a National Meeting at some time, and over half could attend future meetings if the programs were of interest.

NEWSLETTER EDITOR: **Peter D. Wanser**

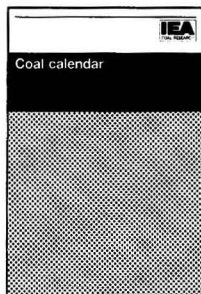
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Comparative Economics For Alternative Fuels And Power Technologies

Irving Leibson

Bechtel, Inc., Box 3965, San Francisco, CA 94119

Comparative economics for the various alternative fuels-technologies including tar sands, oil shale, coal gasification, and direct coal liquefaction.

INTRODUCTION

The fluctuations in the price of crude oil and refined petroleum products have created substantial uncertainty and turbulence in the decision making involved in inter fuel substitution and selection of energy resource base both in near term operational situations and longer term investment decisions. The business climate generated by the precipitous change in oil prices is not conducive to the commitment of financial resources to the longer term research and development expenditures required to develop and demonstrate alternative fuels and power technologies for potential use in the decade of the 90s and beyond.

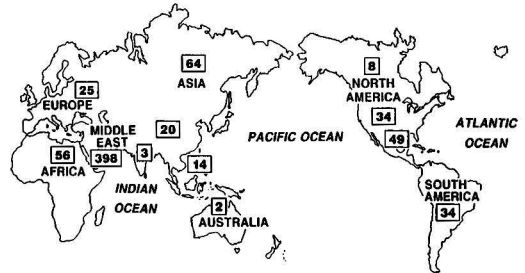
This article assesses the commercial status and comparative economics for the various alternative fuels technologies including tar sands, oil shale, coal conversion via gasification, direct and indirect coal liquefaction, methanol, and biomass derived fuels. Facility sizes have been selected which represent state-of-the-art for scale of operation for the year of investment decision. The evaluation of those technologies which are fully commercially demonstrated is based on 1986 as the first year of operation.

In the case of the power technologies, the cost of electricity at the busbar is analyzed for the conventional technologies such as pulverized coal fired boilers with flue gas desulfurization, oil fired, natural gas fired, and nuclear generating stations as well as those technologies under current development such as integrated coal gasification-combined cycle and fluidized bed combustion.

World Energy Reserves

To gain perspective concerning the various fuels to be considered for diversity and security of fuel supply, let's review the location and size of the various fuel resources worldwide. Figure 1 shows the geographic distribution of the world's economically recoverable proven reserves of crude oil.

Proven recoverable reserves are defined as deposits which have been identified and can be extracted using available technology at prevailing prices. More than half of these reserves are located in the Middle East. Furthermore, the lifting costs of crude oil in the Middle East are



TOTAL: 707 BILLION BARRELS
SOURCE: BRITISH PETROLEUM STATISTICAL REVIEW OF WORLD ENERGY, 1985

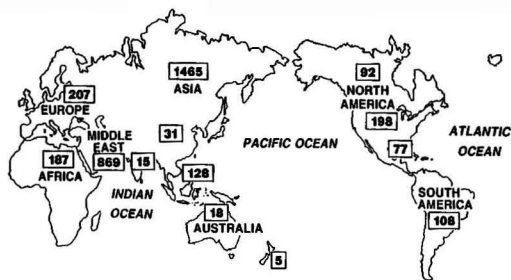
Figure 1. World proven recoverable oil reserves (billions of barrels).
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substantially smaller (because of the comparative huge size of the reservoirs) than that of many of the other producing areas. The total economically recoverable proven oil reserves worldwide are currently estimated to be about 707 billion barrels. World oil production is currently running approximately 21 billion barrels per year [1]. World oil production peaked in 1979 at 24 barrels per year.

Additional substantial resources in the form of heavy oil, tar sands, and oil shale are virtually untapped. In Canada, two facilities are in operation, surface mining tar sands, recovering the bitumen, and upgrading it to synthetic crude oil. Substantial deposits of heavy oil exist in Canada and the United States. Major heavy oil deposits also exist in the Orinoco Belt of Venezuela. Furthermore, substantial quantities of oil shale exist in the United States, Australia, Brazil, China, and the Soviet Union.

The world's proven economically recoverable reserves of natural gas are shown in Figure 2. World gas production for 1984 was approximately 57 trillion cubic feet [1]. Because natural gas is presently being consumed at about one half the rate of oil, the peak in natural gas production seems likely to occur later than oil. The worldwide production capability for natural gas from conventional resources is expected to increase substantially over the next 15 years.

New gas discoveries are running at a rate approxi-



TOTAL: 3,400 TRILLION CUBIC FEET
SOURCE: BRITISH PETROLEUM STATISTICAL REVIEW OF WORLD ENERGY, 1985

Figure 2. World proven gas reserves (trillion cubic feet)

mately twice that for oil. Substantial gas has been discovered on the Northwest Shelf of Australia and a project is being developed for conversion of that gas to liquefied natural gas (LNG) for potential shipment to Japan, Korea, and Taiwan.

Substantial activity is currently underway in many companies around the world to develop the technology necessary for the conversion of natural gas to wide boiling hydrocarbon liquids. The joint venture of Mobil and the government of New Zealand has recently started up a facility to convert natural gas from the Maui gas field to gasoline using new technology developed by Mobil. Shell is currently considering initiation of a project in Malaysia for the conversion of natural gas to wide boiling hydrocarbon liquids using new technology developed by them for this purpose. These same technologies are applicable to the conversion of synthesis gas (produced via coal gasification) to liquid products. Thus it appears that we shall soon have the technical capability to selectively convert fuel forms such as natural gas and coal to hydrocarbon liquids and that this sort of interfuel conversion could become important by the end of the century.

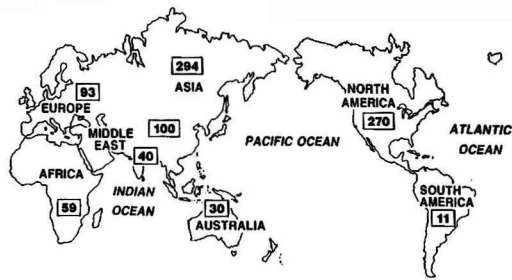
Figure 3 shows the world's economically recoverable coal reserves. Many of the major industrial countries in Western Europe and the Pacific Basin have only small coal reserves. Almost two-thirds of the world's coal reserves are concentrated in the United States, China, and the USSR. The estimated total world proven recoverable reserves are 897 billion metric tons (2). World coal production is currently approximately 3.2 billion metric tons annually.

Turning to uranium, Figure 4 shows the estimates of the free world's reserves of uranium recoverable at costs up to \$30/lb of yellowcake [3]. The OECD and the IAEA estimate that about 40% more uranium would be available at costs up to \$50/lb of yellowcake. An estimate of these additional resources recoverable at less than \$50/lb of yellowcake is shown in the lower boxes in Figure 4. North America has about one half of these additional reserves. To a degree, this reflects increases which have occurred in production costs which have moved resources into higher cost categories. The Asia Pacific region has about 17% of the free world's total known uranium reserves with very significant reserves in Australia. Very little is known about the uranium reserves in the countries with centrally planned economies.

Table 1 summarizes world economically recoverable energy reserves and the corresponding 1984 production rates.

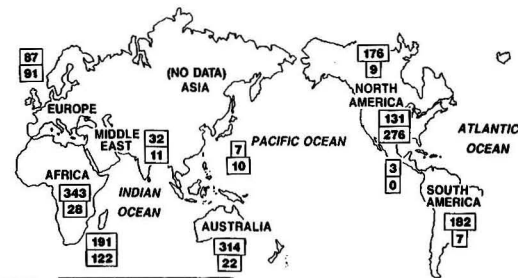
ALTERNATIVE FUELS TECHNOLOGIES

Table 2 gives the economical premises set for comparison of the alternative fuel technologies. The investment



TOTAL: 897 BILLION METRIC TONS COAL EQUIVALENT
SOURCE: ENERGY INFORMATION ADMINISTRATION
U.S. DEPARTMENT OF ENERGY INTERNATIONAL ENERGY ANNUAL, 1984

Figure 3. World economically recoverable coal reserves (billion metric tons coal equivalent).



TOTAL: 1466 METRIC TONS URANIUM
576 METRIC TONS URANIUM
KEY: RECOVERABLE <math>\le 80/\text{kgU} = $\le 30/\text{lb U}_2\text{O}_8$
RECOVERABLE $80-30/\text{kgU} = 30-50/\text{lb U}_2\text{O}_8$
SOURCE: URANIUM—OECD/IAEA, DECEMBER 1983, p. 19

Figure 4. World uranium reserves and resource (metric tons uranium).

and operating cost data have been derived for the most part from conceptual studies and design work done over a period of the past ten years for a wide variety of client organizations. The cost of the fuel produced for each technology is developed on the basis of the capital investment being comprised of 20% equity, 80% debt with a 10% after tax return on equity and an 8% interest on debt. A general inflation rate for all costs is taken as 2.5% per year. Further details of the premises are shown as footnotes in Table 2. In Figure 5, the cost of the product fuel is segmented to show the corresponding portions of cost attributable to capital charges, operations and maintenance, and feedstock or fuel costs. Table 3 provides a summary of the premises for fuel and feedstock costs, facility capacity, capital investment, and first year of operation used as the basis for Table 2 and Figure 5.

Tar Sands

Two commercial scale facilities are in operation in Canada for the open pit mining of tar sands, hot water process extraction of bitumen from the tar sands, and coking of the bitumen with hydrogenation of individual product fractions to produce synthetic crude oil. Figure 5 shows a total product cost of \$24.9/barrel for a 50,000 barrel per day tar sands facility. Developmental activity sponsored by Alberta Oil Shale and Tar Sands Research Authority (AOSTRA) jointly with various oil companies is currently aimed at developing in situ approaches for bitumen recovery from tar sands, alternative processing technologies to replace the Clark hot water extraction

TABLE 1. WORLD ENERGY RESERVES AND UTILIZATION
(BILLION BARRELS OF OIL EQUIVALENT)

Fuel Type	Proven Reserves	1984 Production
Oil	707	21
Gas	636	11
Coal	4,380	16
Uranium		
With light water reactors	200	3
With Breeders	14,000	—

process (e.g., solvent extraction), and the use of catalytic hydrogenation to replace coking for improved product yields [4].

Oil Shale

Figure 5 shows a shale oil product cost of \$26.2/barrel for a 50,000 barrel per day shale oil facility based on U.S. western oil shale deposits. Unocal Corp. has been operating a 10,000 barrel per day shale oil facility located at Parachute Creek, Colorado intermittently for the past several years. They are currently in the process of modifying the facility to permit improved heat recovery from the spent shale fines [5]. Petrobras has operating a semi commercial shale oil facility in Brazil using their own shale retort technology. In Morocco, the governmental oil shale company (ONAREP), is developing an 80 ton per day demonstration project based on the semicontinuous batch T-3 retort technology at the Timahdit deposit. In Australia, projects are being considered by the Southern Pacific Petroleum Company in potential joint ventures respectively with Exxon for the Rundle deposit and Japan Australia Oil Shale Company (JAOSCO), a Japanese consortium, for the Condor oil shale deposit, both located in Queensland. It will be difficult for any of these projects to proceed without special governmental subsidy and support as long as crude oil prices remain unstable and below 25 dollars per barrel.

Coal

Commercially demonstrated technology exists for the production of high BTU gas (equivalent to natural gas in composition and heating value). In the United States, the Great Plains coal gasification facility has been operating for several years producing 125 billion BTUs per day of

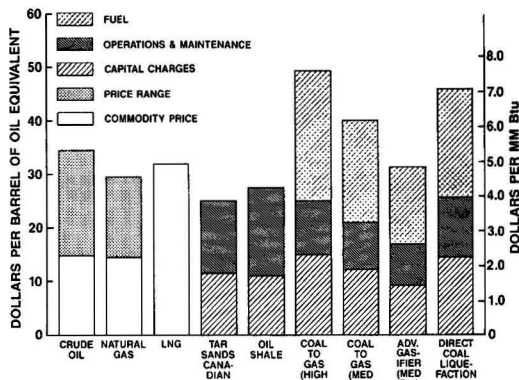


Figure 5. Comparative fuel technologies costs.

high Btu gas using the Lurgi dry bottom coal gasification technology [6]. However, the economics for production of this gas from coal in the United States are not competitive with current commercial sources of natural gas.

Technology for the production of medium Btu gas (300 to 500 Btu per cubic foot heating value) via coal gasification has also been demonstrated on a commercial scale during the past few years. A facility utilizing Texaco coal gasification to produce medium Btu gas (e.g., carbon monoxide and hydrogen) which is used as synthesis gas in the production of methanol and acetic anhydride is being operated by Tennessee Eastman in Kingsport, Tennessee [6]. Acetic anhydride is the raw material used by Eastman in producing photographic film base at the Kingsport site.

Furthermore, a 1,000 ton per day Texaco coal gasifier has been operating for several years in an integrated gasification combined cycle 100 Mwe power facility in Daggett, California owned by a consortium comprised of Southern California Edison, Texaco, the Electric Power Research Institute, GE, Bechtel, and a Japanese utility consortium [6]. Ube Industries has been operating four 500 ton per day Texaco coal gasifiers to provide synthesis gas for ammonia production in a facility in the southern part of Japan for several years. Dow Chemical is constructing a coal gasifier for operation in their Plaquemine, Louisiana chemical plant [5]. Shell Oil is constructing a 200 ton per day coal gasifier using their proprietary pro-

TABLE 2. ECONOMIC PREMISES FOR COMPARISON OF FUEL TECHNOLOGIES
(CONSTANT 1986 DOLLARS)

Fuel Technology	Facility Size	Year of Operation	Total Capital Investment (\$ Millions)	Unit Capital Investment (\$ thousands per Bbl/Day)	Unit Capital Investment (\$/Annual MM Btu)	Product Cost	
						\$/Equiv. Bbl	\$/MM Btu
Crude Oil						15-35	2.6-6.0
Natural gas						14.5-29	2.5-5.0
LNG (regasification)						32	5.5
Tar sands (Canadian)	50,000 ^(a)	1986	2,200	44.1	23.1	24.9	4.3
Oil shale (U.S.)	50,000 ^(a)	1990	2,620	52.4	27.5	26.2	4.5
Coal gasification							
Commercially demonstrated							
High Btu gas	250 ^(b)	1986	2,560	59.4	31.2	49.3	8.5
Medium Btu gas	250 ^(b)	1986	2,180	50.4	26.5	40.0	6.9
Advanced gasifier							
Medium Btu gas	250 ^(b)	1990	1,590	37.0	19.4	31.3	5.4
Coal liquefaction							
Direct	50,000 ^(a)	1990	2,630	52.5	27.6	45.8	7.9

(a) Barrels per stream day
(b) Billions of Btu's per stream day

TABLE 3. COMPARISON OF ALTERNATIVE FUEL TECHNOLOGY PRODUCT COSTS
(CONSTANT 1986 DOLLARS)

Fuel Technology	\$/Equivalent Bbl				\$/MM Btu			
	Capital Charges ^(a)	O&M	Fuel Costs	Total Product Cost	Capital Charges ^(a)	O&M	Fuel Costs	Total Product Cost
Tar sands (Canadian)	11.6	13.3	—	24.9	2.0	2.3	—	4.3
Oil shale (U.S.)	12.9	13.3	—	26.2	2.2	2.3	—	4.5
Coal gasification								
Commercially demonstrated								
High Btu gas	15.1	9.9	24.4 ^(b)	49.3	2.6	1.7	4.2 ^(b)	8.5
Medium Btu gas	12.2	8.7	19.1 ^(b)	40.0	2.1	1.5	3.3 ^(b)	6.9
Advanced gasifier								
Medium Btu gas	9.3	7.5	14.5 ^(b)	31.3	1.6	1.3	2.5 ^(b)	5.4
Coal liquefaction								
Direct	14.5	11.0	20.3 ^(b)	45.8	2.5	1.9	3.5 ^(b)	7.9

(a) Based on 20 percent equity, 80 percent debt with 10 percent aftertax return on equity, and 8 percent interest on debt.
General inflation rate for all costs is 2.5 percent per year.

(b) Coal delivered price of \$50/metric ton or \$2.50/MM Btu.

cess technology for operation in their Deer Park, Texas facility. British Gas Corporation has been operating a slagging coal gasifier in Westfield in the United Kingdom for several years based on joint developments with Lurgi and the technology is believed to be ready for commercial scale demonstration [6]. Of course, SASOL has been operating Lurgi dry bottom coal gasifiers in South Africa as part of the Fischer Tropsch facilities for the production of transportation fuels and chemical feedstocks for some years now in South Africa. Thus, there has been substantial activity in the demonstration of commercial scale coal gasification technology for the production of synthesis gas during the past decade.

Except for special situations, Figure 5 shows that the economics of producing either high Btu gas or medium Btu gas (i.e., synthesis gas) from coal are not currently competitive compared to either natural gas delivered by pipeline or regasified LNG. However, since natural gas market value is highly regional in nature, and the establishment of a network of pipelines or LNG production, transportation, and regasification facilities are highly capital intensive, judgments concerning the desirability of installing coal gasification as opposed to other alternatives require overall study in the light of the total energy situation for the region involved. For example, in China, substantial interest currently exists in utilizing coal gasification to produce town gas as a way of alleviating environmental problems in the large urban areas arising from emissions released to the atmosphere in the direct combustion of coal.

In the past, when oil supply has been tight, invariably the public has become aware of the problem through shortages in the supply of liquid fuels for transport. Motor gasoline, diesel, and jet fuels have been the materials most affected in these situations. Technology recently has been commercialized for the production of motor gasoline from synthesis gas in New Zealand via a 14,500 barrel per day facility owned by a joint venture of the New Zealand government and Mobil [4]. This facility utilizes new technology developed by Mobil for the catalytic conversion of methanol (produced from synthesis gas) to motor gasoline. Further modifications of this technology at the research and development stage may permit the production of diesel and jet fuel and also may eliminate the necessity of producing methanol as an intermediate. The feedstocks for the New Zealand facility is natural gas from the Maui gas field. However, synthesis gas could alternatively be produced via coal gasification. Thus, Mobil's

new technology permits the conversion of either natural gas or coal to liquid transportation fuels. Shell has announced new technology ready for commercialization for the conversion of synthesis gas to liquid transport fuels and they are currently considering several potential locations for commercial scale demonstration. These approaches represent the use of indirect coal liquefaction technologies. Synthetic liquid fuels are not competitive with petroleum derived transportation fuels in today's economic climate. However, special situations exist for those countries that have significant natural gas or coal reserves but little or no petroleum. In these cases, the need to conserve hard money credits in foreign trade may foster the development of indigenous projects for the production of transportation fuels from the available indigenous resources.

In spite of major development activities in direct coal liquefaction during the past decade, it has not been possible to develop and execute a project for commercial scale demonstration of this technology. Ashland Oil operated a 300 ton per day H-coal pilot plant facility for several years in Cattlettsburg, Kentucky. This plant is now mothballed [6]. Exxon operated a similarly sized pilot plant on donor solvent technology in Baytown, Texas for several years. This facility has now been dismantled [6]. A 6 ton per day pilot plant has been operating in Wilsonville, Alabama to demonstrate two-stage coal hydrogenation technology [6]. A 150 ton per day solvent refined coal pilot plant operated for several years in Tacoma, Washington.

In West Germany, Ruhr Kohle has been operating a 200 ton per day direct coal liquefaction pilot plant at Bottrup [6]. Japan, motivated by a desire for diversification of energy supplies, has launched a major development in direct coal liquefaction. A 50 ton per day brown coal liquefaction facility is operating in Victoria, Australia with Japanese financial support. Furthermore, a 250 ton per day Nedol process demonstration plant is being designed and will be built in northern Japan [4]. Figure 5 shows a potential cost of synthetic crude oil of about \$46/barrel from the operation of a 50,000 barrel per day direct coal liquefaction commercial scale facility. Commercialization of direct coal liquefaction is not likely to occur until the mid 90s at the earliest unless unforeseen major changes occur in the world energy situation.

COMPARATIVE COSTS OF ELECTRIC POWER GENERATION

In many countries, the combustion of residual fuel oil

TABLE 4. ECONOMIC PREMISES FOR COMPARISON OF ALTERNATIVE ELECTRIC POWER GENERATION STATIONS
(CONSTANT 1986 DOLLARS)

Item	Power Plant						
	Oil-fired	Regasified LNG	Nuclear	Pulverized Coal With Scrubber	Pulverized Coal Without Scrubber	Coal Gasification Combined Cycle	Coal Atmospheric Fluidized Bed
Plant size, number of units × MWe each	2 × 550	2 × 550	1 × 1100	2 × 550	2 × 550	2 × 550	2 × 550
Total capital requirement including owner's costs, \$ millions	910	610	1,960	1,330	1,140	1,650	1,300
Dollars/kWe including interest during construction	715	510	1,630	1,130	970	1,390	1,100
Dollars/kWe without interest during construction	670	480	1,490	1,040	880	1,310	1,025
Delivered fuel price, \$/MM Btu	2.00-4.80	5.50	0.75	2.50	2.50	2.50	2.50
Plant capacity factor, %	65	65	75	65	65	80	65

TABLE 5. COMPARISON OF ALTERNATIVE ELECTRIC POWER GENERATING COSTS
(CONSTANT 1986 DOLLARS)

Item	Power Plant							
	Oil-Fired \$12/bbl ^(a)	Oil-Fired \$28/Bbl ^(a)	Regasified LNG	Nuclear	Pulverized Coal with Scrubber	Pulverized Coal without Scrubber	Coal Gasification Combined Cycle	Coal Atmospheric Fluidized Bed
Plant size, number of units × MWe each	2 × 550	2 × 550	2 × 550	1 × 1100	2 × 550	2 × 550	2 × 550	2 × 550
Time elapsed from start of engineering to commercial operation, years	4	4	4	7	5	5	4	4
System life, years	30	30	30	30	30	30	30	30
Total capital investment including owner's costs, \$ millions	910	910	610	1,960	1,330	1,140	1,650	1,300
Levelized cost of electricity at the bushbar, mills/kWh								
Capital charges ^(b)	11.8	11.8	6.6	23.0	17.3	14.3	17.8	16.9
O&M	1.1	1.1	1.0	8.0	8.0	6.0	6.2	9.1
Fuel costs	20.5	47.8	57.0 ^(c)	7.6 ^(d)	24.3 ^(e)	23.8	22.9	24.2
Total	33.4	60.7	64.6	38.6	49.6	44.1	46.9	50.2

(a) Residual fuel oil price

(b) Based on 20 percent equity, 80 percent debt with 10 percent aftertax return on equity, and 8 percent on debt. General inflation rate for all costs of 2.5 percent per year

(c) LNG price delivered of \$5.50/MM Btu

(d) Based on yellow cake (U₃O₈) price of \$30/lb and including enrichment, fuel fabrication, reprocessing, and disposal

(e) Coal delivered price of \$50/metric ton

is the primary method for electric power generation. In Japan, regasified LNG, steam coal, and nuclear power are all important power technologies in use. Japan has 29 nuclear power plants in operation with 20 more under construction or at the advanced planning stage. Both Korea and Taiwan have important nuclear power programs with nine nuclear plants in Korea and, five in Taiwan. Japan has cut its dependence on imported oil for electric power generation (principally from the Middle East) from 90% before the 1974 oil crisis to 55% of total fossil fuels consumed as a result of its programs to diversify energy types and sources [7].

To provide perspective for decision making concerning capital expenditures for new power generating facilities, each power technology is evaluated on a consistent economic basis in constant 1986 dollars so as to show the relative cost of power from each technology. Of equal importance, the portions of the costs which are directly attributable to fuel, operations and maintenance, and cap-

ital charges are developed. The economic premises for this analysis are shown in Table 4. In developing these comparisons, facility sizes have been selected in multi unit combinations totaling 1,100 MWe for each technology.

Figure 6 and Table 5 show the comparative electricity generating costs for each of the options considered. At a residual fuel oil price of \$12/barrel, about 60% of the cost of electricity generated from oil fired power stations is attributable directly to the cost of the fuel. At a residual fuel oil price of \$28/barrel, about 80% of the cost of electricity generated from oil fired power stations is attributable to fuel cost. For existing generation capacity with the capability to burn either oil or coal, a substitution or crossover point from oil to coal (or vice versa) occurs at a price of about \$14/barrel for oil (referenced to \$2.50/MM Btu for coal).

For regasified LNG at a delivered price at the point of use of \$5.50 per million Btu's, approximately 87% of the

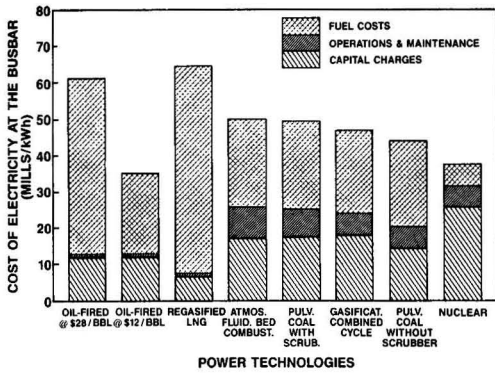


Figure 6. Comparative electric generation costs (constant 1986 dollars).

cost of the electricity generated is directly attributable to the cost of the fuel (based on utilization in combined cycle turbines).

For electricity generating technologies based on consuming coal, at a delivered price of \$2.50 per million Btu's, the cost of fuel represents about half of the cost of electricity depending upon the particular technology involved. In the case of the cost of electricity is attributable to the cost of nuclear fuel (with yellowcake at \$30 per lb.) including the costs for the elements of the fuel cycle required for spent fuel reprocessing and subsequent disposal of the high radioactive level residues. Furthermore, it is apparent that electricity produced from new generating stations using any of the coal technologies is substantially less costly than that from either the combustion of regasified LNG.

In spite of the higher comparative cost of using LNG as a fuel source, its use in Japan is increasing substantially. In fact, total LNG imports to Japan increased 40% from 1983 to 1984. Much of the increase was attributable to the increased use of LNG in electric power generation. The use of LNG is likely to continue increasing because of its quality as a clean burning fuel for environmentally sensitive areas and its value as a feedstock. Security of supply dictates reliance on a variety of fuel types, including LNG.

It is evident that nuclear power is an attractive alternative for those countries that can go this route. The key to these economics is a stable regulatory environment that permits the engineering design to be frozen after the necessary review by regulatory agencies. This permits project completion to occur in seven years or less without the serious cost escalation which has occurred on some recent U.S. projects.

It is important to recognize that capital investment requirements can be projected with a reasonable level of certainty for new power stations where the technology

has already been commercially demonstrated. Greater uncertainties exist in estimating fuel costs over a thirty year facility life. A range of residual fuel oil prices of from 12 to \$28 per barrel has been considered in this analysis.

Several of the coal technologies evaluated have not yet been operated at the 550 MWe capacity level. The coal gasification combined cycle technology has been demonstrated at a 100 MWe capacity level in the Coolwater project in Daggett, California with excellent operating results. It is anticipated that by the mid 1990's, similar units at the 550 MWe capacity level will have been demonstrated in commercial operation. Atmospheric fluidized bed combustion units at the 100 to 200 MWe capacity level are currently being designed for installation at several different locations in the United States including the Gilberton, Colorado-UTE, and TVA facilities.

The principal advantages of these new technologies reside in their capability for improved emission control for SO_x , NO_x , and nonleachability of the resultant coal ash [8]. Table 6 provides a comparison of emissions from integrated coal gasification combined cycle, atmospheric fluidized bed combustion, and pulverized coal fired boilers (with stack gas scrubbing) based on utilization of a 3 weight percent sulfur coal. It is clear that for environmentally sensitive areas, integrated coal gasification combined cycle technology is superior to the other coal consumption technologies. Furthermore, facilities using this technology are constructable in less time and operate with a higher availability (80% capacity factor) than most of the other technologies involved. It is important to note that actual capital requirements are strongly dependent on site specific conditions so that the conclusions reached in this analysis required verification for any specific project.

SUMMARY

Fluctuations in the price of crude oil and refined petroleum products have created instability in energy markets. The development of alternative fuel technologies requires a continuity of substantial financial commitments over periods as long as 10 years or more. The commercialization of these technologies involves not only accurate capital investment forecasts but also projections of commodity fuel prices over a 20 to 30 year period (e.g. the life of the project). This paper shows that crude oil prices greater than \$25 per barrel (in 1986 dollars) are necessary for commercial scale projects to be developed using any of the alternative fuels technologies (except for special situations involving governmental subsidy and/or policy).

The economics of the production of synthetic crude oil from Canadian tar sands and shale oil from U. S. western oil shale appear to be somewhat better than coal conversion to either gas or liquids. However, depending upon natural gas source and market value, synthesis gas produced via coal gasification can be a desirable feedstock source for the production of various chemicals.

The economics of electricity generation favor the use of

TABLE 6. TYPICAL EMISSIONS FROM 3% SULFUR COAL IN ELECTRIC POWER GENERATION

Technology	SO_x (Tons/MWe Yr)	NO_x (Tons/MWe Yr)	Solid Wastes (Tons/MWe Yr)
Pulverized Coal-Fired without Scrubbers	120	25	240
Pulverized Coal-Fired with Scrubbers (90% Removal)	12	8	650
Integrated Gasification Combined Cycle	0.14-3.5	3	300
Atmospheric Fluidized Bed Combustion	6	4	1,000

coal and nuclear power for new electric generation capacity as long as residual fuel oil price remains above \$14 per barrel and natural gas price remains above \$3.50 per million Btu (versus steam coal at \$2.50 per million Btu). New technologies such as integrated coal gasification-combined cycle and atmospheric fluidized bed combustion of coal have a significant potential advantage in greatly reduced emissions of SO_x and NO_x pollutants for those situations with strongly restrictive emission control regulatory requirements.

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Use of Liquid CO₂ for Fine Coal Cleaning

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A novel process for fine coal cleaning which takes advantage of the preferred wettability of "clean" coal by liquid CO₂ to produce a low ash and low sulfur content product.

INTRODUCTION

During the past few years, a great deal of interest has been developed in fine coal cleaning techniques. The concern about the environmental effect of sulfur dioxide emissions and the increasing proportion of fine coal (-28 mesh) in the run-of-mine product are some of the reasons for the increased research in this area. Indeed, there is a national effort to develop technologies that will produce "super-clean" coal, i.e., coal containing less than 2% ash and 0.8% sulfur. The cleaned coal could then be used as a replacement for oil and natural gas.

Several chemical and physical coal cleaning techniques have been suggested for the production of super-clean coal. These include the TRW Gravimelt Process, the Microbubble Flotation, and the Otisca Process [1, 2, 3]. The LICADO (LIquid CARbon DiOxide) process, currently being studied at the University of Pittsburgh, is a new alternative in this field.

The LICADO process uses liquid CO₂, at about 850 psig and room temperature, to separate mineral matter from coal. The operation is carried out by contacting a coal-water slurry with liquid CO₂. Because of the differences in their surface properties, coal particles (hydrophobic) are preferentially attached to the liquid CO₂-water interface and then transferred to the liquid CO₂ phase while the mineral particles (hydrophilic) remain in the aqueous phase.

The main advantages of the LICADO process are:

1. The low cost of liquid CO₂ (\$65/ton) reduces the overall process cost [4].
2. The fugitive CO₂ poses no environmental problem.
3. Carbon dioxide is readily available in large quantities and, it can be generated on-site by burning coal, if necessary.
4. Carbon dioxide is adsorbed on clean coal acting as an inert blanket that minimizes the pyrophoric tendency of ultrafine coal in storage or during transit [5].
5. This process can be incorporated as an integral part of a liquid CO₂/coal slurry pipeline system for transporting coal [6].
6. The low viscosity and surface tension of liquid CO₂ permit its separation easily from clean coal with very low operation costs.

There are also two disadvantages that should be recognized:

1. The process is carried out under pressure (up to 900 psig) which means a high equipment cost.
2. Although CO₂ is not toxic, it must be used with care because of its asphyxial effect.

The technical and economical feasibility of the LICADO process will be determined by the research project now underway. The process has been tested at laboratory scale (between 40 and 2000 ml operation volume). Initial experimental results suggest that the LICADO process is a promising new method for fine coal cleaning.

PROCESS PRINCIPLES

The LICADO process is a surface-property driven process in which the liquid CO₂-water interface is used for the differentiation and separation between clean coal and mineral matter.

Similar concept is widely used in flotation where air bubbles constitute the moving interface. The attractive advantage of air is that it can be easily removed, leaving the solids in two aqueous slurries. However, fine particle slurries thus produced often require costly dewatering operations in some industrial operations.

This disadvantage can be overcome by the use of a water immiscible liquid instead of air. The oil agglomeration process uses this principle to form aggregates of clean coal particles that can be easily removed from the aqueous phase. Nevertheless, in these type of techniques, the removal of the binding liquid (oil, fluorocarbons, etc.) is also a complex and costly operation.

Liquid CO₂, on the other hand, has a very low viscosity (0.07 cP) and surface tension (0.5 dyne/cm) which permits an easy separation from coal particles.

In preliminary experiments, it was observed that coal particles are completely "wettted" by liquid CO₂ while the mineral matter is preferentially wetted by water. Thus, when liquid CO₂ is dispersed into a coal-water slurry, it tends to form agglomerates with the "clean" coal particles and to float them to the liquid CO₂ phase (specific gravity = 0.71). The mineral particles tend to remain in the aqueous phase as refuse.

The mechanism of the separation will then be governed by the interaction between the interfacial, shear and body forces present in the coal particle-liquid CO₂-water system.

The interfacial interactions will depend on the surface composition and morphology of coal particles. By applying Young's equation to the system, the following inequalities can be postulated [7]:

$$\theta_C > \theta_F > \theta_R$$

where θ is the contact angle between water/liquid CO_2 (measured in the water phase) on coal, and the subscripts C, F and R indicate clean coal, feed coal and refuse respectively. As previously discussed, this means that clean coal is more wettable by liquid CO_2 while refuse is more wettable by water.

The structure and composition of the agglomerates will be determined by the surface properties of the individual coal particles as well as by the hydrodynamic conditions in the separation vessel.

Because of the unique properties of liquid CO_2 , shown in Table 1, the effect of mixing, injection mode and flowrates can not be predicted based on literature data of other systems. The influence of these variables on the process performance must, therefore, be determined experimentally.

EXPERIMENTAL

A clean coal with ash content lower than 2% was obtained in the initial tests of the LICADO process. Based on such encouraging results, further experiments were carried out at laboratory scale. High pressure units with operating volumes ranging from 40 ml to 2000 ml have been used in this study.

Equipment Description

The initial test of the LICADO process was carried out in a glass cylinder (0.85 inch I.D.) enclosed in a high pressure vessel. This test cell was later replaced by a Batch Research Unit (BRU). The BRU, shown in Figure 1, is composed of two high pressure cells (2-inch I.D.) connected by a 2-inch ball valve. The ball valve allows quick and complete isolation of the two cells, so that clean coal and refuse can be removed completely without remixing of the products when the system is depressurized.

The size of the valve opening is chosen to be the same as the internal diameter of the cells so that there is no obstruction to the transfer of clean coal and refuse between the upper and lower cells during the operation.

The system is provided with two variable speed mixers (up to 1200 rpm) for agitation of both the aqueous and liquid CO_2 phases. A bottom injection port is provided for introducing liquid CO_2 to the test cell.

TABLE 1. PHYSICAL PROPERTIES OF CO_2

	Reference
Critical Temperature:	31.04°C (10)
Critical Pressure:	72.8 atm (10)
Liquid CO_2 Properties at 25°C	
Vapor Pressure:	63.45 atm (932.7 psia) (10)
Surface Tension	0.5 dynes/cm (11)
Viscosity:	0.064 cP (12)
Density:	0.71 g/cm ³ (10)
Liq. CO_2 /Water	28 dynes/cm*
Interfacial Tension:	
CO_2 Solubility in Water:	5.5% (13)
Water Solubility In Liquid CO_2 :	0.1% (14)

* Measured in our laboratory

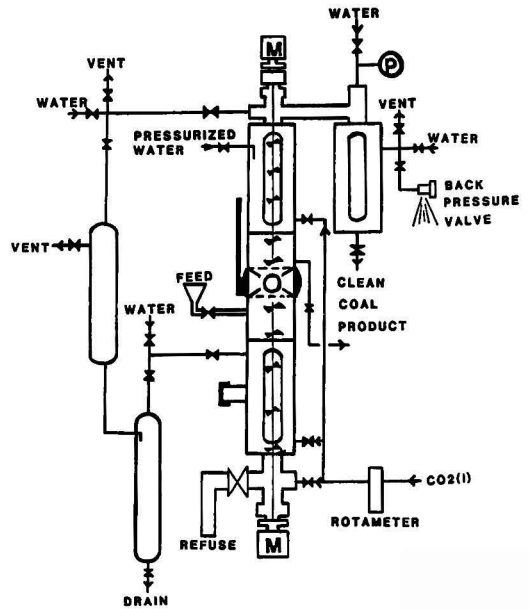


Figure 1. Batch research unit.

Experimental Procedure

The experiments are carried out at room temperature. A coal-water slurry is prepared by mixing coal and water for sufficient time to attain a complete dispersion and wetting of the coal particles. The slurry is fed to the test unit prior to pressurizing the system. The operation pressure is 850 ± 50 psig (CO_2 vapor pressure at room temperature).

Approximately one liter of coal-water slurry is used for each test run. Liquid CO_2 is injected from the bottom while both phases are mixed. After a pre-determined time, the liquid CO_2 injection and the agitation are stopped and phases are allowed to separate. The center valve is then closed and the clean coal product and the refuse fraction are removed.

The experimental range of the operation variables studied in the batch research unit is shown in Table 2. The ash contents of feed coal, clean coal and refuse are determined for every experiment using the standard ASTM [8] method. In some cases the sulfur content and particle size distribution of the samples are also determined using a MINECO Rapid Sulfur Analyzer [9] and a Leitz TAS-Plus image analyzer, respectively.

RESULTS AND DISCUSSION

Three different coals, Upper Freeport coal, Pittsburgh-Bruceton Mine coal and Illinois No. 6 coal were used for the initial tests. These experiments were carried out in

TABLE 2. OPERATING VARIABLES FOR THE 2-INCH BATCH RESEARCH UNIT

Variables	Range
Slurry Concentration, wt%	3-17
Liquid CO_2 Flowrate, ml/min.	20-160
Injection/Mixing Time, min.	5-35
Top Mixing Speed, rpm	0-1050
Bottom Mixing Speed, rpm	0-1200

TABLE 3. INITIAL TEST RESULTS IN 40 ML GLASS TEST CELL

Coal	Ash Content %			Yield %	
	Feed	Product	Refuse	Overall	Clean coal
Pittsburgh-Bruceton Mine (-32 mesh)	4.7	2.3	16.4	83	85
Illinois No. 6 (-100 mesh)	4.7	3.0	22.3	91	93
Upper Freeport (-35 mesh)	16.2	6.6	31.8	62	69
	7.2	1.6	—	—	—
	34.4	12.9	61.4	56	74
	34.4	14.8	92.9	75	97

the glass test cell. The experimental results, shown in Table 3, confirmed the effectiveness of the LICADO process. Indeed, the ash content of clean coal product can be reduced to as low as 1.6% for Illinois No. 6 coal and 2.3% for Pittsburgh-Bruceton Mine coal. For Upper Freeport coal, the ash content was reduced from 34.4% to less than 15% with an overall yield of 75%.

The experiments carried out in the batch research unit have been designed to study the influence of operating variables on the clean coal ash content and product yield. Upper Freeport coal was chosen as the main test coal because it has a sufficiently large ash content so that the effect of operating parameters on the quality of clean coal can be easily discerned in terms of differences in ash content of the product.

A series of experiments was carried out by varying one parameter at a time while keeping the others constant. The results are presented below.

The effect of the liquid CO₂ injection time on the performance of the process is shown in Figure 2. As expected, longer operation time results in a higher yield and an increase in the product ash content. The latter is caused by the increase in the slurry ash content as clean coal is removed.

The experimental results show a small effect of the coal-water slurry concentration on the separation performance. The differences in the clean coal ash content are less than 1% while the product yield varies within less than 5% when the feed slurry concentration is

changed from 3 to 17%. These results, shown in Figure 3, indicate that the case of even high slurry concentration should be explored. The use of more concentrated slurries translates into larger operating capacity per unit volume.

Figure 4 shows that lower liquid CO₂ flowrate improves the clean coal quality without significantly decreasing the product yield. It is speculated that a restricted availability of liquid CO₂ in the slurry limits the agglomeration of ash particles. Consequently, cleaner product is obtained.

The effect of the mixing speed in both the liquid CO₂ and water phase is depicted in Figures 5 and 6. It can be seen from Figure 5 that by increasing the liquid CO₂ dispersion in the aqueous phase (i.e., increasing the mixing speed in the water phase), and therefore, the interfacial area available for particle-droplet contact, a higher product yield is obtained. The product yield increases from 10% to 65% as the agitation speed is increased from 0 to 800 rpm with no significant change in the clean coal ash content.

The mixing in liquid CO₂ phase, however, has a different effect on the separation process. It is observed in Figure 6 that cleaner coal is obtained at mixing speed between 400 and 700 rpm, while the yield decreases linearly from 74% at 0 rpm down to 61% at 1050 rpm. These results are believed to arise from a balance between the tendency to release ash particles from the coal-liquid CO₂ agglomerates by the shear forces and the re-entrainment

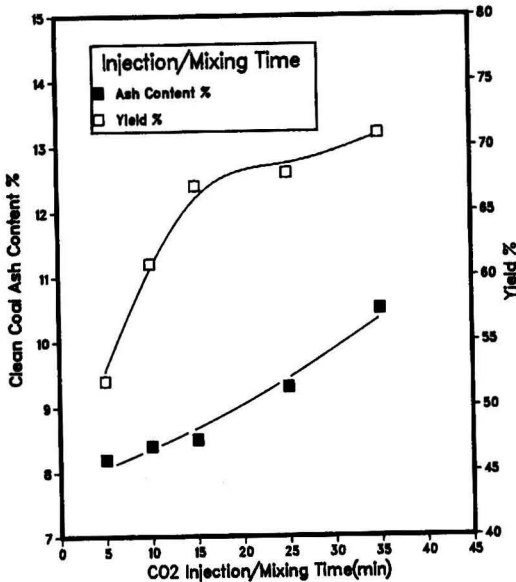


Figure 2. Effect of CO₂ injection/mixing time on separation.

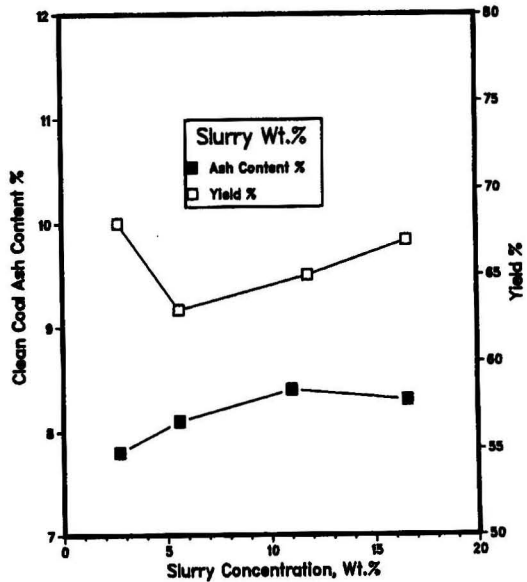


Figure 3. Effect of slurry concentration on separation.

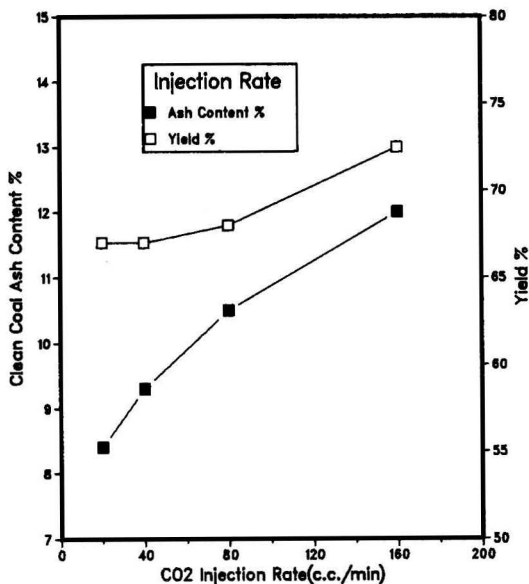


Figure 4. Effect of CO₂ injection rate on separation.

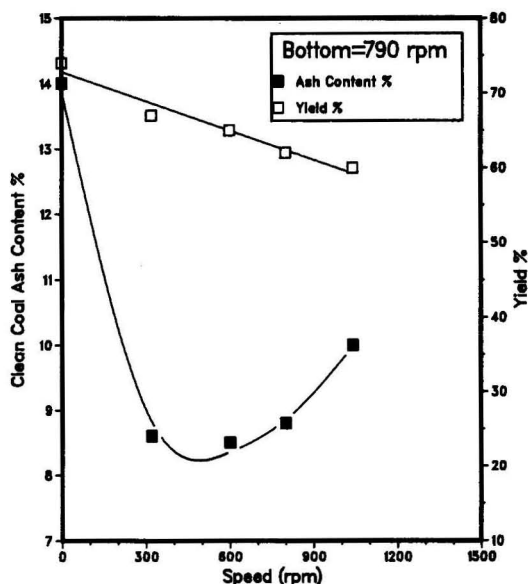


Figure 6. Effect of mixing speed in liquid CO₂ phase on separation.

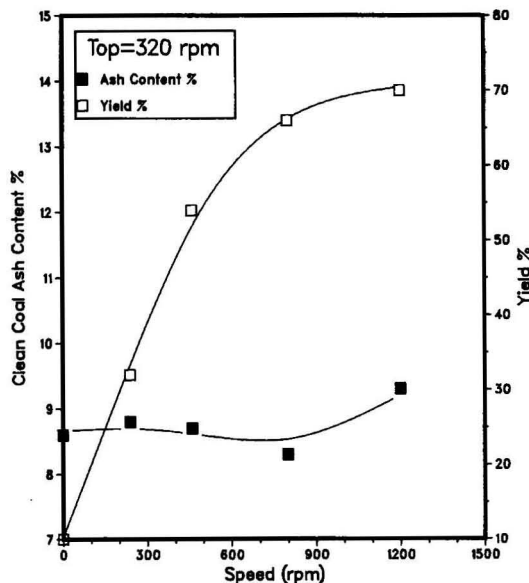


Figure 5. Effect of mixing speed in water phase on separation.

of ash particles into the liquid CO₂ phase caused by the turbulent mixing near the interface.

Pittsburgh-Bruceton Mine coal has also been tested in the batch research unit. The experimental results are summarized in Table 4. These experiments were carried out under semi-continuous conditions, in which the clean coal was continuously transferred to a sampler connected to the top of the cell. The clean coal ash content was reduced from 4.7% down to 1.5%. A low yield of 43% was obtained because of incomplete removal of clean coal from the interface.

A significant reduction in the sulfur content was also

observed in the cleaning process. As shown in Table 5, for Upper Freeport coal, the total sulfur content is reduced from 1.30% down to 0.74% in one single step. The remaining sulfur is believed to be mostly of the organic type which is not removable by physical cleaning processes.

CONCLUSIONS

Although the study of the LICADO process is far from complete, a number of preliminary conclusions may be drawn:

1. The LICADO process can be used to produce clean coal with ash content less than 2.0%.
2. There is no significant effect of the slurry concentration on the process performance in the range studied. Further tests using concentrations higher than 20% is suggested.
3. For Upper Freeport coal, a significant reduction in pyritic sulfur content (estimated to be 80% reduction) was achieved in one cleaning step.

TABLE 4. SEMI-CONTINUOUS TEST PITTSBURGH-BRUCETON MINE COAL, -200 MESH

Feed Ash Content	: 4.7%
Clean Coal Ash Content	: 1.5%
Refuse Ash Content	: 7.1%
Overall Yield	: 43%

TABLE 5. EXPERIMENTAL RESULTS UPPER FREEPORT COAL, -200 MESH

	Feed coal	Clean coal	Refuse
Ash content, %	23.5	8.4	54.2
Sulfur content, %	1.30	0.74	2.44
Mean particle size, μm	19	17	26
Density, g/cm ³	1.45	1.31	1.88

- The mixing plays a pivotal role in influencing the performance of the LICADO process in terms of both product yield and clean coal quality. Further study should be carried out to determine the optimal mixing condition.

The study of the LICADO process will be continued by developing a continuous operation system. The experimental data will be used to evaluate the technical and economical feasibility of this novel coal beneficiation technique.

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Commercial Application of Wood Derived Oil

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Commercial scale-up of an entrained flow, fast pyrolysis system will be installed at a pulp and paper mill. The oil produced will be utilized in a combustion turbine/combined cycle electrical generation project and for the replacement of natural gas in a lime kiln. Research into cracking kinetics will be part of the project.

INTRODUCTION

Pyrotech Corporation is a company whose purpose is to develop energy projects for large commercial or small utility customers. Pyrotech's participation is the turn-key evaluation, design, construction, funding, and operation of power related projects for end users. The project may be entirely owned and operated by Pyrotech or owned as a joint venture with the end user. All project economics are based on conventional technology which is readily available. However, some proprietary Pyrotech technology will be demonstrated in most projects in order to research new ideas and applications of wood derived oil.

THE PROJECT

A west coast pulp mill was unable due to equipment limitations, to produce sufficient electricity to meet an existing contract for sale of electricity. The contract has been signed to export/sell up to 25 MW of electricity to the utility. The pulp mill signed the contract anticipating an increase in its ability to produce electricity and installed a total turbine capacity of 50 MW. Twenty-five MW is used internally and 9 to 15 MW was sold to the utility. After installing a new tie-line in 1985 in order to export an additional 10 MW it was found that the existing waste wood and bark (hog fuel) boiler systems were unable to provide the additional steaming capacity necessary for the production of the additional 10 MW. At this time Pyrotech was asked to evaluate the conditions and present a plan that provided a solution to the problem at minimum risk and capital cost to the pulp mill. Four options were evaluated which included:

1. fuel drying to increase existing boiler efficiency
2. new waste wood boiler
3. third party boiler or gasifier/boiler project
4. PYROTECH PLAN — which is a combined cycle system utilizing combustion turbine and package boiler to produce 150,000 # of steam per hour.

The Pyrotech plan provides the most fuel efficient option with the proposed combined cycle converting wood to electricity at an estimated 40% thermal efficiency compared to 20% for a wet hog fuel system. This plan will screen the existing wood fuel which is being used in the hog fuel boilers and then dry the smaller wood particles to 15% moisture content utilizing waste heat from a new package boiler and from the existing hog fuel systems. The wood will then enter a pyrolysis system which produces a synthesis gas and oil vapors. The vapors which

are not directed to the boiler will be condensed and made available to run the combustion turbines. Turbine exhaust at approximately 700°F and 16% oxygen will provide the combustion air to be used with the synthesis gas. Since the proprietary pyrolysis units are a relatively new technology, a multi-fuel burner will be used which can be fired independently as a suspension burner for the fine dry wood particles. By providing a backup system which is commercially proven, the project provides minimum risk in the production of steam. Additional backup systems of stored wood oil, #4 fuel oil, and natural gas provide five individual fuel systems which can each independently fire the boiler. All components of the project are modular with the longest lead item, the boiler, available from stock with the steam conditions matching the existing turbine system. The installation of a new package boiler system will provide for significant savings in time of construction and capital expenditure. It is anticipated that the total steam system will be on-line within an eight month period from start of construction.

The engineering, construction, financing and operation of the project will be completely overseen by Pyrotech. The pulp mill incurs a minimal risk in the project and no hard capital cost. In return for a share of the earnings from the project, Pyrotech was able to present a plan designed to meet the criteria established by the project and provide a reliable, efficient system with the capacity necessary to fulfill the utility contract.

RESEARCH ASPECTS OF THE PROJECT

Wood oil fired combustion turbine system

A significant additional benefit will be realized from the project by the pulp mill. The pyrolysis units or petroleum synthesis units (PSUs) which will be used can produce a high percent of light oil vapors, up to 70% [4]. This vapor produces a medium Btu/cu ft gas stream for use in the boiler, while a condensing system will give this design the ability to produce a primary oil which can be used to fire the combustion turbines and increase the overall system efficiency [10].

The redundant fuel backup system enables this project to scale up and prove new proprietary entrained flow fast pyrolysis system in a large commercial application and not risk the project profitability if downtime on the new technology is more than anticipated.

The ability of storing wood as a liquid and utilizing it in conventional oil systems provides several interesting

possibilities to the pulp mill. First, it makes available the use of the highest efficiency conventional means of generating electricity — combined cycle; “They are the most efficient of all thermal power plant designs, converting up to 47% of the fuel energy into electricity” [11]. The use of pyrolysis oil as a fuel for turbine engines has been investigated and shown to be “96% efficient as conventional JP-4” (Figure 1) by Teledyne Corporation in their standard Model J69-T-29 combustor which is utilized in the J69 turbine series [9, 10]. The ability to produce electricity from “a state of the art” conventional system (Figure 2) could help reduce cost of wood fuel purchased by the pulp and paper industry. The wet hogged fuel boiler system which is typically used by the industry is substantially less efficient than a combined cycle approach. The improved efficiency of this system will provide an option to each mill for the additional sale of electricity to the utilities under the existing PURPA regulations.

One major energy expense which is incurred in the production of most pulp is in the operation of conventional lime kilns. For many years this industry has investigated the possibility of producing lime by using wood as a fuel instead of oil and gas. However, because of the ash associated with conventional wood combustion, this option has not been available. A substantial ash loading of the lime could possibly produce scaling conditions in the digester and evaporators and temporarily shut down a plant. The typical wood oil which is produced (Table 1) has an extremely low ash content and will greatly reduce the potential scaling and allow the pulp mill to use wood for lime production [4]. One aspect of the wood oil conversion of a lime kiln is the fact that most of the kilns already have the ability to fire oil. A few modifications at a relatively small capital cost would be necessary to convert the kiln to wood oil. The wood oil has a viscosity similar to #4 fuel oil and combustion characteristics similar to kerosene. It is an oxygenated oil which is acidic and a Btu content similar to wood (9000 Btu per pound). The PSUs which will be installed at this project have additional capacity to produce 45 million Btu/hr of condensed

oil above the energy required for the electrical production. This oil will be stored and then burned in the lime kiln, substituting for the natural gas now being fired.

All phases of the wood oil use will be monitored and evaluated in the lime production. The ability to convert wood to oil will provide the pulp industry with an option to fossil fuels now being used. Wood oil can provide a low cost fuel which can be produced conventionally from a raw material which the pulp mills generally control.

Oil upgrade for vehicle use

While the wood oil which is produced might theoretically be used directly in a conventional diesel engine, it is relatively acidic and would require modifications to tanks, fuel lines, etc. in order to be utilized. Extensive research and development has been done at several major institutes in the United States and Canada to upgrade these pyrolytic oils catalytically to conventional gasoline and diesel fuels [1, 2, 3, 6].

As part of the research to be done by Pyrotech, a continuous slip stream of the oil vapors will be directed to a process demonstration unit catalytic reactor. Various commercially available catalysts will be tested in order to evaluate their potential for upgrading these oils. After selection of the catalysts which produce the desired hydrocarbon products, an evaluation of the reactivation characteristics of each will be made. It is anticipated that after selection of the best performing catalysts that the PDU will be put into continuous operation in order to evaluate scaleup potential and production cost analysis. The bench scale studies done by various institutions have been positive for the production of gasoline type liquids, from some tested catalysts [6, 7, 8]. Reactivation has also shown to be successful for some of these catalysts [6]. Once the small reactor is put on line continuously, Pyrotech will initiate a vehicle performance test to document the “real world” use of the fuel.

The probability that a vehicle type fuel can be produced from the oxygenated wood oil appears good. The demonstration of this technology in a continuous flow process demonstration unit will produce the data that is necessary to qualify the cost of production and conversion potential of scaling up the process to produce a “gasoline type” fuel from wood.

The ability for the pulp and paper industry to produce a fuel which can be used in transportation will be a significant accomplishment and can possibly contribute greatly

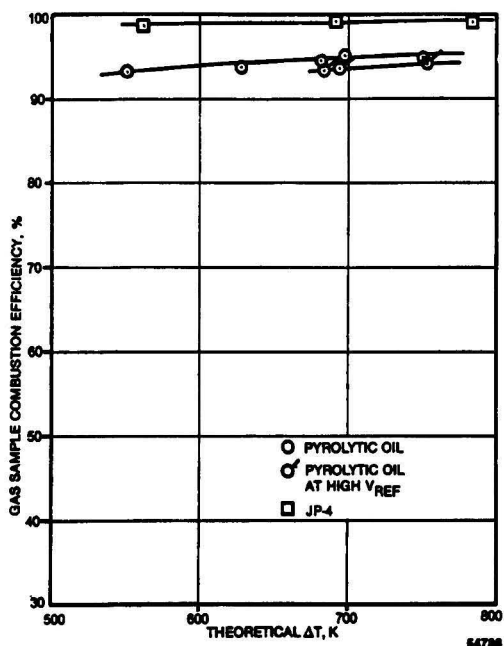


Figure 1. Variations of combustion efficiency with combustor temperature rise for pyrolytic oil and JP-4.

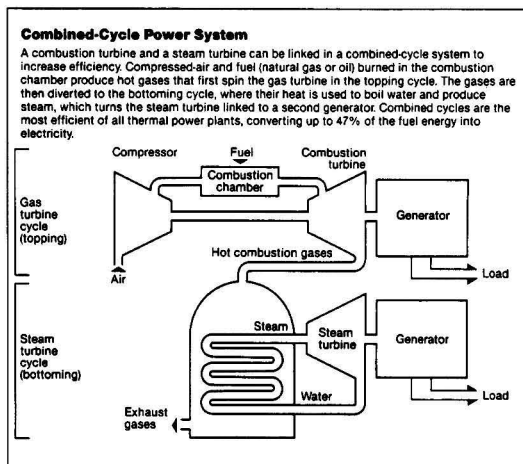


Figure 2. Wood oil fired lime kiln operations.

TABLE 1. DATA ACQUIRED FROM SERI (5)

Material	Moisture (wt %)	C	wt %* (dry basis)		Ash	Empirical Formula	Volatile wt %	(Dry)
			H	O				HHV, kJ/g (Btu/lb)
Feed	6.6	50.97	6.01	42.47	0.47	CH _{1.4} O _{0.62}	81.39	20.1 (8637)
Primary Oils**								
Run 35 (cyclone scrubber)	15.3	54.43	5.73	39.77	0.05	CH _{1.3} O _{0.55}	77.83	20.3 (8735)
Run 40 (cyclone scrubber)	8.6	57.40	5.87	36.68	0.02	CH _{1.2} O _{0.48}	81.93	22.8 (9793)
Run 40 (packed scrubber)	9.4	56.95	5.62	37.11	0.32	CH _{1.2} O _{0.49}	77.59	22.2 (9548)

to a reduction of costs associated with the forest products industry.

CONCLUSION

The primary objective of the project to produce steam at an economical rate is being initiated for the pulp mill. The project provides a minimum risk by utilizing conventional fuel systems as back-up to the new technology. Pyrotech is proceeding with this project at no capital cost to the pulp mill and has been able to receive financing based upon the sales contracts for the energy to be produced.

By utilizing conventional technologies as the basis for the economics of the project Pyrotech has also been able to initiate a research and development project on a large commercial scale. The research results of this project could have a major impact upon the pulp and paper industry in the future. The ability to convert a raw material which is typically controlled by the industry into a liquid fuel for storage or use in conventional systems could provide flexibility, additional sales potential, saving and energy independence for many in the forest products industry.

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WIPP — A Geologic Nuclear Waste Repository Nearing Construction Completion

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Design and construction status of the Waste Isolation Pilot Plant near Carlsbad, New Mexico for the disposal of defense transuranic waste.

BACKGROUND

The Waste Isolation Pilot Plant (WIPP), as authorized by Public Law 96-164, is a defense activity of the Department of Energy designed for the express purpose of providing a research and development facility to demonstrate the safe disposal of radioactive wastes resulting from the defense activities and programs of the United States. The WIPP is exempted from regulation by the Nuclear Regulatory Commission. The site, located in the southeastern corner of New Mexico about 25 miles east of Carlsbad, resides in the midst of a very extensive bedded salt formation which is approximately 3000 feet thick. The WIPP waste storage horizon is at a depth of 2150 feet below the surface about in the center of the salt beds.

The WIPP is designed to receive and retrievably store defense transuranic (TRU) waste, and to conduct experiments with defense high-level waste (DHLW). All DHLW must be retrieved and shipped off the WIPP site at the completion of the DHLW experiments. A decision to permanently dispose of the defense transuranic wastes at the WIPP site may be made at the end of a five-year operations demonstration period.

Nuclear waste is received and handled at WIPP either as contact-handled (CH: no shielding required) or remote-handled (RH: shielding required), based on the surface dose rates of the waste containers. The detailed requirements for the waste to be shipped to and disposed of at the WIPP are defined in the WIPP waste acceptance criteria. The CH waste is shipped in special shipping containers called TRUPACTs (TransUranic PACKage Transportable), and the RH waste is shipped in more heavily shielded shipping containers. When the waste is received at WIPP, it is taken into the waste handling building where the shipping containers are opened and the waste containers removed. The CH waste containers are placed on pallets, moved onto the waste hoist for lowering underground and then transported to and stacked in the waste storage rooms. The RH waste containers are removed into a hot cell, where they are handled remotely; then they are loaded into a shielded facility cask for movement onto the waste hoist and underground lowering. In the underground, the facility cask is transported to the storage area where the waste container is placed in a horizontal storage hole in the wall of the storage room.

SITE AND PRELIMINARY DESIGN VALIDATION PHASE

The construction of the WIPP has proceeded in two phases. The first phase is the Site and Preliminary Design Validation (SPDV) Phase, and the second phase involves Full Facility Construction (FFC). The SPDV Phase construction was initiated in April 1981 and was completed in July 1983. It consisted of two shafts, underground drifts and crosscuts necessary to support *in situ* experimentation, an exploratory drift to the south and a four-room test panel to the north. The twofold purpose of the SPDV program was, first, to confirm that the subsurface geology was consistent with prior expectations—based on the interpretation of surface investigations. Second, the SPDV program provided for the initial *in situ* confirmations of the underground facility design, design criteria and design bases which would permit the construction of the full facility on a safe, environmentally acceptable, timely and cost-effective basis.

The two shafts consisted of a 10-foot-diameter exploratory shaft, outfitted with a hoist, guides and underground loading facilities for future muck removal, and a six-foot-diameter ventilation shaft. These shafts were constructed by blind drilling. The exploratory drift was excavated to the south, using a continuous mining machine to the southernmost extent of the planned waste storage area, so as to confirm the geology of the salt beds. The four-room test panel was excavated to the same size and configuration as the planned waste storage rooms in the northern portion of the facility. Instrumentation was also installed to monitor the behavior to the underground openings.

FULL FACILITY CONSTRUCTION

The major effort at the start of FFC involved construction of the waste shaft, the exhaust shaft and the nonradioactive experimental areas.

The waste shaft was constructed by enlarging the existing 6-foot-diameter ventilation shaft to 21-foot diameter, lining it with concrete from the surface to the top of the salt bed. This shaft was constructed by drilling and blasting methods, allowing muck to fall to the bottom through the existing 6-foot-diameter shaft. The muck was then moved to the existing outfitted shaft, where it was hoisted to the surface and hauled to the salt storage pile.



Figure 1. WIPP surface facilities.

The exhaust shaft was an entirely new shaft, constructed by drilling a pilot hole from the surface to the facility depth. Then, using the pilot hole, the shaft was upreamed to a six-foot diameter. In the same manner as the waste shaft, the exhaust shaft was drilled and blasted to a 15-foot diameter, and concrete lined to a 14-foot diameter from the surface to the top of the salt formation.

As part of FFC, 31 miles of water pipeline were run to the site, and two water tanks of 160,000 gallons each were constructed. In addition, 14 miles of new highway and eight miles of new railroad were constructed for transportation of waste to WIPP once the WIPP becomes operational. Nine miles of 69-kV power line were run to the site, and two warehouse buildings and numerous temporary offices were constructed.

In the underground, mining and outfitting of a series of experiments in the north end of the underground facility was undertaken, under the cognizance of Sandia National Laboratory (the Scientific Advisor for the WIPP Project). These experiments primarily deal with the properties of salt and its response to (1) mined openings (under both ambient and heated conditions); and (2) experiments relating to nuclear waste emplacement, including materials interactions and brine migration.

At the present time, all of the presently authorized experimental areas have been excavated and the majority of the experiments are in place and producing data. None of these experiments use actual radioactive material; instead, electrical heaters and other simulations are used. WIPP plans to conduct authorized experiments with ac-

tual radioactive material after the facility becomes operational in October 1988.

After completion of the exhaust shaft, work began on the exhaust filter building. A large 10-foot-diameter duct leads from the exhaust shaft to a plenum on the east side of the building. Three main mine fans are positioned outside the plenum; these fans exhaust air from the underground through the duct work and into an exhaust stack. The building contains two trains of High Efficiency Particulate Air (HEPA) filters, which are only used if radioactive material is sensed by instrumentation in the exhaust air. This is accomplished by closing a large diverter valve in the 10-foot-diameter duct, which diverts air through a second duct to the west side of the building, passing it through the HEPA filter trains before exhausting it to the atmosphere through the exhaust stack. Construction of this facility was recently completed.

After completion of the waste shaft, construction began on the support building and the waste handling building. The waste handling building is constructed around the waste shaft, with the 45-ton capacity waste hoist between the two major parts of the building. The largest part of the building is used for CH waste, and contains approximately 25,000 square feet of floor space. Three air locks provide entry of TRUPACT waste shipping containers, while maintaining the building at a slight negative pressure to prevent any possible radioactive contamination spread. All exhaust air from the building is HEPA filtered, to prevent possible radioactive contamination spread. The interior of the building is a large, open, ware-

house-type space where the shipping containers are unloaded, the waste palletized and moved onto the waste hoist from the CH waste side of the waste handling building.

The smaller part of the building is used for RH waste. It has a 140-ton bridge crane for offloading heavy shielded shipping containers. It also contains the hot cell with 5-foot-thick walls, shield windows, manipulators and special handling equipment. After unloading the waste from the shipping container into the hot cell, personnel perform inspections and complete checks before the waste is loaded into a shielded facility cask. This cask is then loaded onto the waste hoist from the RH waste side of the facility. The waste handling building has completed construction, and has been turned over to the Department of Energy for testing, checkout and training of waste handling personnel.

The support building was constructed at the same time as the waste handling building and immediately adjacent to it. The support building provides administrative and technical office space, laboratories, computer facilities and locker and change facilities for workers. The support building was completed and turned over to the Department of Energy in June 1986. The status and arrangement of the surface facilities are shown in Figure 1. In the un-

derground excavation progresses with a continuous miner operating on the necessary drifts and crosscuts to support the final facility underground ventilation routing, for hauling of muck from continuing waste storage area construction and for transport of waste to the waste storage areas. Underground shops, maintenance areas, fuel stations and decontamination stations have also been constructed and outfitted. The mining operation has proceeded into the waste storage area.

The WIPP will store the TRU waste in waste storage panels, each including seven waste storage rooms. This underground layout is shown in Figure 2. Each storage room, and the entries leading to them, is 13 feet high by 33 feet wide. The RH TRU waste will be stored in horizontal holes in the walls of the rooms and CH TRU waste will be stacked within the room.

The waste storage panels are mined, as needed, for waste storage. A panel will be completely mined out before the start of waste storage in that panel, and mining will be in progress for the next waste storage panel while waste is being stored in the panel most recently mined. This permits separation of the waste storage and mining operations, and also minimizes the amount of time that a waste storage room is open before waste is placed in it, so that creep closure of the rooms is minimized. In addition,

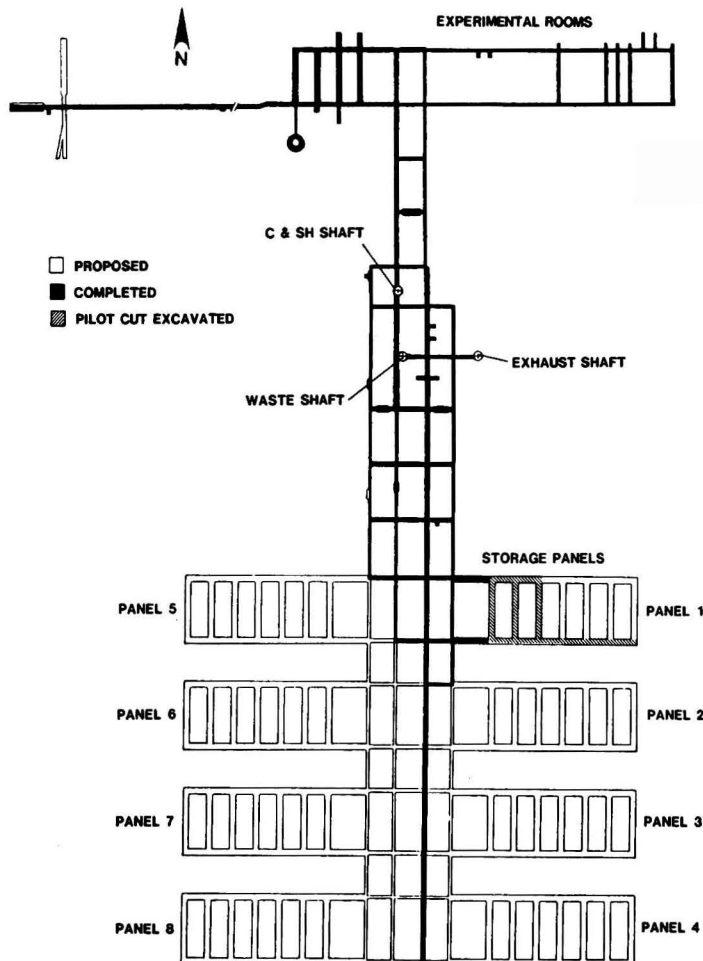


Figure 2. WIPP underground arrangement.

the mining of the waste panel is performed in two phases. The first phase makes a pilot cut 12 feet high by 14 feet wide. The second phase mines the room out to the final 13 feet high by 33 feet wide configuration. The extent of mining, to date, is also shown in Figure 2. A total of approximately 10 linear miles of drift and the three 2150-foot shafts have been excavated, for a total of approximately 600,000 tons. The completion of mining of the first waste storage panel is expected in mid-1987.

SUMMARY

Construction of the majority of the WIPP facilities and systems has been completed, and these facilities and systems turned over for operation. At this time, extensive testing, checkout, training and mock waste handling oper-

ations are in progress. WIPP is scheduled to start to receive actual radioactive waste in October 1988.



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Conversion of Natural Gas to Methanol in Southern Chile

J. M. Rovner and L. J. Opalka

The M. W. Kellogg Company Houston, Texas

A scheme to utilize natural gas reserves in southern Chile to produce 2500 STPD of methanol.

PROJECT STRUCTURE/FINANCING

The Cape Horn Methanol (CHM) plant in southernmost Chile along the Straits of Magellan represents a joint effort of various diverse groups in a project structure which is defined as "project financing". This private investment involves U.S., Chilean and Japanese investors as well as the World Bank's International Finance Corporation (IFC). The attraction to ventures of this type is aggressively priced natural gas provided by Chile's state owned oil company Empresa Nacional Del Petroleo (ENAP). The project organizers have proceeded with the philosophy that they will be one of the low cost producers of a commodity product and therefore assured a place in the methanol market. The project also enjoys private investor independence in that supplemental gas payments will be made to ENAP on the basis of project profitability. Take or pay provisions assures ENAP the debt servicing of a 180 km pipeline financed separately by the Inter-American Bank. Additionally, gas prices are keyed to the wholesale producer price index of the U.S.

Sources of funds made available to the project include syndicated loans by Citibank and three Japanese companies (Nissho Iwai, Marubeni and Kawasaki Heavy Industries), IFC senior and subordinated loans, and equity from joint venture partners. The main partners in this project are The Henley Group, Chilean Investors and the IFC.

This financing package is also enhanced by material pricing from Japan through Kawasaki Heavy Industries (KHI) at a time when the world economy for equipment and bulk materials is very competitive. The combination of all these elements helps overcome the sizable transportation cost in a product export market with consumers in the U.S., Japan and Europe.

Kellogg's involvement in the project began in 1984 with sufficient engineering design to support a defined equipment estimate. During this phase, it was recognized that the technical development and commercial pricing of the project would need to be demonstrated at "arms length". With this concept as a basic definition, M. W. Kellogg's support of the project development was key in reviews by independent consultants to verify that the project scope was thoroughly organized and all technical elements of the project were properly considered. Formal technical and commercial proposals were presented to the Venture and became the basis of awarded contracts.

Engineering and procurement operations are centered at M. W. Kellogg's offices in Houston with close ties to KHI offices in Kobe, Japan for material purchases. Construction will be handled by Chilean subcontractors with supervision by Kellogg. With a turnkey contract award of December 30, 1985, procurement of all major equipment was completed in November 1986. Delivery to the project site began in September 1986 and will carry through the end of 1987. The construction schedule calls for mechanical completion by June 1988. Precommissioning and startup operations will then be completed during the second half of 1988. The overall significant consideration to the turnkey contract is that the contractor has an unrestrained release to build the project with all project documents being executed prior to release to proceed. Sources and utilization of funds were thoroughly analyzed and forecasted to allow the most efficient project execution.

NATURAL GAS GATHERING/CHARACTERIZATION

The natural gas feedstock for this plant originates from the Straits of Magellan reserves of the Empresa Nacional Del Petroleo (ENAP), the Chilean state-owned oil company. The reserve is located at the southern end of Chile near the eastern entrance to the Straits of Magellan, with an estimated capacity of 50 billion standard cubic meters.

The associated gas is initially separated from crude oil in eight local on and offshore production fields which operate at varying pressures from 0.56 to 8.4 MPa. The gas is collected by a pipeline system for transport to a gas treatment plant. Six intermediate compressor stations boost the pressure to 10.5 MPa before drying over molecular sieve/silica gel units. The dried gas is pipelined to a cryogenic gas treatment plant which condenses out the natural gas liquids. The condensates are de-ethanized and further fractionated into propane, butane and natural gasoline. The residual gas is re-compressed to 10 MPa for well re-injection and the user pipeline which feeds the methanol plant. Typical natural gas compositions are shown in Table I. The sulfur content of the treated gas is relatively low. However, the presence of organic sulfur compounds requires hydro-desulfurization prior to entering the process unit. The raw natural gas is relatively light, and can be fed directly to the methanol plant even when the gas treatment plant is out of service.

TABLE I. TYPICAL NATURAL GAS COMPOSITIONS (v%)

Feed to Gas Treatment Plant (From Fields)		Residual Gas From Gas Treatment Plant (To Users)
Methane	90.84	Difference
Ethane	4.7	3.0-5.3
Propane	1.65	0.12-0.25
i-Butane	0.39	0-0.03
n-Butane	0.55	0.0-0.5
i+n-Pentane	0.33	0-0.03
Hexanes	0.14	
Heptanes	0.11	
Octanes +	0.08	
Nitrogen	1.06	0.9-1.35
CO ₂	1.15	0.12-0.30
	100.00	
Water Content = 150 ppm(v)		
Sulfur (H ₂ S + Organics) = 20-25 ppm(v)		

PROJECT ENGINEERING CONSIDERATIONS

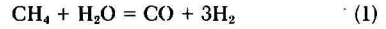
The plant is sited at Cabo Negro, near the 53rd southern parallel, about 20 kilometers north of the city of Punta Arenas. Punta Arenas is a well developed city of about 100,000 with modern telecommunication and sea/air transport facilities. Most of the heavy equipment will be imported through the port of Punta Arenas and trucked to the jobsite. The roads and bridges to the jobsite have been determined to be adequate to handle the expected loads without further reinforcement or rehabilitation. In spite of the extreme southern hemisphere location, the range of weather conditions is not particularly severe. The summer average maximum temperature is 300°K (81°F), with a winter minimum temperature of 260°K (9°F). Therefore, although the average winter temperature is 275°K (36°F), some winterization of the plant will be necessary. Annual snowfall is about 38 cm (15 in.).

A distinguishing feature of the project climatology is the wind. High sustained daytime winds of up to 30 knots can be expected during the spring and summer (September through December). The prevailing wind direction is west-southwest. The plant is situated such that this wind will blow from the inland out to the sea, thereby avoiding engineering problems due to saline wind conditions. Apart from wind loading criteria for tower thickness, this also has constructibility implications for tall equipment. The product from this plant is primarily intended for export, so access to world shipping lanes is important. The Straits themselves pose no significant navigational hindrance. The minimum channel depth is 50 meters so that vessels up to 200,000 DWT can be accommodated.

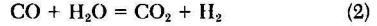
METHANOL PROCESS

The production of methanol from natural gas is relatively straightforward, as indicated schematically in Figure 1. Although the raw natural gas has a relatively low

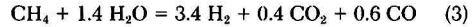
sulfur content, it still requires pre-treatment to protect downstream reforming and methanol catalysts; a fixed bed hydrodesulfurization system will suffice. Natural gas is reformed with steam at a high (1100°K) temperature over a nickel catalyst according to the reaction:



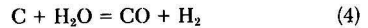
The water gas shift reaction is also catalyzed to near-equilibrium over this catalyst:



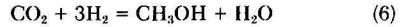
Therefore, the reformed gas contains a significant amount of CO₂ from reaction (2). In practice, the overall stoichiometry in the reformer can be approximately represented by



This equation indicates that 1.4 moles of steam are consumed for every mole of methane reformed. However, much more steam is provided to the reformer to inhibit carbon deposition on the catalyst by the reaction:



After cooling and compression, methanol can be produced directly from the synthesis gas components in equation (3) by the overall reaction system:



Balancing equations (3), (5) and (6) yields the overall reaction:



The process utilizes a conventional copper-zinc methanol synthesis catalyst provided by BASF which is active in the range 5-10 MPa. The catalyst is contained in adiabatic reactors of a spherical geometry. Intercoolers remove the heat of reaction by raising intermediate pressure motive steam for a turbine driver.

Excess hydrogen, which is produced in equimolar amounts with methanol according to reaction (7), and inerts, primarily residual methane and nitrogen from the reformed gas, are purged from the synthesis loop to the plant fuel system. Although the synthesis catalyst has a high specificity to methanol, small amounts of organic by-products, together with the water of reaction from equation (6), must be separated from the final product. This is accomplished in a conventional distillation column sequence. This section is designed to produce chemical quality methanol, since this product generally has more stringent purity specifications than anticipated requirements for fuel grade methanol.

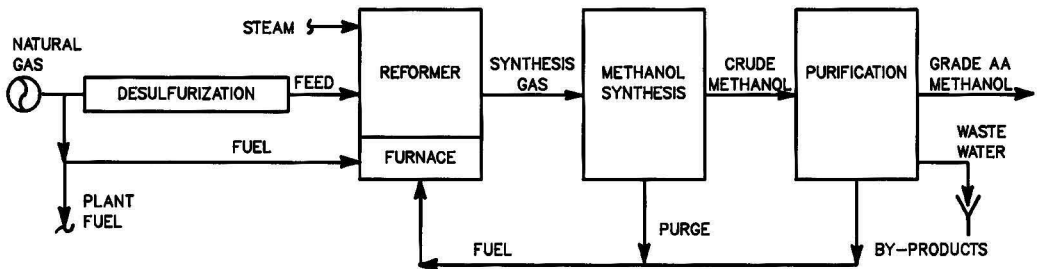


Figure 1. Methanol process conceptual block flow diagram.

PROCESS HEAT INTEGRATION/STEAM BALANCE

The overall stoichiometry of the methanol process given by equation (7) gives rise to some rather unique considerations in plotting an energy efficiency strategy for the plant. The excess hydrogen adds to the volume of reformed gas that must be cooled before compression to synthesis loop pressure. Process heat requirements, primarily for the purification section reboilers, is relatively minimal and at low levels. This leaves most of the heat in the reformed gas train to be recovered in the high pressure motive steam system.

Even at high synthesis loop carbon conversions, the hydrogen rich purge will very nearly satisfy the entire firing requirement of the reformer, which is by far the largest fuel user in the plant. It is possible for the methanol process to be both a net fuel and steam exporter, especially as designs become more efficient. This is no problem in an integrated plant where credit can be claimed for exporting this energy to other process or utility units. The challenge in a grassroots plant such as this was to integrate the process with offsites in the most efficient way possible without compromising the operability and reliability of the plant. Additionally, in this location, there is no existing local infrastructure to which co-generated energy, for example, in the form of electricity or steam, could be exported.

The overall plant steam balance is shown in Figure 2. The entire high pressure steam requirement is raised in a process waste heat boiler. This steam is superheated in the convection section of the reformer and used to power the make-up gas compressor turbine. Steam is extracted from this turbine to provide reforming steam and power smaller process drivers from a medium pressure header. Although all the steam required by the plant could have theoretically been generated within the methanol process battery limits by including an auxiliary boiler on the reformer, an additional source of steam would still be required for start-up. This is normally generated by an offsites package boiler. In this plant, the package boiler is

sized for this requirement, but is kept continuously on-line to control the medium pressure steam header. Intermediate pressure steam raised in the reactor intercoolers drives a steam turbo-generator which provides the bulk of the electrical load for the plant. Turbines using medium and intermediate pressure steam exhaust to a low pressure header which primarily provides latent heat to the desalination plant and a distillation section reboiler. A small amount of low pressure steam will also be used in tracing for freeze protection.

ENGINEERING DESIGN

Due to the grassroots nature of the project, only natural gas and seawater are imported across the plant battery limits. All other utilities for start-up and shut-down as well as normal operation must be provided internally. A once-through seawater system provides the bulk of cooling duties in the plant. A small indirect fresh water closed loop cooling system is used in services where a salt water leak into the process side would be intolerable. Air coolers are used for larger duties where the cost of including this service in the fresh water cooling loop would be prohibitive. Seawater is also directly used for firewater. Seawater must also, of course, be the starting point for all fresh water in the plant. An evaporator-type desalination plant using low pressure steam produces the desalinated water. Further treatment makes this water suitable for potable uses and boiler feedwater make-up.

Due to the remote location of the plant, a fairly large storage capacity was included. Large fixed roof tanks were deemed the most economical for this requirement, which require inert gas blanketing to prevent explosive mixtures from forming particularly during ship loading. This is by far the largest use of nitrogen, with smaller amounts made available for process purging. Nitrogen is produced by a cryogenic separation unit with a motor driven air compressor. A Dewar is included to provide a comfortable reservoir of liquid nitrogen for emergency situations.

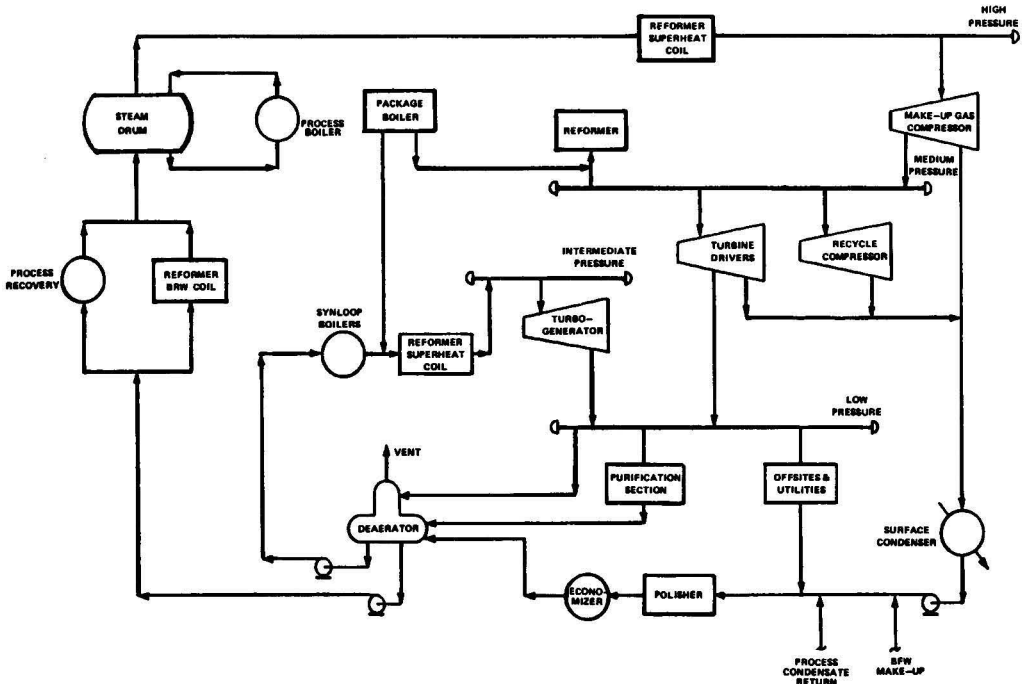


Figure 2. Conceptual steam balance.

In order to enhance the reliability and operability of this plant, this design includes a gas turbine generator which shares the electrical load and has the capacity to keep the plant running should the steam driven turbo-generator shutdown. The gas turbine generator is also particularly useful in providing electrical power for the initial "black start", so that critical utilities can be motor operated until the main steam system is available. Similarly, the steam driven turbo generator is designed to support full plant operation when the gas turbine generator is out of service. In this case, the package boiler provides the additional steam required by the increased load on this machine. A diesel powered generator is also provided to maintain essential services (such as cooling, instrument air, motor operated valves, etc.) for a safe and orderly emergency shutdown of the entire process unit.

The integration of process and offsites utilities is indicated in Figure 3. For a grassroots plant, the classical distinction between these classes of energy consumers becomes less meaningful. The energy required to make methanol truly is the heating value of the natural gas crossing the plant battery limits and will be in the range 8-8.9 MM KCAL/MT.

CONCLUSIONS

Conversion to methanol can offer an effective way to utilize natural gas as an easily transportable bulk commodity. This is logistically advantageous for remote locations such as southern Chile. Methanol also has market growth potential for international trade, making conversion projects of this type attractive to developing countries with little internal demand for natural gas or its derivatives. The investment opportunity for this specific project has suited the objectives of both the World Bank and private investors. More importantly, natural gas con-

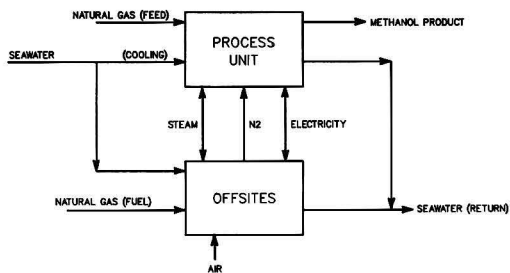


Figure 3. Process/utilities integration.

servation is encouraged by providing an economic alternative to non-exploitive uses (flaring, in the extreme) of this non-renewable resource.



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L. J. Opalka is a Project Manager for The M. W. Kellogg Company with overall project responsibilities including contract, procurement and engineering design. He holds a B.S. degree in Civil Engineering from the University of Illinois and is a registered Professional Engineer in Illinois and Texas.

Cogenerate With Coke

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Midrange petroleum fractions can be produced from heavy crudes by hydrogen addition or by carbon removal. The latter has increased the supply of fuel grade, high sulfur coke. This, combined with energy regulations, provides incentives for coke-fired cogeneration plants.

INTRODUCTION

As refiners turn to heavier, sourer feedstocks, the production of fuel grade green coke from delayed cokers is increasing. The market for fuel grade coke in the United States is not strong, most of the fuel grade coke produced is exported. These market conditions create an environment where a cheap source of fuel is available for use in cogeneration facilities producing electricity and process steam for use in the coke producing refinery. The refinery steam and electricity requirements can be satisfied by the cogeneration plant and excess electricity can be sold to the local electric utility at avoided cost rates. The refinery gains by having an economical source of steam and electric power. Possibly of more importance, the refiner has a reliable means of disposing of petroleum coke with an economic return based upon the value of the steam and power produced in the cogeneration plant rather than the uncertain or low value that can be received in the fuel grade coke market.

This article examines the technical and economic feasibility of a coke based cogeneration project for a U.S. Gulf Coast refinery. Cases are developed comparing coke based cogeneration with import of electricity and natural gas fuel to produce steam. Two coke burning technologies are examined: a circulating fluid bed boiler and a pulverized coke fired boiler. The refinery is assumed to require 145 MW of electricity, 400,000 lb/hr of 600 psig steam and 1,200,000 lb/hr of 250 psig steam. Furthermore, it is assumed that the refinery produces 2800 T/D of coke that must be disposed of in the cogeneration plant.

TECHNOLOGY FOR BURNING COKE

Two technologies for burning petroleum coke to produce steam are considered. Both technologies are based upon adaptation to coke firing of well known commercial plant experience. They represent Combustion Engineering's continued efforts in developing boiler applications for fuels other than traditional gas, oil and coal.

Circulating Fluid Bed Boiler

The Lurgi/Combustion Engineering Circulating Fluid Bed (CFB) Boiler is a well developed commercial technology for producing steam. The CFB boiler has been de-

veloped to efficiently burn solid fuels previously considered environmentally unacceptable. These fuels include coal and lignite residue, anthracite culm, sludges and of interest here, petroleum coke.

The design of the CFB boiler avoids the need for installation of costly scrubbing facilities normally required for controlling sulfur oxide emissions. The inherently low combustion temperature in the CFB boiler has the advantage of low nitrogen oxide formation. The process assures stack emissions that meet stringent environmental regulations. Solid waste is in the form of a mixture of dry ash, calcium sulfate and calcined lime.

Combustion in a CFB boiler takes place in a vertical combustion chamber. Fuel and limestone are fed into the combustion chamber, fluidized, and burned at temperatures of 1500-1600°F. Limestone reacts with the sulfur dioxide released by burning the fuel to form calcium sulfate. The bed material in the combustion chamber consists primarily of calcium sulfate, excess calcined lime and ash from the fuel. Limestone is added at a conservative 1.6 Ca:S ratio to ensure greater than 90% removal of sulfur compounds.

The bed material is fluidized with primary air introduced through a grate at the bottom of the combustion chamber and by the combustion gases generated. Secondary air is added to the combustion chamber to achieve complete and staged combustion.

The velocities in the combustion zone of the CFB are set so that the combustion gases entrain a considerable amount of solids from the combustion chamber. The solids are separated from the gases in a cyclone and are returned to the bed directly or after passing through an external fluid bed heat exchanger (FBHE). High internal and external circulating solids rates characteristic of CFB's, result in uniform temperatures throughout the combustion chamber and effective temperature control, facilitating good SO₂ control.

Long solids residence and contact times, with the high heat and mass transfer rates achieved, result in a high combustion efficiency. These conditions also allow both the complete decomposition of the limestone and subsequent capture of SO₂ at a low calcium to sulfur molar ratio. Staged combustion, at controlled low temperatures, effectively controls NO_x formation.

Heat for steam generation and superheat is removed from the combustion chamber walls, the FBHE and a convective pass of heat exchangers including an economizer where heat is removed from the flue gas exiting the

solids recycle cyclone. After passing the convective heat exchangers the gases are further cooled in an air preheater. The flue gases are cleaned in a bag house to meet New Source Performance Standards for particulate emissions of 0.03 lb/10⁶ B fired and then are vented via an induced draft fan and stack. Although higher efficiencies are expected, a conservative boiler efficiency of 88% (LHV) was assumed for this study.

Excess solid waste is removed from the system via a fluid bed cooler. Materials are removed from the fluidized bed in order to control solids inventory and remove sulfur oxides captured in the form of calcium sulfate. Material removed is taken offsite for disposal.

Pulverized Coke Boiler

The Combustion Engineering Pulverized Coke Boiler (PCB) is a design adapted from Combustion Engineering's long experience with conventional pulverized coal burning boilers. Changes are made from standard coal burning boilers to handle the low ash, high metals and low volatiles content of the coke.

Petroleum coke is pulverized and burned with primary air in a modified tangentially fired boiler. Secondary air and recycled flue gas are added to the combustion radiant chamber to help minimize NO_x formation. Modifications to standard coal boiler design include larger fire box volume to increase the residence time which is required for complete burn out of the low volatiles containing coke, and refractory lining portions of the radiant fire box to protect the tubes from high corrosion due to the high sulfur and vanadium content of the coke. Supplementary firing with natural gas at a rate of 5% of the fuel value of coke is included to ensure complete combustion.

Ash and combustion gases pass overhead into convective heat exchangers for additional steam generation, superheating and economizer. The gases are passed through an air preheater. Ash is then removed in a venturi scrubber before the flue gases are scrubbed for sulfur oxides removal in an Air Quality Control System (AQCS) followed by a wet precipitator for acid mist removal. Limestone is added to the AQCS for sulfur oxides removal. The waste stream from the AQCS is a gypsum sludge containing fly ash at approximately 30% water content. Overall thermal efficiency for the PC Boiler was taken as 88%, similar to the CFB.

Coke Quality

Coke fuel for both of the boiler cases is based upon the following typical fuel grade coke characteristics:

Volatile content	12.5%
Sulfur	7%
Moisture	10%
Ash	0.2%
Heating Value BTU/lb (HHV)	15,000
Hardgrove grindability	70

These characteristics are similar to that of an anthracite coal except ash is much lower and Hardgrove grindability is higher (i.e., coke is more friable).

DESCRIPTION OF POWER CYCLES

High pressure superheated steam from the coke fueled boilers is fed to turbogenerators for production of electricity and steam. Steam is extracted at nominally 610 psig to supply 400,000 lb/hr of 600 psig steam to the refinery. Steam at nominally 260 psig is extracted to supply 1,200,000 lb/hr of 250 psig steam to the refinery and to

satisfy boiler feed water preheating requirements. Extraction steam at nominally 50 psig is used to strip dissolved gases from the BFW in the deaerator. Flash steam from boiler blowdown is also used by the deaerator. Steam from the exhaust of the turbine is condensed at 3.5 inches mercury absolute against cooling water. A diagram of the overall power cycle is shown in Figure 1.

It is assumed that 50% of the steam exported to the refinery will be returned as condensate. The condensate and makeup water are treated to remove dissolved solids and then fed to the deaerator. In the deaerator the water is stripped of dissolved gases. The water is then pumped, pre-heated against 260 psig steam and fed to the boiler.

Electricity is generated by the turbogenerators at 13.8 KV. A certain amount of electricity is used in the cogeneration plant for water treatment, cooling tower, materials handling and waste treatment. This power is fed to transformers within the cogeneration plant and used internally. Excess power is exported to the refinery and the utility at 13.8 KV. In this analysis the refinery requires 145 MW of power.

Table 1 summarizes the operating parameters for the Circulating Fluid Bed Boiler and the Pulverized Coke Boiler cases.

The two power cycles are similar in most respects. For the Pulverized Coke Boiler, supplemental firing with natural gas is included to ensure complete burning of the coke fuel. Steam temperature is reduced to 950°F compared to 1000°F for the Circulating Fluid Bed Boiler because of concerns relating to corrosion of superheater surfaces due to the metals content of coke. Overall cycle efficiency is 71.8% for the CFB and 68.3% for the PCB case.

The CFB boiler plant is based on four boilers rated at 750,000 pounds per hour (MCR) of 1810 psig/1000° steam. The boilers are sized so that the plant can meet the refinery operating requirements for steam and electricity with three boilers operating and one boiler down for inspection or maintenance. The boilers produce steam which is fed to two extracting condensing turbogenerators. Each turbogenerator is capable of generating 125 MW of power. Normally each turbogenerator produces 91.5 MW of electricity. The additional capacity is provided for flexibility when the refinery does not require the design export steam quantity. When the refinery requires less steam, the steam will be condensed. The additional power produced will be sold to the electric utility.

The PCB plant consists of three boilers each capable of producing 1,125,000 pounds per hour (MCR) 1810 psig/950°F steam. The boilers are sized so that the plant

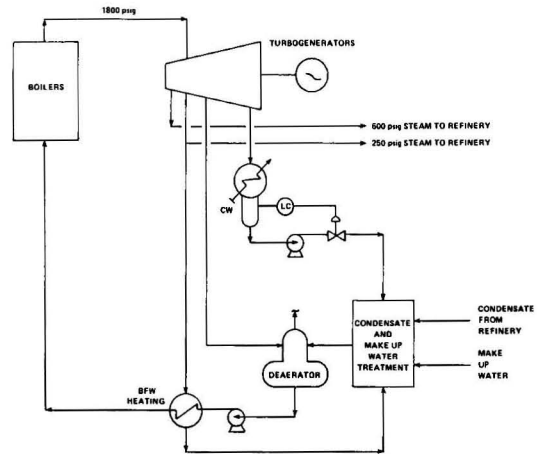


Figure 1.

TABLE 1. SUMMARY OF OPERATING PARAMETERS

	Circulating Fluid Bed	Pulverized Coke
Fuel Fired:		
Coke	2800 T/D	2800 T/D
Natural Gas	0	4200 × 10 ⁶ BTU/D
Utilities & Other Materials:		
Limestone	972 T/D	680 T/D
Water	3400 GPM	4200 GPM
Waste:		
Ash	1188 T/D	1536 T/D (30% water)
Steam Produced:		
Quantity	2,706,000 lb/hr	2,926,000 lb/hr
Temperature	1000 °F ₆	950 °F
Pressure	1810 PSIG	1810 PSIG
Power Generated:		
Gross	183 MW	200 MW
Internal Usage	31 MW	36 MW
Available for Export	152 MW	164 MW
Export to Refinery	145 MW	145 MW
Export to Utility	7 MW	19 MW
Steam Export to Refinery:		
610 psig	400,000 lb/hr/740°F	400,000 lb/hr/690°F
260 psig	1,200,000 lb/hr/590°F	1,200,000 lb/hr/530°F
Condensate Return	800,000 lb/hr	800,000 lb/hr
Overall Efficiency for Power Cycle	71.8%	68.3%

can meet the refinery operating requirements for steam and electricity with two boilers operating and one boiler down for inspection or maintenance. The boilers produce steam which is fed to two extracting condensing turbogenerators. Each turbogenerator is capable of generating 135 MW of electricity. Normally each turbogenerator produces 100 MW of power. As for the CFB case, this additional capacity is provided for operating flexibility when the refinery does not require the design quantity of steam.

The configurations of the two facilities have been designed with operating flexibility and overall reliability in mind. Operating flexibility is enhanced by the capability to operate easily at different steam export and power production rates than the design scenarios.

Reliability is provided for by building spare boiler capacity and sparing critical equipment to ensure the plants can meet the refinery requirements for steam and power even when one boiler is not operating, or when critical equipment fails or requires maintenance.

INVESTMENT

Investment cost for the CFB and PCB plants was arrived at using equipment factored estimates. This type of estimate produces a capital cost which is considered accurate to ±20% of eventual cost of the installed plant.

To prepare the equipment factored estimate the two plants were defined by performing process studies which set the philosophy and major equipment sizes for items such as the boilers, turbogenerators, cooling tower, fuel handling and storage. Major equipment items were estimated using vendor quotations. An equipment list for each plant was developed and the remaining equipment was estimated using in-house data from other projects for similar equipment. These data are also used to check vendor quotes on major items. The equipment was broken into five different areas to facilitate the estimate. The areas were offsites, steam and power, utilities, environmental and buildings.

Commodity costs associated with equipment include civil, structural steel, piping, instrumentation, electrical, fireproofing, painting and insulation. These costs are esti-

mated using cost ratio factors, developed by analyzing return cost data for projects of a similar nature. Buildings were estimated on a square foot basis.

Certain utility systems and offsite costs are estimated on a system basis (i.e., water treatment and waste handling). Take-offs for interconnecting piping were made based upon a conceptual plot plan developed and are used to develop quantities and costs. Field labor costs for installation work were determined by pro-rating from corresponding equipment and materials costs for similar projects. These costs reflect labor wages and productivity anticipated for the Gulf Coast area.

Construction and miscellaneous field expenses including construction equipment, tools and temporary facilities are estimated as a percent of direct field labor costs. Field staff and office costs are similarly figured. Historical information is again used in determining these cost items.

Engineering costs are developed from manhour estimates from each discipline. The manhours were then priced at current rates and non-payroll home office costs added.

All of the above mentioned expenses were summed up to arrive at an estimated total installed cost (December 31, 1985) for each plant. Once the estimated TIC was established, escalation, contingency and financing expenses were added to arrive at an expected investment for the plant.

The results of the analysis are presented in Table 2. Expected investment for the CFB boiler case is \$322,000,000; for the PCB case \$416,000,000.

TABLE 2. INVESTMENT FOR COGENERATION PLANTS (\$X10⁶)

	CFB Boiler	PCB Boiler
Plant Cost	240	310
Escalation	30	40
Contingency	25	30
Subtotal	295	380
Interest During Construction	27	36
Total Capital Investment	322	416

ECONOMICS

Analysis

An econometric model was developed to evaluate the economic performance of the Circulating Fluid Bed Boiler and the Pulverized Coke Boiler cogeneration plants. The proposed facilities were each compared against present day steam and electricity costs taking into account appropriate escalation for inflation. A three year construction period and a 10 year project life is considered. Inflation for the 13 year time frame was assumed at 3.5%/year.

The model was developed taking into account capital costs, operating costs, expected revenues, loan financing and taxes.

While obviously both the CFB and the PCB cases were calculated using the model, only the final results for both cases are presented. Also shown are some of the significant intermediate calculation tables for the CFB case only.

Capital costs include the cost of the plant, interest during construction and cost of site preparation. Working capital is based on interest required to cover two month's revenues. Plant capital costs are used as a basis for depreciation calculations. Sources and uses of funds during construction for the CFB case are shown on Table 3.

Operating costs include the cost of coke, natural gas, operations and maintenance, limestone and waste disposal. A sensitivity analysis was performed varying the unit price of coke between \$1 and \$20 per ton. Limestone and natural gas were priced at \$15/ton and \$2/10⁶ BTU, respectively. Waste disposal costs for ash and sludge were estimated at \$13/ton for the Circulating Fluid Bed boiler dry ash and \$15/ton for the Pulverized Coke boiler ash/sludge mixture (30% water). Operation and maintenance are taken as 4% of the plant capital cost and include insurance, labor, administration, supplies and spare parts. These costs are in 1986 dollars. The model adds inflation to each cost for each year of project life to arrive at current values for all cost parameters. The model then calculates yearly values for total operating costs.

Revenues are similarly based on present values for Gulf Coast electricity and steam costs. Refinery electric energy costs in 1986 are taken at \$0.05/Kwh. Refinery

steam costs are \$3.75/1000 lb for 600 psig steam and \$3.00/1000 lb for 250 psig steam. Cogenerated electricity bought by the public utility is estimated to return \$0.04/Kwh. Each product is considered separately. The inflation rate of 3.5% per year is taken into account to arrive at annual revenues.

The model is designed so that both operating costs and revenues can be adjusted to reflect plant operating rates above or below the design output. Since sufficient spare capacity was included for each plant studied, operating rate was not varied over the life of the project in this analysis. The economic returns are not particularly sensitive to operating rate unless operating rates are significantly below design. Operating costs and revenues (1986) are presented in Table 4 for varying coke costs between \$1 and \$20 per ton.

The analysis to calculate return on investment is based upon the following parameters. Equity to debt ratio is 25/75% based upon total capital cost. Total capital cost includes plant cost, site preparation and interest during construction. Debt is financed by a 10 year bank loan at 10% interest rate with equal yearly payments to be paid over the first 10 years of the project operating life. Depreciation is calculated using the 10 year accelerated cost recovery system (ACRS). Taxes are assessed at 46% of gross profits. The model calculates the yearly payments required to satisfy the loan requirements allocating the proper amounts for interest and principal.

On a yearly basis, gross profit is figured by adding total revenues and subtracting total operating costs, interest paid on the bank loan and depreciation. Taxes are then subtracted to calculate net profit. Cash flow is arrived at by adding depreciation to net profit and subtracting the principal payment for the bank loan. The internal rate of return (IRR) on owner's equity is then calculated from the cash flow for a ten year operating period.

Results of Economic Analysis

The econometric model was used to analyze various cases for the CFB and PCB configurations. Studies were performed to see how different parameters and assumptions affect the economic analysis. The cash flow analysis

TABLE 3. SOURCES AND USES OF FUNDS DURING CONSTRUCTION

Coke Price 10\$/T Circulating Fluid Bed Boiler	(000's US\$)							
	Year 1		Year 2		Year 3		Total	PCT of TOT Plant
	1st Half	2nd Half	1st Half	2nd Half	1st Half	2nd Half		
Source of Funds:								
Owner's Equity	\$5,563	\$5,775	\$14,664	\$15,224	\$19,362	\$20,102	\$80,690	25.0%
Bank Loans								
Loan 1	\$16,688	\$17,326	\$43,992	\$45,673	\$58,085	\$60,305	\$242,069	75.0%
Total Bank Loans	\$16,688	\$17,326	\$43,992	\$45,673	\$58,085	\$60,305	\$242,069	100.0%
Total Sources	\$22,251	\$23,102	\$58,655	\$60,897	\$77,447	\$80,407	\$322,759	100.0%
Uses of Funds:								
Plant Construction	\$21,834	\$21,834	\$55,855	\$55,855	\$69,811	\$69,811	\$295,000	91.4%
Construction Interest (M-O-P Drawdown)	\$417	\$1,268	\$2,801	\$5,042	\$7,636	\$10,596	\$27,759	8.6%
10% int rate: Loan 1								
Total Uses	\$22,251	\$23,102	\$58,655	\$60,897	\$77,447	\$80,407	\$322,759	100.00%

TABLE 4. 1986 OPERATING COSTS AND REVENUES
Circulating Fluid Bed Boiler (1986)

Operating Costs					
\$MM					
Coke Cost \$/T	1.00	5.00	10.00	15.00	20.00
Cost of Coke	1.02	5.11	10.22	15.33	20.44
\$MM					
Natural Gas	2.00	2.00	2.00	2.00	2.00
Unit Cost					
Gas Cost	0.00	0.00	0.00	0.00	0.00
\$MM					
Operation/	15.20	15.20	15.20	15.20	15.20
Maint \$MM					
Limestone	15.00	15.00	15.00	15.00	15.00
\$/T					
Limestone	5.32	5.32	5.32	5.32	5.32
Cost \$MM					
Waste	13.00	13.00	13.00	13.00	13.00
Disposal \$/T					
Waste	5.64	5.64	5.64	5.64	5.64
Disposal					
Cost \$MM					
Total Operat-	27.18	31.27	36.38	41.49	46.60
ing Cost					
\$MM					
Revenues \$MM					
C/kWh	5	5	5	5	5
Refinery					
C/kWh Utility	4	4	4	4	4
\$/1000 lb 600	3.75	3.75	3.75	3.75	3.75
\$/1000 lb 250	3.00	3.00	3.00	3.00	3.00
Ref Elec Rev	63.51	63.51	63.51	63.51	63.51
\$MM					
Util Elec Rev	2.45	2.45	2.45	2.45	2.45
\$MM					
600 Rev \$MM	13.14	13.14	13.14	13.14	13.14
250 Rev \$MM	31.54	31.54	31.54	31.54	31.54
Total	100.64	110.64	110.64	110.64	110.64
Revenues					
\$MM					

and the internal rate of return for \$10/ton coke is shown on Table 5.

For the CFB configuration, using \$10/ton coke (1986) projected operating revenues exceed operating costs by 84 million dollars in the first year of operation. This increases to 114 million dollars in the tenth year of operation. Cash flow grows from 19 million dollars in year one to 48 million dollars in year ten. The internal rate of return over this period equals 28.9%.

For the PCB facility, considering \$10/ton coke (1986), projected operating revenues exceed operating costs by 79 million dollars in the first year of operation. This increases to 108 million dollars in the 10th year of operation. Cash flow grows from 9 million dollars in the first year to 41 million dollars in the tenth year. The internal rate of return on owner's equity is 16.6% over a 10 year operating period.

Comparing the return on investments for the two boiler types, the IRR for the CFB boiler is better than that for the PCB case. This result is attributable to the considerable capital savings associated with the CFB compared to the PCB. High capital costs for the PCB are due to the equipment required for sulfur oxide removal and sludge handling. These costs are avoided with the CFB as no "add-on" equipment is required to control sulfur oxide emissions.

A sensitivity analysis was done to quantify how the internal rate of return changes as the value of the coke fuel is varied between \$1/ton and \$20/ton (1986 \$). The results are presented in Table 6 for the CFB and PCB cases. As would be expected, when the cost of coke fuel decreases, IRR increases and vice versa.

Another analysis was performed to see how IRR is affected by an increase or decrease in plant total capital cost. Total capital cost for \$10/ton coke cases was decreased 10% from the base to calculate the low case and increased 10% from the base to form the high case. The results of this analysis are shown on Table 7. For the CFB the 10 year average IRR is 28.9% for the base case, 33.3% for the low capital case and 25.0% for the high capital

TABLE 5. ECONOMIC ANALYSIS—PER YEAR CIRCULATING FLUID BED BOILER

Coke Price 10 \$/T	(000's US\$)						
	Year 1	Year 2	Year 3	Year 4	Year 5	Year 6	Year 7
Sales Revenue							
Value of Products	\$0	\$0	\$0	\$124,810	\$129,178	\$133,700	\$138,379
Value of Feeds & Operating Cost	\$0	\$0	\$0	\$41,040	\$42,476	\$43,963	\$45,502
Margin Ex Working Cap	\$0	\$0	\$0	\$83,770	\$86,702	\$89,737	\$92,877
Interest on Working Cap 2 Months Sales				\$2,080	\$2,153	\$2,228	\$2,306
Net Operating Margin	\$0	\$0	\$0	\$0	\$0	\$0	\$0
Interest on Bank Loans During Operation	\$0	\$0	\$0	\$23,602	\$21,181	\$18,760	\$16,340
Total Inter Payments	\$0	\$0	\$0	\$23,602	\$21,181	\$18,760	\$16,340
Depreciation	\$0	\$0	\$0	\$25,821	\$45,186	\$38,731	\$32,276
Income Before Taxes				\$32,267	\$18,182	\$30,017	\$41,955
Income Taxes				\$14,843	\$8,364	\$13,808	\$19,299
After Tax Income				\$17,424	\$9,818	\$16,209	\$22,656
Depreciation				\$25,821	\$45,186	\$38,731	\$32,276
Principal Payments	\$0	\$0	\$0	\$24,207	\$24,207	\$24,207	\$24,207
Equity Investment	(\$11,338)	(\$29,888)	(\$39,463)				
Net Cash Flow	(\$11,338)	(\$29,888)	(\$39,463)	\$19,038	\$30,797	\$30,733	\$30,725

Internal Rate-of-Return on Owner's Equity: 28.9% P.A.

TABLE 5. (CONTINUED)

Coke Price 10 \$/T	(000's US\$)						
	Year 8	Year 9	Year 10	Year 11	Year 12	Year 13	Total
Sales Revenue							
Value of Products	\$143,222	\$148,235	\$153,423	\$158,793	\$164,351	\$170,103	\$1,464,195
Value of Feeds & Operating Cost	\$47,094	\$48,743	\$50,449	\$52,214	\$54,042	\$55,933	\$481,456
Margin Ex Working Cap	\$96,128	\$99,492	\$102,975	\$106,579	\$110,309	\$114,170	\$982,739
Interest on Working Cap 2 Months Sales	\$2,387	\$2,471	\$2,557	\$2,647	\$2,739	\$2,835	\$24,403
Net Operating Margin	\$0	\$0	\$0	\$0	\$0	\$0	\$0
Interest on Bank Loans During Operation	\$13,919	\$11,498	\$9,078	\$6,657	\$4,236	\$1,816	\$127,086
Total Inter Payments	\$13,919	\$11,498	\$9,078	\$6,657	\$4,236	\$1,816	\$127,086
Depreciation	\$32,276	\$32,276	\$29,048	\$29,048	\$29,048	\$29,048	\$322,759
Income Before Taxes	\$47,546	\$53,248	\$62,292	\$68,227	\$74,285	\$80,471	\$508,490
Income Taxes	\$21,871	\$24,494	\$28,654	\$31,384	\$34,171	\$37,017	\$233,905
After Tax Income	\$25,675	\$28,754	\$33,638	\$36,843	\$40,114	\$43,454	\$274,585
Depreciation	\$32,276	\$32,276	\$29,048	\$29,048	\$29,048	\$29,048	\$322,759
Principal Payments	\$24,207	\$24,207	\$24,207	\$24,207	\$24,207	\$24,207	\$242,069
Equity Investment							(\$80,690)
Net Cash Flow	\$33,744	\$36,823	\$38,479	\$41,684	\$44,956	\$48,296	\$274,585

Internal Rate-of-Return on Owner's Equity: 28.9% P.A.

TABLE 6. SENSITIVITY ANALYSIS OF IRR TO COKE PRICE

Coke Price (\$/st)	CFB	PCB
1	33.9	21.4
5	31.7	19.3
10	28.9	16.6
15	25.9	13.6
20	22.7	10.5

TABLE 7. SENSITIVITY ANALYSIS OF IRR TO INVESTMENT

Capital Cost	CFB	PCB
Low (-10%)	33.3	20.6
Base	28.9	16.6
High (+10%)	25.0	13.0

case; for the PCB: 16.6% for the base case, 20.6% for the low capital case and 13.0% for the high capital case.

CONCLUSIONS

A study is presented looking at the economics of steam and power cogeneration using fuel grade green coke in a U.S. Gulf Coast Refinery. The economics are not embellished by the various tax benefits that are or were available to projects of this type, but rather looks at the basic common economic driving potentials.

The results of the study show that cogeneration presents a viable avenue to create a profitable market for an otherwise undesirable byproduct of a refinery. The study results could be equally valid for any other complex requiring power and steam in an area where fuel grade coke is available from a nearby refinery.

The better economics shown for the CFB indicate this to be the preferred route. This conclusion is further supported by some operating characteristics of the CFB, especially its flexibility and adaptability to different fuels. By switching to different fuels the operator could take advantage of fluctuating relative fuel prices and substitute other fuels (e.g., heavy residue or even gas) for the coke, if a temporary shortage should present attractive sales opportunities for coke.

While this paper presented two viable means of coke utilization, there are other processes that warrant exami-

nation. One of these is combustion in an atmospheric "bubbling" fluid bed (AFB) similar to the one under design for the TVA. This cycle would be similar to the CFB.

Another possibility to utilize the coke is gasification and combustion in a gas turbine based combined cycle. While the investment for such an installation is expected to be higher than that for straight combustion, the additional revenue potential due to improved cycle efficiency may justify the incremental investment.

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Catalytic Control of FCC Regenerator SO_x Emissions

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Chevron's Transcat reacts with sulfur oxides in the regenerator to form a solid compound which is subsequently reduced in the reactor, releasing H_2S . The result is effective, stable sulfur oxide control with no measurable effect on plant capacity, yield, or conversion. The technology is very cost effective relative to alternatives.

INTRODUCTION

Many of the recent advances in fluid catalytic cracking have contributed to reduced emissions. Attrition resistant catalysts and complete combustion regeneration have reduced particulate and CO emissions. Catalysts with improved selectivity and riser cracking have reduced coke make. This in turn reduces combustion-related emissions such as sulfur and nitrogen oxides. In spite of these developments, the fluid catalytic cracker (FCC) remains a recognized source of sulfur oxide (SO_x) emissions in many refineries. The FCC process, as a consequence, has not escaped the attention of regulatory agencies. In 1978, the South Coast Air Quality Management District in California adopted a stepwise limit on FCC SO_x . On January 1, 1987, this limit is scheduled to drop to 60 kg (132 lb) $\text{SO}_x/1000$ bbl feed or about 300 ppmv in the regenerator flue gas. The Environmental Protection Agency is currently formulating a New Source Performance Standard (NSPS) for FCC SO_x . The limit currently being considered is 9.8 kg $\text{SO}_x/1000$ kg coke burnoff, which again is about 300 ppmv in the regenerator flue gas.

Refiners presently have three options for controlling FCC SO_x emissions: hydrofining, flue gas scrubbing and catalyst technology. Hydrofining for SO_x control and flue gas scrubbing are expensive options—ranging upward from about \$0.40/feed bbl in total (capital plus operating) cost. FCC feed hydrofining has many process benefits including FCC yield gains. Where these benefits provide sufficient economic incentive, hydrofining may be the process of choice for controlling FCC SO_x . Scrubbing, on the other hand, provides no yield benefits. With its associated sludge or spent solution disposal requirements, scrubbing will be exceedingly expensive in some locations, particularly on the West Coast.

Conceptually, control of FCC SO_x via catalyst technology has several attractive characteristics:

1. Little or no capital investment is required.
2. Operating cost is low.
3. Reliability is high, equal to that of the FCC itself.
4. Waste disposal problems are minimal.

For these reasons, Chevron has actively pursued the catalyst technology alternative. This paper summarizes the success of this approach.

What is Catalyst Technology?

Use of catalyst technology causes sulfur that would normally appear as regenerator SO_x to appear instead as H_2S in the reactor offgas. The characteristic of an FCC that makes this possible is the circulation of catalyst between oxidizing conditions in the regenerator and reducing conditions in the reactor. Catalyst technology takes advantage of this redox cycle by incorporating in the circulating inventory a material that can form a stable compound with sulfur under oxidizing conditions but not under reducing conditions. From a practical standpoint, material selection is also limited by the need to avoid adverse catalytic effects in the FCC reactor. The material should also possess high physical and chemical stability, and should be nontoxic.

Chevron's proprietary catalyst technology, called Transox, was developed to meet these criteria. Transox operation is illustrated schematically in Figure 1. It involves replacing a small portion of the normal cracking catalyst inventory with special Transox catalyst (Transcat). In the oxidizing atmosphere of the regenerator, Transcat reacts with sulfur oxides to form a solid com-

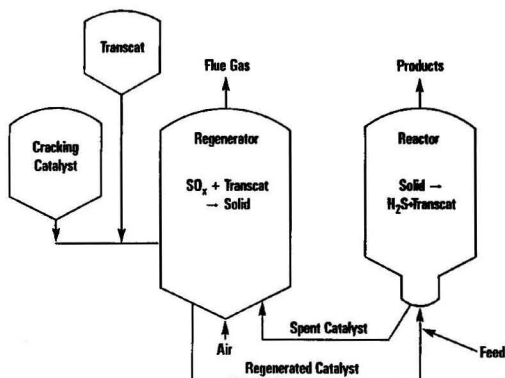


Figure 1. Schematic diagram of transox operation.

pound. In the reactor, the solid compound is reduced, releasing H₂S. The overall effect is that regenerator SO_x emissions are reduced while reactor H₂S make is increased. Typically, the incremental H₂S will be less than 20% of that normally produced in the cracking reaction. It can usually be handled in existing sulfur recovery facilities.

Commercial Experience

The Transox process has very recently been used for routine control of SO_x emissions from the FCC at Chevron's Cincinnati Refinery. The confidence to begin this application was based on extensive pilot plant testing and experience gained from two plant tests at Chevron's Hawaii and El Paso Refineries in 1980 and 1981. The plant tests were several weeks long and demonstrated stable control of SO_x emissions throughout the tests.

The FCC's at Hawaii and El Paso are both Esso Model IV's of about 20,000 BPD capacity. Both units operate in the promoted complete CO combustion regeneration mode, with typically 2-3% oxygen and essentially no CO in the regenerator flue gas. Both are equipped with electrostatic precipitators and CO boilers.

Table 1 lists feedstock properties for the two units. Both units run an unhydrofined vacuum gas oil containing about 1% sulfur. El Paso runs a small amount of metals-containing residuum and uses metals passivation technology to control gas make.

Transcat was added to each unit by replacing a small portion of the normal cracking catalyst inventory. Both the normal catalyst and Transcat were then made up to maintain their relative proportions in inventory. No other operational changes were made. There was no detectable effect of Transox operation on plant capacity, yield, or conversion in either test.

Table 2 shows typical SO_x levels for the two tests. Both units were operating prior to the test with a commercially available cracking catalyst which is known to possess some intrinsic activity for SO_x reduction. As Table 2 shows, Transox achieved a reduction of greater than 65% in both tests. It is estimated that Transox can achieve up to 90% SO_x reduction with cracking catalyst with low intrinsic activity for SO_x reduction, i.e., a base case flue gas SO_x level of 1000 ppm could be reduced to 100 ppm with Transox.

Metals deposition and the use of a metals passivator at El Paso apparently had little, if any, effect on Transox performance. In fact, the unique adsorptive and flushing characteristics of Transcat may actually enhance unit performance in a high metals content feed application by selectively trapping and removing some of those metals.

In developing Transox, every effort was made to achieve the goal of controlling SO_x while not increasing the emission of other pollutants. CO and particulates are of primary concern and are already covered by Federal NSPS. Emissions of CO were controlled by the use of combustion promoter. Table 3 shows the results of EPA Method 1-5 particulate testing at Hawaii. Within experimental error, no increase in mass rate of particulate emission was detected.

In some localities, NO_x emissions can be a problem. Under certain conditions, an increased NO_x level in the regenerator flue gas with the use of Transox was observed. However, the NO_x from the regenerator can be substantially reduced by proper operation of the CO boiler.

Table 4 shows representative NO_x levels before and after Transcat addition for the Hawaii and El Paso tests. In these tests, NO_x increased in the regenerator flue gas. At Hawaii, passing the regenerator flue gas through the CO boiler raised the NO_x in the base case and lowered

the NO_x during Transox operation. In part, this is due to mixing the regenerator flue gas with the flue gas produced by the fired burners in the CO boiler. However, the mixing effect does not account for all of the NO_x decrease that occurred across the two CO boilers during Transox operation. In both cases, over half of the total flue gas exiting the CO boiler came from the regenerator, yet the CO boiler NO_x level was less than half of the regenerator flue gas NO_x level. Also shown in Table 4 are the results of a test of CO boiler NO_x reduction at El Paso. A reduction of over 50% was achieved by a relatively inexpensive modification. Subsequent additional testing has demonstrated that any increase in NO_x resulting from the addition of Transcat can be avoided if appropriate modifications are made to the CO boiler.

TABLE 1. TYPICAL FEED ANALYSES

	Hawaii	El Paso
Sulfur, wt%	0.98	1.06
API Gravity	24.8	22.9
Ramsbottom Carbon, wt%	0.16	0.43
Distillation, LV%		
Start, Deg F	528	542
10	619	686
30	711	756
50	782	823
90	934	982
END	1044	1035
Metals, ppmw		
Nickel	—	0.8
Iron	—	6.0
Vanadium	—	0.4

TABLE 2. SULFUR OXIDE LEVELS IN REGENERATOR FLUE GAS

	Hawaii	El Paso
Base, ppmv	412	303
Transox, ppmv	86	103
Reduction, %	79	66

TABLE 3. PARTICULATE EMISSIONS—HAWAII FCC

	Base	Transox
Concentration, mg/dscf	6.0 ± 0.7	6.6 ± 0.3
Mass Rate, lb/hr	36 ± 4	37 ± 2

TABLE 4. NO_x LEVELS NORMALIZED TO 2% OXYGEN

	Hawaii	El Paso
FCC Flue Gas		
Base, ppmv	110	610
Transox, ppmv	700	1080
CO Boiler Stack		
Base, ppmv	140	—
Transox, ppmv	270	420
Modified CO Boiler, ppmv	—	190

TABLE 5. TRANSOX TRIAL AT CINCINNATI

	Partial Burn	Complete Burn	Complete Burn	TRANSOX
Feed S, Wt %	0.74	0.56	0.67	0.53
O ₂ in Regenerator	~0.7	2.9	3.3	3.9
Flue Gas, Vol %				
SO ₂ , vppm	—	711	959	260
SO ₂ , Lb/MBFF	824	566	632	170
Reduction*	Base	9	Base	66

*Weight Basis, Adjusted to Constant Feed Sulfur

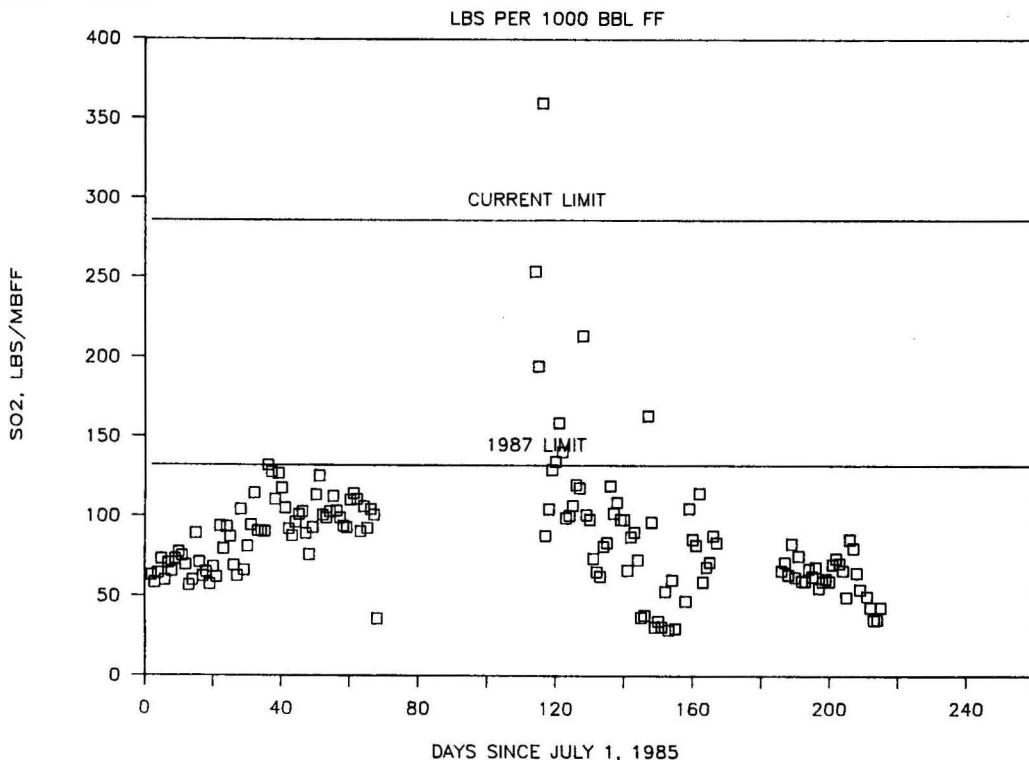


Figure 2. SO₂ emissions from El Segundo FCC.

The most recent commercial experience with Transox was a routine day-in-day-out SO_x emission control application on the FCC in our Cincinnati refinery. This unit was shut down in late May 1986 but ran for four months with over 60% reduction in SO_x emissions relative to the base case (no Transox) operation. The SO_x reduction program on this unit was actually executed in three stages:

1. Operation was shifted from partial to complete CO burn.

2. High sulfur slurry recycle was reduced to a minimum.

3. Transcat was introduced. The results of this program are summarized in Table 5. The shift to complete combustion reduced SO_x emissions about 9%. It turned out that slurry recycle was already at a practical minimum and as a result no reduction in SO_x was achieved from this step. The initial charge of Transcat was added over a two-day period. The unit ran smoothly during the Transcat addition and no change in conversion or selectivity was observed. An additional 66% reduction in SO_x emission relative to a base case in complete CO combustion was observed as a result of the Transcat addition. These results were obtained on a full riser, slide valve controlled, single regenerator cracker.

Chevron's El Segundo Refinery operates in the South Coast Air Quality Management District (SCAQMD). The current SO_x emission limit from FCC's operating within this District is 286 lb per 1000 bbl of feed. On January 1, 1987, this limit drops to 132 lb per 1000 bbl of feed. In Figure 2, both of these limits are superimposed on eight months of emissions data submitted by the refinery to SCAQMD.

It is obvious from Figure 2 that on some days in 1985 and 1986 SO_x emissions exceeded the new 1987 limit of 132 lb/1000 bbl of fresh feed. Present plans are to use

Transox as required during 1987 to give the flexibility of running higher sulfur feedstocks when that is an economic choice. It is fully expected that Transox will keep SO_x emissions below 132 lb/1000 bbl during these periods. No adverse process or environmental effects are anticipated as a result of the Transox application.

ECONOMICS

Recent estimates indicate that, for a 50,000 BPD FCC, a feed hydrofiner will require a capital investment of \$40 to \$50 million and a flue gas scrubber, \$15 to \$25 million. Combined operating costs and capital charges amount to about \$0.80/feed bbl for a hydrofiner and \$0.40/feed bbl for a scrubber. The hydrofiner estimate does not include credits for process and product quality benefits. These credits would reduce the cost attributable to SO_x control to an extent that is specific to the particular application and will in some cases make hydrofining an attractive alternative. In contrast to scrubbing and hydrofining, Transox involves essentially no capital investment. Typically, Transox operating cost (including Transcat, royalty, and NO_x Control if required) will be 3-6¢/bbl feed. This is more cost effective than the competing catalytic products currently being offered to the refining industry. Except under very special conditions, Transox is clearly more cost effective than either feed hydrofining or flue gas scrubbing.

APPLICABILITY TO OTHER UNITS

Catalytic SO_x control systems developed to date are not cost effective in all FCC units. In units operating in a partial CO combustion mode, the SO_x reduction achieved is

typically only 0-25% of that observed if the same unit were running in a full CO burn. Also, although many commercially available catalysts can be used with Transox, not all are compatible. Finally, due to design and operating differences, the level of SO_x reduction will

vary from unit to unit. However, one of the real advantages of Transox is that it requires essentially no modifications to the FCC. Thus, its performance in any FCC unit can be determined by a relatively inexpensive, full-scale test.



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Cool Water: Performance & Economics

Wayne N. Clark, Ronald L. Litzinger and I. R. Straughan

Cool Water Coal Gasification Program, Daggett, CA 92327

Cool Water is the United States' first Integrated Coal Gasification Combined Cycle (IGCC) power plant. This demonstration program has operated on many coals including low sulfur western coals and high sulfur eastern coals.

INTRODUCTION

The Integrated Gasification Combined-Cycle (IGCC) technology utilized at Cool Water represents a "marriage" of the chemical process industry with the electric utility industry. The chemical industry has brought its gasification, sulfur removal and sulfur recovery technologies together to provide a clean gaseous fuel for the electric utility industries' combined-cycle power generation technologies.

The combined technologies provide the electric utility industry with a means to produce power from coal, our nation's most abundant energy resource, in an environmentally acceptable manner.

We are pleased to report that the commercial-scale demonstration at Cool Water has been a successful one. Cool Water design confirmations, to date, have met and surpassed initial projections. The Cool Water Program has successfully demonstrated IGCC on a commercial-scale and has resolved many of the uncertainties previously associated with the technology. Overall emissions from the gasification/combined-cycle plant have been well below U.S. EPA New Source Performance Standards for coal-fired plants. IGCC represents a shelf-available technology for amelioration of acid rain concerns associated with the burning of coal. Preliminary economics indicate that IGCC technology is competitive with conventional coal-fired power plants equipped with flue gas desulfurization (FGD) units.

BACKGROUND [7]

Texaco Inc. (Texaco), which has developed its proprietary gasification technology over the last 30 years, and Southern California Edison Company (SCE), among the most innovative of utilities in the search for alternative power sources, began preliminary discussions in 1977 about the possibility of utilizing integrated gasification combined-cycle technology for the commercial production of electricity.

The Cool Water Coal Gasification Program (CWCGP) was formally initiated in July, 1979, upon execution of the original Texaco/SCE agreement which has since been amended to provide for contributions of capital and expertise by other participants. (See Table 1 and Figure 1 for Capital Funding and Participant's Program Functions, respectively.) The Electric Power Research Institute

TABLE 1. PROGRAM CAPITAL FUNDING
(MILLIONS OF DOLLARS)

Participants	Committed Funds
SCE	25
Texaco	45
EPRI	69
Bechtel	30
GE	30
JCWP	30
Subtotal	229
<u>Contributors</u>	
ESEERCO	5
Sohio	5
<u>Program Loan</u>	<u>24</u>
Subtotal	34
TOTAL COMMITMENT	263

(EPRI), Bechtel Power Corporation (Bechtel), and General Electric (GE) joined the Program in 1980. The Japan Cool Water Program Partnership (JCWP) joined in 1982 (JCWP is a consortium of interested Japanese industrial organizations led by Tokyo Electric Power Company, the world's largest private utility). Two non-equity contributors, The Empire State Electric Energy Research Corpo-

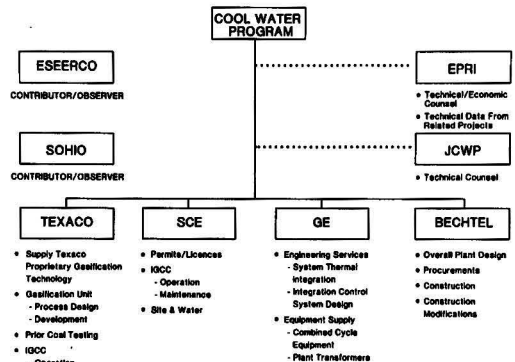


Figure 1. CWCGP participant's program functions.

ration and SOHIO Alternate Energy Development Company also joined the Program at a lower level of funding to obtain access to particular information from the Program. All capital costs were thus funded by private industry.

Detailed engineering for the plant began in February, 1980. Bechtel began construction in December, 1981, immediately after project funding was secured. Construction of the plant was completed ahead of schedule and under budget on April 30, 1984. The original concept of a \$300 million plant was pared to a \$294 million dollar capital budget. Final capital cost of the plant was \$263 million. A relatively good economic environment and effective cost management contributed to the downward revision of the plant's cost.

The plant, located adjacent to SCE's existing Cool Water Generating Station in the Mojave Desert approximately mid-way between Los Angeles and Las Vegas, began commercial production on June 24, 1984. The location gives CWCGP an excellent opportunity to prove an environmentally superior process for utilizing coal in a state with stringent environmental regulations. It is believed that if one can obtain the necessary permits to construct and operate a coal-based facility in Southern California, one should be able to do so most anywhere.

The Program began a five-year demonstration phase of commercial production on behalf of its joint owners in June, 1984. SCE has an option to purchase and operate the plant, assuming regulatory permit availability, for the subsequent 15 years.

The Program's principal objectives during the five-year demonstration phase (June, 1984, to June, 1989) are as follows:

1. Demonstration of acceptable system and equipment performance at a commercial scale.
2. Confirmation of system compliance with environmental criteria.
3. Verification of controllability of the integrated plant under all operating conditions.
4. Assessment of equipment and system reliability.

5. Preparation of operating, maintenance, safety and training procedures which could be applied to future plants.
6. Development of a complete economic and technical data base.
7. Demonstration of feedstock flexibility.

Operating successes in the two years of commercial production to date have placed the Program well underway in the accomplishment of these objectives.

PROCESS DESCRIPTION

The plant utilizes an entrained bed, oxygen-blown Texaco gasifier to convert 1,000 tons (907×10^3 kg) of coal per day to a medium-Btu synthesis gas. After particulate and sulfur removal, the gas is combusted in a gas turbine to produce electricity. In addition, steam is produced by recovering heat from the hot product gas in syngas coolers and from gas turbine exhaust gas in the heat recovery steam generator (HRSG). Steam from both sources is combined and superheated in the HRSG and then utilized in a steam turbine for production of additional electricity. A simplified block diagram of the CWCGP process is shown in Figure 2.

The Program coal is a specified Utah run-of-mine coal with approximately 0.5 wt. % sulfur. The Program has also tested two high sulfur Eastern United States coals. Illinois #6 coal, containing 3.1 wt. % sulfur, and Pittsburgh #8 coal, containing 2.8 wt. % sulfur, were successfully run in early 1986. Up to six other coals may be tested during the remaining portion of the five-year demonstration period.

COAL RECEIVING, HANDLING, AND SLURRY PREPARATION

Coal is delivered to the plant by rail in unit trains, bottom dropped from each hopper car and conveyed to storage. The coal is stored in two 6,000 ton ($5,433 \times 10^3$ kg) storage silos. Coal is transferred from the silos to live stor-

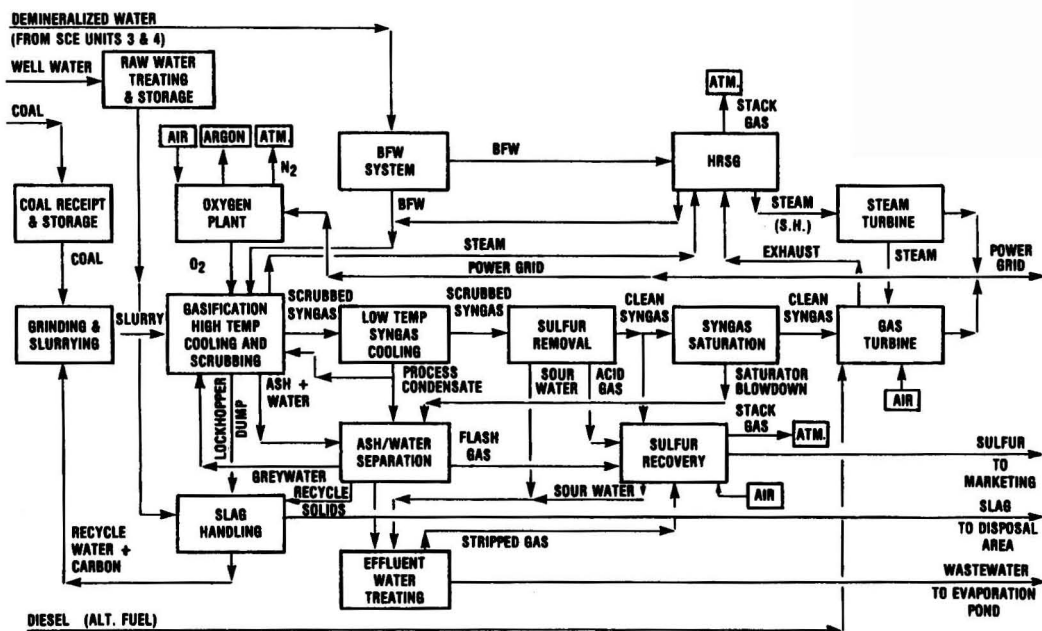


Figure 2. Cool Water Coal Gasification Program, block flow diagram.

age in the coal grinding area. The plant has been very successful in eliminating hot coal problems by using an inert gas system in the silos. The unloading facility and all conveyors are enclosed and equipped with dust collection and suppression equipment to minimize fugitive dust emissions.

Coal is combined with recycled ash and water from the gasification plant and pulverized in a wet rod mill. The plant has discontinued the use of cage mills for coal pre-crushing as modern continuous mining equipment produces coal with a size distribution making the pre-crushing unnecessary. Rod mill slurry is then transferred to one of the three slurry storage tanks. A transfer pump withdraws slurry from storage and feeds it to one of two high pressure, positive-displacement charge pumps. The charge pumps deliver the coal slurry to the gasifier. Slurry concentrations at the plant have typically ranged from 60% solids to 66% solids.

COAL GASIFICATION, SYNTHESIS GAS COOLING AND CARBON SCRUBBING

The coal-water slurry is combined with oxygen in a specially developed burner and fed into the refractory-lined gasifier. The partial oxidation reactions take place in the gasifier at 600 psig (4.1 MPa) and at temperatures in excess of 2000°F (1090°C). A medium-Btu synthesis gas consisting mainly of CO, H₂, CO₂ and steam is produced. A small amount of methane is also formed. Fuel-bound nitrogen is converted to nitrogen gas with some ammonia formed. Fuel-bound sulfur is reduced to H₂S with a small amount of COS formed. Table 2 shows the clean gas composition. The gas contains molten slag, some unconverted carbon and fine fly ash. The hot gas is first cooled in the radiant syngas cooler where 1,600 psig (11.0 MPa) saturated steam is produced. The molten slag droplets solidify and drop into a water sump at the bottom of the vessel where a lockhopper is used for its removal. The raw syngas is then cooled further in the convection syngas cooler. Additional 1,600 psig (11.0 MPa) saturated steam is produced in the convection cooler's evaporator section and boiler feed water is preheated in its economizer section. Raw syngas is then routed to the carbon scrubber where essentially all of the fine particulate material is removed by direct scrubbing with water.

Fly ash water from the radiant cooler sump and the carbon scrubber is routed to a settler where solids and water are separated. Recovered settler water is recycled to the carbon scrubber and/or rejected to the evaporation ponds. The settler bottoms are recycled to the coal grinding unit.

Slag from the lockhopper is discharged to a slag sump where a drag conveyor removes and dewateres the slag. Water from the slag sump is recycled to the coal grinding unit and the slag is discharged to a storage pad where it is loaded into trucks for disposal in a slag pit equipped with an impermeable clay liner.

Syngas from the carbon scrubber is cooled to 100°F (38°C) by successive heat exchange with saturator circulating water, steam turbine condensate, air and cooling water. Condensate is removed after each cooling step. A

portion of the condensate is recycled to the carbon scrubber while the remainder is rejected to a sour water stripper prior to discharge to the evaporation ponds. The cooled syngas flows to an acid gas removal unit where the sulfur compounds are removed.

The clean fuel gas goes to a saturator where the gas is reheated by direct contact with hot water. The water saturation provides the moisture required to control NO_x formation at the gas turbine. The saturated fuel gas is then superheated against economized boiler feed water. The saturator/clean gas heater system provides NO_x control while providing an excellent means to recover low level heat from the gasification plant and HRSG.

COMBINED-CYCLE UNIT

The superheated clean fuel gas flows to the gas turbine where it is combusted with air. Steam injection is also provided, as a backup to the saturator, for NO_x control. The turbine drives an electric generator producing approximately 65 MW of electric power.

The hot exhaust gas from the turbine is cooled in the HRSG by producing 1,450 psig (10.0 MPa) superheated steam at 950°F (510°C). The HRSG is composed of three sections: superheater, evaporator and economizer. The saturated steam raised in the HRSG evaporator is combined with saturated steam from the syngas coolers and routed to the superheater. The cooled flue gas is discharged to the atmosphere through the HRSG stack.

Superheated steam from the HRSG is utilized in the steam turbine which drives an electric generator producing approximately 55 MW of electric power. Steam for various plant uses is extracted from the turbine at a nominal 300 psig (2.1 MPa) and 15 psig (103 kPa). The steam turbine is a condensing type, with condensate recovered from the vacuum system being collected in the condenser hotwell. Makeup boiler feedwater is added to the hotwell to hold the system in balance.

The plant generates 117 MW of gross electricity at 13.8 kV. Approximately 6 MW of electric power is drawn through the plant's auxiliary transformer for internal use. The balance of the plant electrical output is transmitted to SCE's existing 220 kV grid through the plant's main transformer.

SULFUR REMOVAL AND RECOVERY

The reduction of fuel bound sulfur to H₂S and COS, during gasification, is a key to IGCC's environmental superiority. The plant uses acid gas removal, sulfur recovery and tail gas treating technologies, used for years by the chemical industry, to remove the sulfur compounds from the fuel gas prior to combustion and recover the sulfur compounds as salable elemental sulfur. The plant uses a Selexol unit for acid gas removal, a Claus unit for sulfur recovery, and a modified SCOT unit for tail gas treating.

The plant is designed to process a range of coals with sulfur contents from 0.35 to 3.5 wt. %. The 10 to 1 ratio of sulfur content has presented many interesting challenges during both design and operation of the plant. The removal of H₂S and COS remains at a constant 97% on the Selexol unit, while the coal sulfur content changes. Solvent conditions are adjusted to minimize CO₂ removal in the Selexol unit. The removal of CO₂ also remains constant at Selexol, while coal sulfur content changes. The syngas CO₂ content, however, is minimally altered as coals change. Therefore, the composition of acid gas varies considerably with changes in coal sulfur content. The plant uses a modified SCOT unit to increase H₂S concentrations to a level acceptable for the Claus unit.

TABLE 2. CLEAN SYNGAS COMPOSITION

Component	Mol % (Dry Basis)
CO	42.5
H ₂	38.2
CO ₂	18.6
CH ₄	0.3
Ar & N ₂	0.4
H ₂ S & COS	50 ppm

The plant's Claus unit processes acid gases from as low as 20 vol. % H₂S to as high as 60 vol. % H₂S.

The sulfur compounds are removed from the syngas at the Selexol absorber. Incoming gas and solvent are chilled to 10°F (-12°C), using an ammonia refrigerant, to enhance COS removal. The chilled solvent is used in lieu of a COS hydrolysis unit upstream of the Selexol unit. The plant was designed for a H₂S/COS ratio of 19/1 in the syngas. The H₂S/COS ratio of the syngas, however, has been 50/1 which enables the plant to maintain 40°F (4°C) at the Selexol absorber and still achieve 97% sulfur removal.

This was welcome news as operation at 40°F (4°C) significantly reduces the "aggravation factor" associated with additional refrigeration compressors and operation below the freezing point of water.

Rich solvent is then steam stripped and the lean solvent is sent back to the absorber. The Selexol unit acid gas is routed to the front amine absorber in the modified SCOT unit where H₂S is removed. The rich amine from the front amine absorber combines with the rich amine from the tail gas treating amine absorber and is charged to the amine stripper. The H₂S is stripped from the amine and routed to the Claus unit. Vent gas from the front amine absorber is routed to the SCOT unit's catalytic reactor. This additional H₂S concentration step in the modified SCOT unit eliminates the need for a recycle compressor at the Selexol unit's rich solvent flash drum to increase the H₂S concentration. Typically a 5-6 vol. % H₂S acid gas is concentrated to 20 vol. % acid gas in the modified SCOT unit.

Acid gas from the SCOT stripper is converted to elemental sulfur at the Claus unit via the classic Claus reaction. Supplemental natural gas firing, acid gas preheaters and air preheaters are used during low sulfur coal operation as the lean acid gas will not maintain desired thermal reactor temperatures on its own.

The preheaters operate the same as the system reheaters using 600 psig (4.1 MPa) steam from the Claus boiler. The richer acid gas produced during high sulfur coal operation is self-sufficient to maintain desired thermal reactor temperatures and the additional heating described above is not required. The liquid sulfur produced at the Claus unit is stored and then pumped to the buyer's truck F.O.B. the plant.

The tail gas from the Claus final sulfur condenser is combined with vent gas from the front amine absorber and is routed to the modified SCOT unit's catalytic reactor. Residual SO₂ in the Claus tail gas is reduced to H₂S in the SCOT reactor. The SCOT reactor also hydrolyzes COS in the front amine absorber vent gas to H₂S. Effluent from the SCOT reactor is routed to the tail gas treating amine absorber in the modified SCOT unit where H₂S is removed. The vent gas from the tail gas treating amine absorber is incinerated for destruction of any remaining sulfur compounds.

Single pass conversion at the Claus unit when operating on high sulfur coal is high enough that tail gas treating is not required to meet permit limits. The Program, however, continues to operate the SCOT reactor to stay well below permit levels and further demonstrate the cleanliness of the IGCC technology.

ADDITIONAL SUPPORT SYSTEMS

Oxygen is supplied to the gasifier from an "across the fence" cryogenic air separation plant which also provides the plant's nitrogen requirements. The oxygen plant typically draws 17 MW of electric power.

Raw and demineralized water are supplied from SCE's adjacent Cool Water Generating Station. IGCC plants use 33% less water than conventional coal-fired power plants

with flue gas cleanup.

The plant cooling tower is of the mechanical-draft, evaporating type and serves to remove heat from three separate cooling water systems. The circulating water system is a high flow-low head system serving the steam turbine surface condenser. The open cooling water system is a low flow-high heat system serving the process plant coolers. The closed cooling water system is a high purity and corrosion inhibited system supplying cooling to specific systems requiring the high purity water such as pump seals. Heat from the closed system is rejected to the circulating water system. Cooling tower make-up is primarily well water with steam system blowdown streams providing additional make-up.

The plant is also equipped with other utilities required for operation including a flare system and plant air systems.

OPERATING EXPERIENCE

Cool Water operating experience to date has exceeded expectations and continues to improve.

The gasifier was first fired on May 7, 1984. Plant commissioning was completed only 47 days later and the Program entered commercial production to begin the five-year demonstration period on June 24, 1984. Table 3 summarizes the plant commissioning accomplishments versus objective dates:

The plant is currently operating at full production rates. Table 4 summarizes production to date:

The plant has during several months had capacity factors greater than 70%, including September, 1984, which was only four months after startup. The capacity factor is the power produced as a percent of design. In September, 1985, the plant had its best production month to date with a capacity factor of 86%. Production goals continue to increase (even though scheduled demonstrative testing has a significantly adverse impact on full design production) and the Program will aggressively pursue them.

PERFORMANCE

Gasifier performance has been higher than originally expected on a commercial scale. Burner developments have eliminated any scale-up concerns. Single-pass carbon conversions have been greater than 98 wt. % when operating on Utah coal. Single-pass carbon conversion on Illinois #6 coal averaged 96.5 wt. %. Single-pass carbon

TABLE 3. CWCGP COMMISSIONING ACCOMPLISHMENTS

	Objective Date	Achieved
First Btu	6/01/84	5/07/84
Initial Electrical Production from Syngas	6/25/84	5/20/84

TABLE 4. CWCGP CUMULATIVE PRODUCTION (6/24/84-2/28/87)

Gasifier Operation (Hrs.)	14,254
Coal Gasified (Tons Dry)	575,071
Coal Gasified (10 ³ kg)	479,000
Gross Electrical Production (kWh)	1,393,363,000
On-Stream Factor	60.6%
Capacity Factor	50.3%

conversions have been high enough that the slag screen classifier, intended to separate fine high carbon content slag for recycle to the gasifier, has not been utilized. Settler bottoms are still being recycled to the gasifier and increase overall conversions 0.5 wt. % above single-pass conversion values.

The high carbon conversions are also being attained at lower reaction temperatures than originally expected. The lower gasification temperatures have lowered oxygen costs and extended refractory life. Actual oxygen consumption has been 6 percent lower than the design value. Gasifier refractory life is presently estimated to be three-year actual versus a one-year design value on low sulfur Utah coal.

Plant heat rates have also been in line with original projections. Overall plant heat rates have been approximately 11,300 Btu/kWh (11,922 kJ/kWh). Overall plant heat rates are given in equation below:

$$\text{Overall Plant Heat Rate} = \frac{\text{Coal Energy Input}}{\text{Gross Pwr. Prod. - Pt. Aux. Pwr. - Oxygen Pt. Pwr.}}$$

The Program projects that the heat rate can be lowered to 10,900 Btu/kWh (11,500 kJ/kWh) with minor plant improvements. EPRI has estimated a mature second generation IGCC plant would have a heat rate of 9,200 Btu/kWh (9,700 kJ/kWh) [2]. EPRI's estimate is based on the design heat rate for the Cool Water plant and various commercial plant adjustments including higher gas turbine firing temperatures, reheat steam cycle and lower oxygen purity. Coupling these with the minor plant improvements discussed above could lower EPRI's estimate even further.

The plant has completed partial load and dynamic load following tests. Partial load tests were conducted at 70% of the design plant throughput. Carbon conversions decreased slightly and heat rates increased approximately 16% at the partial load.

Dynamic load following tests have also gone well. Load changes of 20% at a ramp rate of 4%/min. have been achieved. Dynamic testing has been conducted with the gasification plant leading and with the combined-cycle unit leading. Utility daily load following requirements are usually 10% to 50% load changes at 1%/min. to 3%/min. [3]. Test results are within the ranges stated above and have demonstrated the IGCC's capability of serving the electric utility industry.

EMISSIONS

Environmental performance at the plant has been outstanding. Sulfur removal and recovery technology have made IGCC an extremely clean process. Total plant SO₂ emissions (HRSG plus incinerator) are shown versus the U.S. EPA New Source Performance Standards in Table 5.

TABLE 5. CWCCP EMISSIONS

	U.S. EPA NSPS	Test Results
Utah Coal (0.5 wt. % S):		
Lb. SO ₂ /million Btu Coal	0.240	0.021
kg SO ₂ /million kJ Coal	0.103	0.009
Ill. #6 Coal (3.0 wt. % S):		
Lb. SO ₂ /million Btu Coal	0.600	0.130
kg SO ₂ /million kJ Coal	0.258	0.056
Pittsburgh #8 (2.8 wt. % S):		
Lb. SO ₂ /million Btu Coal	0.600	0.086
kg SO ₂ /million kJ Coal	0.258	0.037

SO₂ emissions are typically 10-20% of the allowable levels under the U.S. EPA New Source Performance Standards. Sulfur removal from syngas has ranged between 97% and 99%. Overall sulfur recovery from coal is typically 96%.

Stack emissions of NO_x and particulates have also averaged about 10% of allowable levels under the U.S. EPA New Source Performance Standards.

Slag from the gasifier has been classified non-hazardous by RCRA standards and by the California Department of Health Services. The Program is currently pursuing markets for its slag as either filler or abrasive material.

ECONOMICS

Economic evaluations at Cool Water continue. One of the Program's principal objectives is to establish a realistic basis for determining the cost of electricity from a mature IGCC plant.

Recent studies by EPRI indicate the capital investment required for a 500 MW IGCC plant is comparable to that for a direct coal-fired plant of the same size with flue gas cleanup [4] and that phased construction of the IGCC plant would result in a cost approximately 10% less than the direct coal-fired unit. Phased construction of an IGCC plant would consist of initially installing a gas turbine to be fired with natural gas or oil while these fuels were available at a low cost. Heat recovery steam generators and a steam turbine would be added as load growth continues and then the gasifier plant would be added to convert the plant to IGCC operation.

Phased construction offers several advantages. Large capital outlays are stretched out by matching load growth requirements more evenly, significantly minimizing the "at-risk" capital due to the short construction time-frame for the modules. Phased construction also improves the potential to take advantage of non-utility ownership of the gasification plant.

EPRI studies further indicate that the operating costs of a mature IGCC plant would be 10% lower than a direct coal-fired plant using flue gas cleanup [5]. It is too early for the Program to make definitive statements, however, the Program feels the estimates are conservative. Preliminary analysis by the Program indicate that production costs continue to decrease as the plant progresses on its learning curve. These operating cost declines combined with atypical expenses for a demonstration plant provide insight to determine projected commercial production costs. Atypical costs include special testing costs, extraordinary record keeping costs and high oxygen costs due to the short amortization period for the oxygen plant.

AWARDS

To date, the Cool Water Plant participants have won several prestigious engineering awards. The Texaco Coal Gasification Process received honorable mention for the "1985 Kirkpatrick Chemical Engineering Honor Award" for its commercialization of the technology. *Chemical Engineering* magazine cited the use of coal gasification for electric power production as an example of the diversity of the process [6].

The Institute for the Advancement of Engineering called the Cool Water Plant "possibly the world's cleanest coal-using plant." The IAE presented the plant with its "Outstanding Engineering Achievement Award" in 1986. *Power Magazine* awarded the plant with its "1985 Environmental Protection Award" which recognizes utility industry achievements in protecting the environment.

In 1984, the National Council on Synthetic Fuels Pro-

duction awarded the first "Walter Flowers Achievement Award in Synthetic Fuels" to Texaco Inc. for development and commercialization of the Texaco Coal Gasification Process. EPRI received the second Walter Flowers Award, in 1986, for its contribution in synthetic fuels related to the Cool Water project.

SUMMARY

The Cool Water Program has demonstrated that IGCC technology is an excellent alternative to produce electric power from coal in an environmentally acceptable manner. Chemical process technology is an integral part of the IGCC technology in that it provides the clean gaseous fuel for the power production. The excellent demonstration at Cool Water has proven the technology and opened the door for future opportunities.

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Wayne N. Clark received a BS in Engineering from San Diego State and received his MBA from the University of Southern California. He is a registered Professional Engineer in both California and New Jersey. He joined Texaco Inc. at Los Angeles in 1968, and held several positions for the Company within Southern California until his transfer to Texaco's Eagle Point Refinery at Westville, NJ. As Assistant Plant Manager of Maintenance and Construction, he was responsible for all engineering, maintenance, construction and environmental activities. He joined the Cool Water Program in 1981, where he initially input into the Engineering & Design and was later responsible for the construction and operation of its first-of-a-kind Coal Gasification, Combined Cycle Plant.



Ronald L. Litzinger is a Process Engineer for Southern California Edison. He holds a BS in Chemical Engineering from the University of Washington. He has been with the Cool Water Coal Gasification Program since the plant started in 1984 and has held positions in both the Engineering and Operating Departments. Prior to joining Edison in 1986, he worked for Texaco Inc. in the coal gasification field. He is a registered Professional Engineer in California.



Ian R. Straughan earned his Bachelor Degrees in Applied Geology and Zoology and his Ph.D. in Philosophy from the University of Queensland. He began his career at Southern California Edison Company in 1974, as a Senior Research Scientist in the Research and Development Organization, with responsibility for environmental and biomass energy systems research. In 1978, he was awarded the title of Consulting Scientist, and in 1980 promoted to Manager, Systems Research and Development. In this position, he had responsibility for environmental research and the deployment of renewable technologies on the Edison system. He was appointed Director of Research and Development in 1984. He is presently Manager of Resources with responsibility for Electric System Planning, and has served on numerous federal, state and industry panels on alternate energy sources.

Reduced Sulfur Emissions With the Cope™ Process

By

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*The Claus Oxygen-based Process Expansion (COPE™) technology
can increase a Claus unit's capacity at substantially reduced capital
cost over conventional alternatives.*

COPE™ TECHNOLOGY: AN IMPROVEMENT TO THE MODIFIED CLAUSS PROCESS

The Claus Sulfur Recovery Unit (SRU) has been used in industry for well over 100 years for the purpose of converting H_2S to elemental sulfur. Typical sulfur recovery in an SRU is about 95-97% for a relatively rich acid gas feed, when using at least three catalytic stages and modern-day reheat schemes (Figure 1).

With the advent of more stringent air pollution regulations and the issuance of the EPA federal regulation entitled "40 CFR Part 60" in 1976 (pertaining to SRU's in refineries), the emission level from such a unit is set at 250 ppm SO_2 in the stack (on the basis of 0% excess air and 0% moisture) or 10 ppm H_2S if an incinerator is not used. On October 1, 1985, revised New Source Performance Standards (NSPS) for SRU's located at gas plants were issued in the Federal Register under "40 CFR Part 60 Sub-part LLL." These new NSPS standards require all future on-shore natural gas process plants (>2 LTPD sulfur) to have higher sulfur recovery efficiencies, the actual recovery determined by the acid gas concentration and the size of the plant. Various Tail Gas Cleanup Unit (TGCU) processes may be required to meet the more stringent emission standards for gas-plant SRU's.

Overall long-term trends indicate an increase in the sulfur content of refinery crude oil and natural gas. In many locations, it may become necessary to install additional Claus SRU and TGCU capacity or to debottleneck existing units while simultaneously meeting more stringent sulfur emission standards.

The COPE technology (Claus Oxygen-based Process Expansion) permits enriching the air stream from 21% up to 100% with pure O_2 , while handling a rich acid gas feed in an SRU without exceeding a typical reaction furnace temperature limitation of about 2700°F. As shown in Figure 2, this process is a simple modification to the conven-

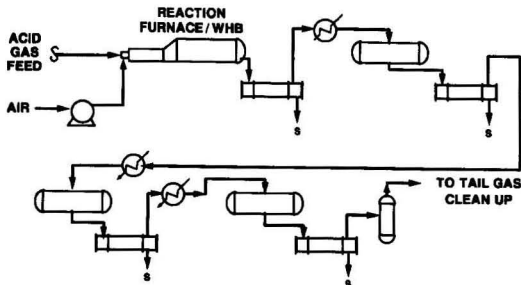


Figure 1. Conventional Claus sulfur recovery unit.

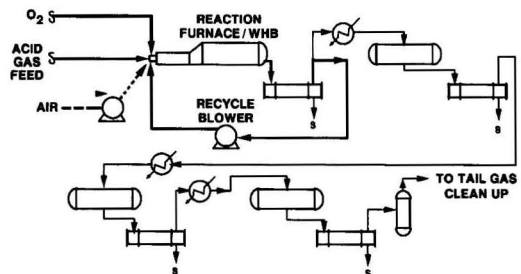


Figure 2. COPE sulfur recovery unit.

ditional Claus process. A portion of the first sulfur condenser effluent gas is recycled to the reaction furnace burner by a hot gas recycle blower. The recycle stream acts as a coolant (in place of N_2) to moderate the furnace temperature to the desired level. Note that the recycle blower could also be located immediately after the first

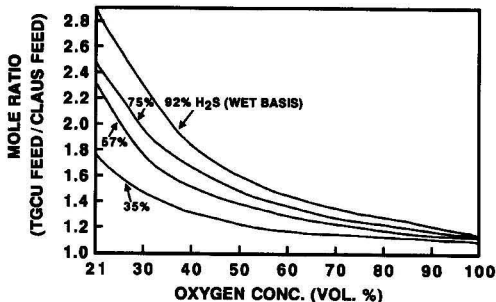


Figure 3. Relative tail gas flow rate.

condenser such that the pressure of the net forward flow could be increased for additional capacity.

While this recycle flow increases the relative contribution of the front-end pressure drop (through the burner, furnace, waste heat boiler, and first sulfur condenser), the flow through the converter section and the TGCU is greatly reduced (Figure 3), yielding a corresponding reduction in the overall system pressure drop. Thus, the use of oxygen to reduce or eliminate the nitrogen flow through the SRU permits an increase in the acid gas feed flow to bring the overall plant pressure drop back to the original air-based value. The absence of diluent nitrogen also yields other benefits: 1) improved sulfur conversion in the Claus process, and 2) reduced sulfur emissions from the TGCU.

COMMERCIAL PLANT DATA

Before the effect of COPE operation on sulfur emissions is examined in greater detail, results of the first

commercial installation will be presented to establish a basis for projected results. The first COPE installation was commissioned on one of two 108 LTPD 3-stage Claus SRU's at Conoco's refinery in Lake Charles, LA and has been in successful commercial operation since March 1985. A second unit was also modified and started up in May 1985.

The test data from the first installation are shown in Table 1 for three oxygen enrichment levels (65%, 54%, and 40% O₂) at two acid gas flow rates (250 and 200 MSCFH, corresponding nominally to 200 and 150 LTPD sulfur, respectively).

Table 1 shows the overall Claus SRU's performance, temperature, and pressure profiles derived from the test results. It is remarkable to note that the capacity was increased by 81-84% at O₂ enrichment levels of 54-65% while improving overall Claus sulfur recovery.

Enrichment to 100% O₂ was not tested due to thermal limitations of the waste heat boiler. The furnace temperatures were effectively controlled within the range of 2350-2570°F by adjusting the recycle gas stream.

Note that the pressure drop through the furnace section increases while the overall pressure drop is reduced. For example, comparing the air-base case to the Test 1 (65% O₂) results shows an increase from 1.8 to 3.6 psi through the furnace section and a corresponding reduction from 9.8 to 8.6 psi for the system pressure drop.

PROJECTED TGCU PERFORMANCE WITH COPE SRU

To take full advantage of the COPE SRU capacity increase, additional sulfur in the tail gas must also be processed. A standard hydrogenation/selective amine tail gas unit using a state-of-the-art, commercially available selective amine can handle the increased sulfur load with minimal alterations. The analysis will show that the sulfur concentration in the TGCU effluent to the incinerator can

TABLE 1. COPE PERFORMANCE DATA (APRIL 22-24, 1985) CONOCO, INC. — LAKE CHARLES, LA

Overall Claus Performance	Test 1	Test 2	Test 3	Projected Air Base ^a
% O ₂ Enrichment	65	54	40	21
Achieved Sulfur Capacity, LTPD	199	196	152	108
Acid gas flow, MSCFH ¹	253	249	195	138
Acid gas H ₂ S, mol% (dry gas basis)	93.7	93.9	93.1	93.7
Air flow, MSCFH	78	125	176	304
COPE O ₂ flow, MSCFH	100	90	55	0
Tail gas flow, MSCFH	320	356	343	381
Overall Claus sulfur recovery, % ²	97.9	97.7	97.6	97.4
Temperature Profile, °F				
Acid gas feed (preheated)	365	361	368	365
Reaction furnace	2568	2515	2345	2374
Waste heat boiler outlet	765	762	725	680
1st Condenser outlet	326	324	328	326
1st Converter in/out	413/643	413/634	413/610	513/632
2nd Converter in/out	438/497	438/492	437/481	438/484
3rd Converter in/out	434/441	433/440	432/439	434/443
2nd Condenser outlet	305	314	307	305
3rd Condenser outlet	287	288	294	287
4th Condenser outlet	286	287	286	286
Pressure Drop, psi				
Preheater/furnace section	3.6 (42%)	3.7 (38%)	N/A	1.8 (18%)
Converter section	3.0 (35%)	3.6 (37%)	N/A	4.4 (45%)
TGCU section	2.0 (23%)	2.5 (25%)	N/A	3.6 (37%)
Total	8.6 (100%)	9.8 (100%)	9.4 (100%)	9.8 (100%)

¹ Saturated with H₂O at 115°F.

² Sulfur conversion is somewhat higher since there is sulfur vapor loss and entrainment from the last sulfur condenser.

³ Computer simulation initialized to the performance test data.

TABLE 2. PROJECTED SULFUR BALANCE
SELECTIVE AMINE TAIL GAS UNIT

Stream:	65% O ₂ COPE Mode	Air-Based Mode
○ SRU acid gas feed, MSCFH	253	138
contained sulfur, LTPD	199	108
○ SRU tail gas, MSCFH	320	381
contained sulfur, LTPD	4.3	2.9
○ Feed gas to amine absorber, MSCFH	104	278
contained sulfur, LTPD	4.3	2.9
○ Net tail gas to incinerator, MSCFH	97	275
contained sulfur, vppm	80	80
lb/hr	0.8	1.9
Overall sulfur recovery	99.99	99.98
OVERALL SO ₂ EMISSIONS:		
Stack gas flow, MSCFH	121	335
SO ₂ , vppm	60	65
SO ₂ , TPY	5.7	16.2

be maintained at the air-based value, even with an 85% increase in acid gas feed in the COPE mode, resulting in a 60-80% reduction in total sulfur emissions.

Table 2 summarizes the detailed tail gas process information shown in Table 1 for both the 65% O₂ COPE operation (Test 1) and a base air case. Due to the increase in sulfur conversion with this technology, the amount of sulfur in the tail gas is proportionally less ($4.3/2.9 = 1.5$) than the feed capacity increase ($199/108 = 1.85$).

Furthermore, with 65% O₂ COPE operation, the tail gas rate declines despite the 85% feed increase, more than doubling the sulfur species concentration.

A typical hydrogenation/selective amine tail gas unit flow scheme is shown in Figure 4. Operation of the tail gas system requires one modification when operating the Claus unit in the COPE mode (in this analysis, at 65% O₂). Since the amount of water in the tail gas is proportional to sulfur production, additional surface area must be added to the quench tower heat exchanger to condense additional water from the COPE capacity increase.

In all other aspects, operation will not be significantly different. The hydrogenation section benefits from the higher hydrogen and carbon monoxide partial pressure and the reduced flow rate. Both of these benefits effec-

tively increase the capacity.

In the amine absorber, the higher H₂S partial pressure and lower gas rate offset the 50% increase in sulfur loading, and no increase in amine circulation rate is required. Since the H₂S concentration in the net overhead gas is controlled by mass transfer from the gas to the liquid phase, the lower gas rate results in increased contacting time and the ability to achieve the same H₂S concentration at the base air operation. Slightly more CO₂ co-absorption results, but this has no significant effect.

Additionally, the tail gas flow rate is so reduced that it could be compressed and fed to a central fuel gas system instead of incinerating.

UTILITY/COST SUMMARY

Tail gas unit utility consumption is summarized in Table 3. The lower COPE tail gas rate results in a large reduction in fuel for both the reducing gas generator and the incinerator, with proportionally less steam produced. The amine stripper reboiler steam consumption is virtually unchanged, since it is nearly proportional to amine circulation rate; and the net steam produced in the SRU

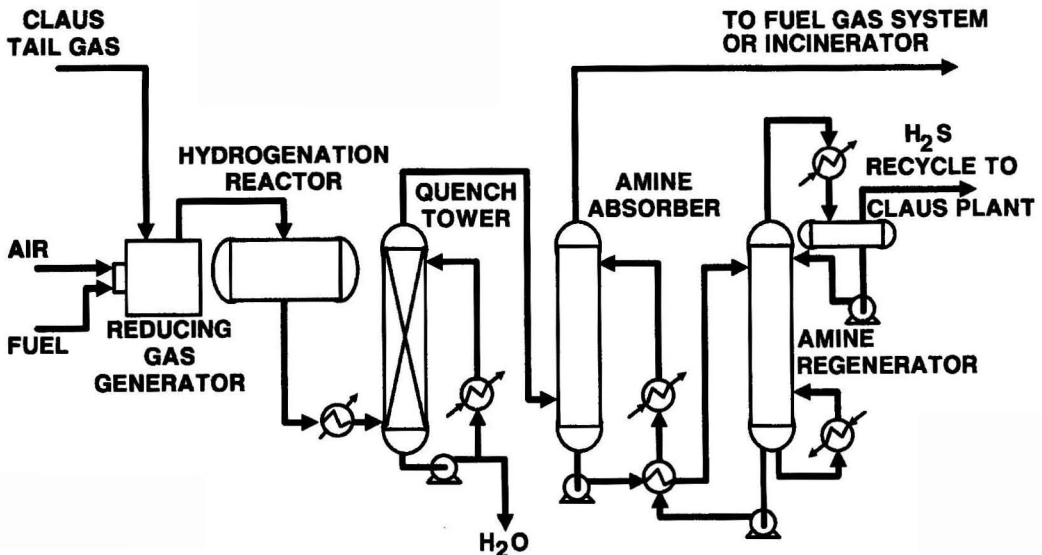


Figure 4. Selective amine tail gas cleanup unit.

TABLE 3. UTILITY COMPARISON (RETROFIT)
BASIS: 93.7% H₂S (DRY BASIS)

	Air-Based	65% O ₂ COPE Retrofit
I. Sulfur Capacity, LTPD	108	199
II. Utility Summary		
A. Claus SRU (3 converters)		
Air blower, ¹ KW	158	40
COPE recycle blower, KW ($\Delta P = 3$ psi)	—	39
Acid gas preheater, MMBtu/hr	0.72	1.30
WHB steam (credit), MMBtu/hr	(25.10)	(48.40)
Reheat steam, MMBtu/hr	3.60	2.55
Condenser Steam (credit), MMBtu/hr	(7.1)	(11.0)
O ₂ , st/d	—	101
B. Selective Amine TGCU		
Cooling water, gpm	1170	1762
Electricity (air blower ¹ & pumps), KW	54	45.8
Fuel gas	4.9	2.4
(hydrogenation & incinerator), MSCFH		
Net 50# steam (credit), MMBtu/hr	(0.40)	(0.14)
450# steam (credit), MMBtu/hr	(5.44)	(0.94)
III. Incremental Sulfur Cost ² (\$/LT)	100	32

¹ 60% adiabatic compressor efficiency and 90% driver efficiency.

² Fully-costed basis, including capital charges.

increases proportionally to sulfur capacity. Assigning representative unit costs to these utility figures yields an incremental cost of \$32/LT for capacity produced by this process, as compared to \$100/LT for new, air-based capacity with a TGCU. The incremental costs cited include the difference between the capital required for the COPE retrofit (approximately \$1.7MM) and the air-based plant with a TGCU (approximately \$13MM).

CONCLUSION

The COPE process represents a commercially demonstrated means for constructing new or expanding existing Claus SRU capacity in refinery and gas plant operations. Improved Claus conversion and higher sulfur recovery result in a 60-80% reduction in SO₂ stack emissions downstream of a typical TGCU.

ACKNOWLEDGMENT

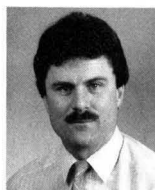
The authors gratefully acknowledge Conoco Inc.'s substantial role in successfully implementing COPE technology at the Lake Charles refinery.



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Landfill Gas: Asset or Liability?

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Management of LFG is a growing concern nationwide. LFG recovery is one option that should be evaluated by landfill owners as a means to meet odor control, closure regulations and/or to help offset long-term closure costs.

INTRODUCTION

The landfill gas (LFG) industry came into being a little more than a decade ago. The energy crisis of 1974 had hit and subsequent public concern over diminishing world energy reserves and rising fuel prices kindled interest in alternative energy sources. Efforts to mechanically control migrating LFG had been successfully carried out at several Los Angeles County Sanitation District landfills by the mid-1970's. Since simply flaring LFG seemed wasteful, ways to utilize it began to be explored. In June 1975, the first commercial recovery project went into operation at the Palos Verdes landfill in the Los Angeles area. The processed LFG was sold to the Southern California Gas Company.

Before the benefits of LFG recovery were generally known, landfill owners, public and private, basically ignored the LFG problem or claimed that their landfill did not generate gas. Throughout the 1970's, most landfill owners were skeptical about LFG recovery due to various technical problems in the early plants. This resulted in slow growth for the industry.

By 1980, only four high-Btu and four medium-Btu plants were in operation, all but two of which were in California. The use of LFG to generate electricity was being discussed, but the first plant was still more than two years away.

As more plants came on line, the pace of development took a dramatic upswing. Today, there are 23 facilities producing high- and medium-Btu gas and 21 LFG-fueled electricity generating plants. Developers and consultants claim to have firm contracts to build and start up more than 24 additional LFG plants—mostly for electricity generation—by mid-1986. Numerous public agencies are also including LFG recovery projects in long-range planning and budgeting. As many as 152 cities and 65 counties have plans for recovery programs at their landfills, according to a 1985 survey.

The environmental benefits that result from LFG recovery projects have become important to landfill owners. Through recovery, for example, landfills can more easily comply with environmental regulations as a result of reduced potential for lateral migration and reduced atmospheric emissions. Also, within the past few years, state design and closure regulations specific to LFG control have increased landfill owner's interest in recovery projects to help offset costs for regulatory compliance.

This paper addresses the basics about LFG and can

serve as a reference on LFG issues. The following topics will be discussed:

1. How is LFG generated?
2. What is the concern about LFG?
3. How can LFG be controlled?
4. Suitable sites for LFG recovery.
5. Utilization options for LFG.
6. "Who's who" in the LFG industry?
7. Main characteristics of the current regulatory climate.

LFG GENERATION

Years ago, open burning dumps were the nation's most prevalent method for solid waste disposal. Although burning resulted in waste volume reduction, increasing public concern regarding the environmental and health impacts of this haphazard refuse disposal technique resulted in the development of sanitary landfilling practices in the late 1960's. Ironically, while many of the nuisances and health hazards associated with burning dumps were eliminated, compacting and covering refuse with layers of soil resulted in a totally new environmental problem—the production of LFG.

When organic matter decomposes, it is transformed by the action of microorganisms abundant in solid wastes into a variety of simpler organic materials. Byproduct gases, principally methane (CH₄) and carbon dioxide (CO₂), are produced along with lesser amounts of ammonia, hydrogen sulfide and other trace gases. The formation of LFG undergoes an evolutionary process as the waste is exposed to aerobic (oxygen-abundant) conditions, followed by anaerobic (oxygen-deficient) conditions. This evolution proceeds in three phases:

1. In the first phase, lasting a few weeks or less, aerobic bacteria, present at the time of waste disposal, begin the decomposition process by consuming the oxygen to a point of depletion. Carbon dioxide and water are the principal byproducts.
2. During the second phase, a group of facultative organisms called acid formers (i.e., they can tolerate oxygen, but are not dependent on its being present) become dominant. Significant amounts of carbon dioxide, volatile acids, and some hydrogen are produced in this phase.
3. In the third phase, the methane-forming bacteria become dominant and steady-state LFG production re-

sults. Oxygen in any amount destroys the activity of these anaerobic organisms. A reduction in the amount of carbon dioxide and a depletion of hydrogen is observed. Methane in the LFG usually ranges from 40-60%, with a corresponding range for carbon dioxide of 60-40%.

The rate of LFG production is dependent on a number of site-specific factors. Among these are refuse composition and age, moisture content, depth, pH and temperature. Steady generation typically occurs within one to two years after original landfilling has taken place and can continue after filling has been completed for 10 to 30 years, depending upon local conditions. Thereafter, gas volumes steadily decline but generation can continue for many more years.

Actual recovery rates vary depending on the porosity and depth of the cover materials, the general landfill configuration, collection system inefficiencies, and withdrawal rate limitations imposed to prevent excessive air intrusion. A commonly used range for potential recovery rates is 80 to 280 standard cubic feet of LFG per ton of solid waste per year.

WHAT IS THE CONCERN ABOUT LFG?

The methane content is explosive in concentrations between 5-15% by volume in air when an ignition source is present. Thus, it is important to control its movement. LFG disperses in all directions from the landfill mass by convection and/or diffusion. Convection is movement and response to a pressure gradient with movement in the direction of decreasing pressure. Diffusion is movement in response to a concentration gradient.

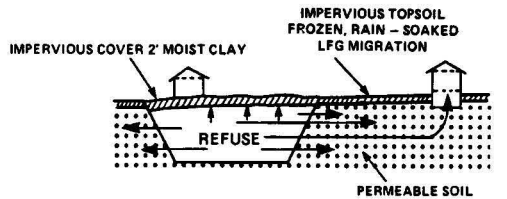
The Environmental Protection Agency (EPA) through the Resource Conservation and Recovery Act (RCRA) in 1976 established guidelines for control at all operating landfills. Local governments have adopted or expanded on the RCRA guidelines, and have applied requirements to closed landfills as well. According to RCRA, methane concentrations must exceed neither 1.25% by volume in an enclosed structure nor 5% by volume (10% LFG) at the landfill property boundary. In response to growing concerns regarding potentially toxic emissions and air quality degradation resulting from LFG production, regulations have been developed in California to limit surficial gas emissions from both active and closed landfill sites.

Gas produced in a landfill escapes from the refuse by either vertical or lateral migration. (See Figure 1.) Since gas is lighter than air, it tends to rise through the landfill cover if the cover is sufficiently permeable. However, if this vertical path is sealed by rain-saturated or frozen cover soil, pavement, or a clay/bentonite or synthetic membrane cap, the tendency towards lateral migration is increased greatly. It is this lateral migration—often to significant distances—which heightens liability problems and hazards.

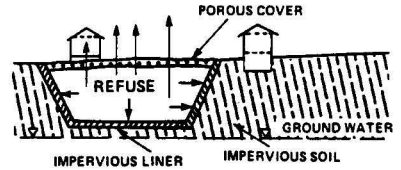
Movement occurs in sand, silt, or clay soils as long as there are continuous voids. The rate of movement decreases as pore size decreases; therefore, movement is greatest through highly porous sands or gravel and slightest in dense clay soils. Drainage culverts and subsurface utility corridors can provide a conduit for LFG migration.

HOW CAN LFG BE CONTROLLED?

To determine whether offsite migration of LFG is occurring, monitoring wells or probes need to be installed at the perimeter of the landfill. Individual monitoring wells are most effective when equipped with multiple probes to determine LFG concentrations at various depths. (See Figure 2.)



LATERAL LFG MIGRATION



VERTICAL LFG MIGRATION

Figure 1. Lateral and vertical LFG migration.

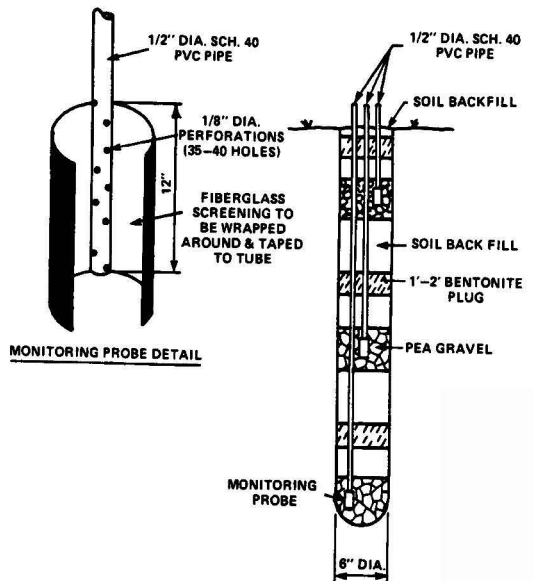


Figure 2. Typical LFG monitoring well.

Selecting a control system that will meet RCRA requirements demands consideration of such site-specific factors as:

1. Landfill depth and depth to groundwater.
2. Age, composition and moisture content of refuse.
3. Distance to property lines.
4. Type and location of proposed on- and off-site developments.
5. Subsurface soil characteristics.
6. Characteristics of closure cap and vegetative cover.

There are two types of LFG control alternatives. Passive controls typically consist of free venting structures or cut off barrier trenches that do not require mechanical components or much upkeep. To be effective, trenches must be excavated to the depth of the landfill. Vent

trenches, backfilled with gravel, serve to intercept laterally migrating LFG, and thus provide a path of least resistance to vent to the atmosphere. If covering of the gravel trenches is likely to occur, pipes need to be installed through this cover to allow the intercepted LFG to vent freely. Passive control systems can slow and/or prevent migrating LFG. However, in general, when a landfill of adjacent property is developed for public use, passive control methods are used only as a backup for an active LFG control system.

An active system consists of a series of vertical wells or trench wells with perforated pipe installed. Depending upon the site, wells are located either inside or outside of the actual landfill limits with well laterals connecting to a main header system. (See Figure 3.) A motor/blower unit or compressor provides an applied suction which draws the LFG to a central point where the gas is discharged to the atmosphere, burned in a flare to control odors, or further processed for utilization. Typically, an active LFG control system is installed in landfills where:

1. Refuse depth is greater than 25 feet.
2. There is little or no distance between the limits of deposited refuse and the property line.
3. Structures adjacent to the property line need to be protected.

Monitoring wells are installed along the property line or between the wells and the area requiring protection to verify system effectiveness. When any piping system is installed in refuse, settlement is a design and maintenance problem. Well and header piping must have sufficient flexibility to accommodate the differential settlement common to landfills. Another problem is condensate, which is produced as the warm saturated LFG cools in the header pipe. If these pipes are not correctly sloped and if condensate drains are not properly designed, the header lines may become blocked with condensate, making the system inoperable. Monitoring and maintenance will need to be carried out for years to ensure that the system continues to function as designed.

Venting collected gases may pose odor and/or air quality problems. Therefore, a flare is often required to combust the collected gases.

The cost for a perimeter active LFG control system will be dependent on the depth of refuse and flaring requirements. Installed costs for such systems can range from \$75 to \$125 per linear foot of perimeter requiring migration control.

In some landfills, existing migration control wells have been modified to operate as recovery wells. At other sites,

recovery systems are operated in conjunction with the control system. There is a basic incompatibility between wells designed for control and those designed for recovery, since control wells often draw in large quantities of air, which is toxic to the methane-producing bacteria. Not only does this reduce potential recovery rates, but it also contributes to a gas of lower quality, requiring costly upgrading to achieve acceptable Btu values for commercial use. If the control and recovery systems are to operate simultaneously, each will preferably have its own collection header. High quality gas from the specially designed or modified recovery wells will be transmitted to the processing facility and from there to the end user. Air-rich gas from the control system will be collected and flared or put to such onsite uses as fueling the motor blower or compressor.

LFG RECOVERY SITE CRITERIA

Critical to the success of a LFG recovery project is the quantity, quality and collectibility of LFG plus its marketability. Until LFG is actually sold, it is merely a waste product representing a liability to the site owner. Market use surveys should be completed as an initial task for any recovery project assessment.

LFG recovery systems can be installed at either an operating or a closed site. Local decisionmakers frequently express concern regarding compatibility of a recovery installation with final land use plans. At several sites in the United States, what once was an ugly waste disposal area has been turned into an attractive recreational area, underlain by a LFG recovery system.

Many landfills are not suitable for a commercial LFG utilization project. The industry uses the following criteria for initial site screening:

1. At least one million tons of in-place refuse.
2. Nearby and willing market users for LFG.
3. A receipt rate during the site's operational life of roughly 400 tons per day.
4. An average refuse depth of approximately 40 feet.
5. An active fill area of approximately 40 acres.
6. Recently closed or preferably with active filling life remaining.

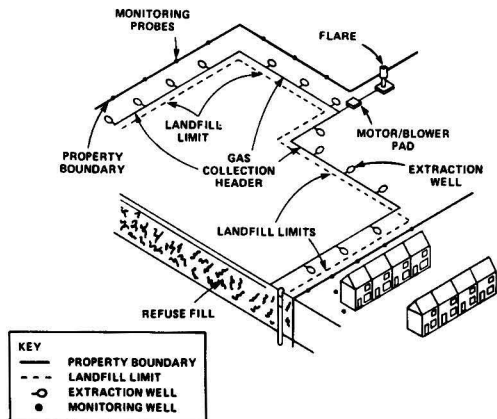


Figure 3. Active LFG control system.

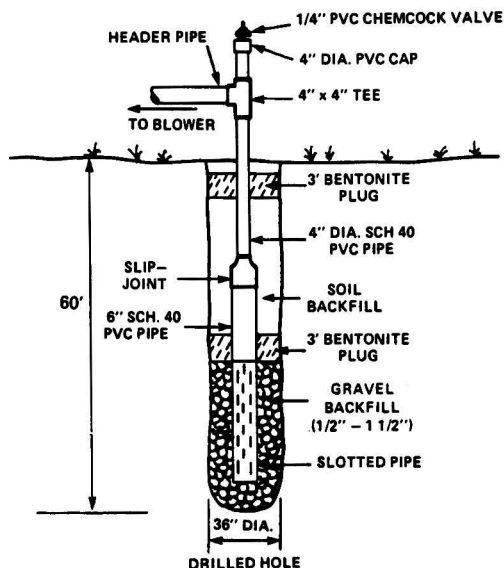


Figure 4. Typical LFG extraction well installation detail.

If the preliminary assessment looks promising, a more accurate analysis can be made from a field pumping test program to both confirm initial projections and provide design criteria for the collection/processing system.

A LFG recovery system typically consists of a series of induced exhaust wells that are designed, constructed and operated in such a manner as to minimize atmospheric air intrusion. These wells are constructed of perforated pipe casing placed in holes drilled in the refuse; the holes, which extend 50-90% of the refuse depth, are then back-filled with gravel and sealed from the surface with bentonite clay or concrete to prevent air intrusion. (See Figure 4.) A suction is applied to each well casing. The LFG withdrawn at each well is conveyed to a central point by means of a pipe network referred to as the "gas collection header." Each wellhead is normally equipped with a butterfly valve for flow rate control. The source of the applied suction and the central point to which the LFG is collected is either a vacuum blower or a compressor.

UTILIZATION OPTIONS

LFG can be used directly as a medium-Btu fuel or upgraded through processing to a higher heating value. There are three primary categories of its use:

1. Direct use as a boiler fuel after partial processing (water and particulate removal).
2. Direct combustion after partial processing to generate electricity, either for sale to the local utility or for onsite use.
3. Upgrading to pipeline standards for injection into utility company pipelines. Processed LFG can also be compressed and used to fuel vehicles or garbage trucks with converted carburetor systems.

The option of using LFG in boilers for space heating, steam generation, or hot water heating is the simplest process. Condensate and particulate removal is the only processing required. LFG direct from the landfill has an energy value ranging from 450 to 550 Btu per standard cubic foot. Several operating recovery projects currently supply LFG to chemical plants, power plants, a hotel complex all of which have a 24-hour 365-day fuel demand.

The production of pipeline quality gas from LFG requires the removal of carbon dioxide, non-methane hydrocarbons, hydrogen sulfide and water vapor. There are a number of "off the shelf" gas processing technologies in use in the LFG industry to achieve pipeline quality gas.

Electricity can be generated by either an internal combustion gas engine or a gas turbine driven generator using the medium-Btu LFG. The alternative to sell electricity to the local utility became more attractive with the passage of federal legislation in 1978 requiring utilities to purchase electricity produced from alternate energy resources at the utilities' avoided cost. However, unless the market price for the LFG-generated electricity is at least 5-6 cents per kilowatt hour, justifying the economics for generating electricity on LFG is difficult.

ECONOMICS

The relative economics of a given landfill gas project depends upon many variables. Important criteria for evaluation are total capital cost, operating and maintenance (O&M) expense, and the energy efficiency of the selected process. The total capital costs of each of the LFG energy options vary according to the location, size, or LFG volume expected, and the ultimate end use of the LFG (i.e., as medium-Btu or high-Btu gas sales or electrical power generation).

The medium-Btu plant processing scheme, in general,

has the most favorable economic projections. The medium-Btu plant typically has the lowest capital cost, lowest O&M costs, the highest energy efficiency and the highest methane recovery factor. However, several conditions affect the economics of such plants including gas sales price, location of the energy user(s) (high or low pipeline costs) and the end user's energy requirements. The medium-Btu plant will generally show the highest rate of return on a discounted cash flow basis and the highest present value for any of the LFG utilization options.

The high-Btu plant is more difficult to evaluate because of the various process schemes available, the overall methane recovery factor differences among the processes, the various energy demands of each process, and consequently, the differences related to capital and O&M costs. Total capital costs for high-Btu plants are higher than for the medium-Btu plant; the main difference is that the associated pipeline costs for the high-Btu facility are not included in project evaluations. The pipeline costs for the high-Btu plant are assumed to be borne entirely by the buyer, while for the medium-Btu plant these costs are the responsibility of the plant operator.

The electricity generation facility is the most expensive and energy-intensive of the LFG utilization options. The overall net thermal efficiency, for instance, is approximately 20-30%, compared to approximately 80% for a high-Btu plant and over 90% for a medium-Btu facility. This option will generally produce the lowest rate of return of any LFG option, except in areas of the country where the electrical utility companies' avoided costs exceed 6 cents per kilowatt-hour.

"WHO'S WHO" IN THE INDUSTRY?

As the LFG industry has grown, so have the support services. Increased state and federal regulatory activity is forcing landfill owners to manage LFG. With a number of management options available, the marketplace can appear somewhat confusing. For example:

1. There are now more than 14 development companies competing for LFG rights. These companies will test the landfill, install the facilities, collect and process the gas, contract to sell the energy product to a user, and provide the site owner with a portion of the revenues. Many of these developers seem to have become "locked in" or affiliated with a particular process.
2. At least eight major gas treatment or processing options are in use today. Proponents of each option usually claim superiority over the others. Landfill owners should be aware that no one process or energy product is automatically best in all situations. The market at a particular site *must* dictate the process chosen.
3. At least eight major engineering and construction firms have been involved in the construction of various LFG facilities. Each offers different capabilities, costs and timetables, all of which should be considered when selecting contractors.
4. A number of brands of large equipment such as compressors, blowers, generators, process skids, etc., are in use at LFG projects today.
5. Increased competition between developers and consultants has brought forth sophisticated marketing and sales techniques, making it difficult to know what information is objective. More than a dozen consulting firms today claim significant expertise in this business and some have ties to developers, suppliers, or process owners.

REGULATORY CLIMATE

Good News

The benefits of LFG recovery have been recognized nationally and by certain states. In the 1985 House version of the Superfund Reauthorization Bill, there is an amendment which specifically relieves LFG recovery operators from potential strict liability for Superfund clean-ups. The proposed bill states: "The purpose of this amendment is to encourage persons to recover and process methane from landfills in recognition of the substantial energy and public health benefits that result from these activities."

In California, the state legislature adopted the following position: "It is the policy of the state to encourage the recovery and utilization of methane gas produced during the natural decomposition of waste in landfills in California and to encourage private sector participation in the effort."

In recognition of the pollution control benefits of LFG recovery, Illinois amended its Environmental Protection Act to provide that "air pollution control equipment means any facility intended to eliminate, prevent, reduce, or control the emission of specified air contaminants and this includes LFG recovery facilities." A second Illinois law requires gas utilities to wheel processed LFG.

Bad News

While there is a lot of support for the development of LFG recovery facilities because of the recognized environmental and energy benefits, there are also a number of regulatory issues which can seriously interfere with the development of these projects. Although many local government officials have not yet had to grapple with permitting LFG recovery projects, in time these issues may have to be faced.

Three regulatory issues stand out as the "hot" topics for 1986:

1. Permitting.
2. LFG condensate handling.
3. Control of emissions from landfills.

Permitting

Multiple agencies in the same state often have overlapping jurisdiction over landfill and resource recovery operations. Different agencies are often responsible for air, water, solid waste and health, each of which impacts LFG recovery operations. Often a project must be cleared by all of these agencies before work can commence.

One Southern California project required over two years to permit, in part because of delays that included months of special testing to experiment with novel liquid control measures for landfills. This delay in permitting then extended through a turnover of the County Engineering staff that led to renegotiation of the basic agreement for the plant site. The combination of multiple regulatory layers and excessive delays resulted in a lost revenue stream in the tens of thousands of dollars per month. During this delay, the citizens around the landfill were subjected to as much as 100 tons per day of hydrocarbon emissions that would otherwise have been removed by the plant.

As another example, in Illinois there was an overlap in jurisdiction between the state control board and the state environmental protection agency. After meetings and briefings with the LFG industry, the two agencies finally held a joint hearing and designated a single office to assume overall responsibility for gas recovery projects.

Basically, a relatively straightforward process of permit consolidation should be considered for LFG facilities so that development of these facilities can proceed on time and on budget. A lead agency could be designated. Otherwise, with multiple permitting layers, developing plans can often be extremely costly and time consuming.

Condensate

Raw LFG is 100% saturated with water when it is recovered from the landfill. Vapor condenses from LFG during the gas processing stages. Traditionally, the condensate produced has been recirculated back to the landfill on the basis that it originates in the landfill and is merely a by-product of the gas that is recovered.

In addition, the condensate represents only a fraction of the total amount of liquids in any given landfill. Experts have calculated that, in any year, no more than 1/20 of 1% of the moisture in a landfill may be condensed out of the gas that is processed, or a total of 1% over a 20-year period.

At LFG facilities where condensate is not allowed to be recirculated, cost for off-site disposal can be substantial. For example, at one California LFG facility, the developer incurs costs, including transportation costs that amount to 75¢/gallon. On the average, the amounts of condensate per million cubic feet of recovered LFG range from 300 to 1,100 gallons. Thus, for facilities generating substantial amounts of condensate, handling costs can seriously impact project economics, and have, in fact, constrained/cancelled development of certain projects.

One major issue facing the solid waste industry is whether restrictions on liquid disposal in landfills should apply to condensate recirculation. Despite the fact that condensate originates in the landfill, and is not the disposal of a new liquid waste stream into the landfill, there has been confusion on whether liquid disposal prohibitions apply.

Recognizing the dilemma, the state of California adopted a regulation allowing condensate recirculation to landfills as an exemption to California's general liquid disposal restrictions. Local districts retain the right to issue waste discharge requirements.

The 1985 House version of the Superfund Reauthorization Bill would also allow condensate recirculation by providing that recirculation is deemed not to be the disposal of a liquid waste.

A further issue is how the condensate should be characterized under RCRA hazardous waste identification rules. EPA has chosen to apply what is referred to as the "characteristic rule" and "derived from rule," depending upon the type of landfill from which the gas is recovered; however, EPA has also noted that when a state has been given authority to enforce its own hazardous waste programs, the state's rules apply in lieu of the federal regulations. Accordingly, states are free to determine how condensate should be characterized and regulated under their hazardous waste rules and regulations.

Control of Emissions from Landfills

Increased concern about the surface emissions from landfills of reactive organic gases, odorous compounds, and/or toxics led to passage of California State Assembly Bill 3525 known as "The Calderon Bill" in 1984. This bill requires all solid waste disposal sites to monitor the hazardous wastes in the air adjacent to and outside their perimeters. The onus of enforcement of standards is on the local air quality management districts, who must receive these monitoring reports on or before January 1, 1987, and then require mitigative measures should air pollution be detected.

Both the South Coast Air Quality Management District (Los Angeles area) and Bay Area Air Quality Management District (San Francisco area) have developed regulations to enforce this legislation. In general, these regulations require:

1. The installation and monitoring of an active gas control system at landfill sites to prevent concentrations of organic compounds on the landfill surface from exceeding 50 ppm.
2. The installation and monitoring of perimeter gas sampling probes.
3. Determinations of concentrations of both the trace toxic air contaminants and total organic gases emitted from the landfill surface, both above the site and at the perimeter.

Although California is a leader in this type of legislation, other states with dense population centers can be expected to adopt some type of landfill emission regulations. The impact of such regulations will be considerable for developers and operators of landfill gas recovery facilities. For instance:

1. There may be conflicts at sites with high Btu plants between meeting required surface emission levels and maintaining Btu levels.
2. There will be a need to increase system reliability. Owners of landfills will have to submit system monitoring reports. If migration or emission control systems are not operational, the owners will be in non-compliance.
2. Both gas migration control systems and gas extraction systems may need to be installed at the same rates to achieve compliance with all sections of these regulations.
4. Owners of certain landfills in California have begun requiring system operators to assist in, or be responsible for, meeting the air quality management district regulations.

CONCLUSION

The management of LFG is a growing concern nationwide. LFG recovery is one option that should be evaluated by landfill owners as a means to meet odor control, closure regulations and/or to help offset long-term closure costs. Key advantages resulting from a LFG recovery project are as follows:

1. *Reduced hazards:* By extracting LFG in a controlled way, the hazard represented is substantially reduced.
2. *Reduced emissions and odors:* LFG recovery/control provides a means for reducing the level of pollutants that are released to the atmosphere and thus air quality improves.
3. *Economic benefits to the local community:* LFG projects result in the payment of taxes and royalties to the owners of the landfill (often local governments). In addition, construction and operation of a facility generates business activity and can create new jobs in a community.
4. *Conservation of non-renewable fuels:* When LFG displaces conventional fossil fuels, these non-renewable fuels are saved for other uses.
5. *Improved vegetation:* The recovery of LFG results in improved soil conditions for vegetation both on and surrounding the landfill.

Decisions on LFG recovery feasibility will always be site-specific. However, in those cases where LFG recovery seems possible, the option to turn this potential liability into a viable asset should be pursued.



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Combustion of Refuse-derived Fuel in Utility Boilers

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INTRODUCTION

Nearly all of the solid waste generated in the United States today is being landfilled. In many communities this method of disposal is no longer acceptable either economically or environmentally. Resource recovery, or the recovery of energy and materials from solid waste, has recently become an attractive alternative to the continued practice of landfilling. By using solid waste as an alternative fuel source, electric utilities are able to recover energy, conserve fossil fuels and reduce the volume of material that must be landfilled.

The purpose of this paper is to review how an electric utility might best use its existing facilities to burn solid waste and to identify some of the technical risks and concerns associated with the use of solid waste as a fuel.

First, the use of refuse as a fuel is reviewed, and then two methods of firing refuse in existing facilities are presented. These methods include co-firing with conventional fossil fuels and converting existing boilers to burn 100% refuse-derived fuel (RDF).

REFUSE AS A FUEL

Early efforts by utilities to use solid waste as a fuel have demonstrated the importance of improving its combustion properties. These early projects attempted to minimize the undesirable effects of refuse combustion by mixing the refuse with more predictable fuels such as coal. Unfortunately, the small amounts of refuse fired introduced significant amounts of low-fusion-temperature ash materials, which increased problems with slagging and tube fouling. There were also problems with the heterogeneous nature of refuse. The occasional introduction of wire stands, rope, rags and streamers caused material hang-ups and bridging problems, and the presence of metals, glass and other abrasive substances eroded conveying lines and boiler tubes. The typically high chloride content of refuse created problems with gas side corrosion and plant emissions.

In an attempt to minimize these problems, process facilities have implemented a number of improvements over the last 10-15 years of operation. The most important is the use of equipment and systems designed specifically for processing solid waste.

The first attempts at processing refuse used equipment from the mining and ore processing industries. This approach was found to be generally unsuitable for processing a heterogeneous material such as solid waste. As a result, equipment and processing systems have since been

developed that can separate and refine the combustible fraction more effectively. Through an improved process using multiple stages of screening and air classification, selection shredding and magnetic separation, a higher yield of refuse-derived fuel product can be produced with lower ash content.

It is important to recognize that while significant improvements have been made in processing refuse, the RDF product is still an unpredictable material with a low heating value, usually containing high amounts of ash and moisture. However, by improving and stabilizing the RDF properties and by tailoring the refining process to meet specific combustion requirements, existing equipment can be adapted to burn RDF while maintaining good availability.

CO-FIRING RDF IN EXISTING BOILERS

Nearly all of the utility experience in burning refuse has been co-firing RDF with coal in existing suspension-fired boilers. This experience is tabulated in Table 1. As shown, the RDF has been co-fired in several types of boiler designs. These include wall-fired and tangential-fired pulverized coal units, and cyclone-fired units. The operating success of these projects has varied. Some installations have been successful while others have been terminated as a relatively negative experience. Many of the combustion problems encountered in these early projects have since been remedied, however, by improved processing technology and by modifications to the combustion equipment.

Some of the major concerns in co-firing RDF in a modern suspension-fired boiler are as follows:

Fuel Feed

The small particle size required for suspension firing will normally allow the RDF to be conveyed and injected directly into the furnace pneumatically. Particle sizes with a top size of one inch are commonly used. Because of the abrasive nature of RDF, conveying lines are subject to high wear rates and should be provided with wear back elbows. Also, 90-degree bends and long verticle runs should be avoided to minimize pluggage.

Combustion

In suspension firing, finely sized fuel particles are pneumatically injected into the furnace, where rapid combustion takes place at very high temperatures. Be-

TABLE I. PRINCIPAL RDF CO-FIRING INSTALLATIONS USING UTILITY BOILERS

Utility	Unit No.	Boiler Size 1000 PPH	Type Combustion	Percent RDF	Status/Comments
Union Electric Company Meramec Plant, St. Louis, MO	1	925	Tangential	5-20	First demonstration 10/74 through 11/75.
Ames Municipal Electric Co. Ames, IO	7 8	360 620	Tangential Wall-Fired	20 25	Operational since 9/75. Slagging and wear reduced by adding disc screens.
United Illuminating Co. Harbor Station, Bridgeport, CT	1	625	Cyclone	30	Shutdown 11/80 due to financial problems. Only 1 year operation.
Commonwealth Edison Crawford Generating Station Chicago, IL	7 8	1300 2000	Tangential Tangential	10 10	Shutdown due to process plant difficulties.
Madison Gas & Electric Blount Street Station Madison, WI	8 9	425 425	Wall-Fired Wall-Fired	12 12	Operational since 1/79. Clinker formation reduced by lowering ash/smaller particle size.
Wisconsin Electric Power Co. Oak Creek Station Milwaukee, WI	7 8	2000 2000	Tangential Tangential	10 10	Shutdown since 1980. Slagging and ash removal problems.
Rochester Gas & Electric Co. Russel Station Rochester, NY	1 2 3 4	350 430 430 560	Tangential Tangential Tangential Tangential	15 15 15 15	Shutdown since 7/84.
Baltimore Gas & Electric Co. C. P. Crane Station Baltimore, MD	2	1360	Cyclone	15	Operational since 2/84. Clinkering problems. General satisfactory operation.
City of Lakeland Dept. of Electric & Water Utilities C. D. McIntosh Jr. Power Plant Lakeland, FL	3	2510	Wall-Fired	10	Operational since 6/82.

cause of the high heat release rates, heat absorption and boiler performance are sensitive to changes in fuel characteristics; that is, sudden changes in moisture or particle size. The RDF must therefore be of comparatively high quality, uniform in size and consistent in ash and moisture content to maintain flame stability and avoid upset conditions.

The ash content of RDF on a Btu basis is much higher than it is for coal; and as a result, significant quantities of low-fusion-temperature ash can be injected. Because of the tight furnace design used in modern suspension-fired units, significant slagging and boiler tube fouling problems can result. The remedy to this problem is to minimize the ash content of the RDF.

The shredded RDF must be injected at a point in the furnace where maximum burnout of the RDF can be achieved. If the material is injected too high in the furnace, the residence time for the lighter particles may be insufficient for complete burn out. This will allow a carry-over of burning material into the conversion section of the boiler and cause overheating, increased deposition and increased flyash emissions. Heavier material, however, will have more time for burnout, since it falls by gravity to the lower furnace. Injecting the RDF at a lower elevation may reverse the problem where insufficient mixing with the coal-air mixture causes a substantial amount of unburned, dense RDF to drop out of the furnace.

The optimum point of injection would appear to be in the area of greatest turbulence and exposure to the burning coal-air mixture, probably near the center of the coal burners.

The addition of dump grates in the lower furnace is required to capture falling debris and allow additional time for complete burnout. This improves combustion efficiency and reduces the presence of floaters or unburned refuse which can clog wet bottom ash removal systems. The ash buildup on the grate must be monitored so that

collected ash can be dumped into the furnace bottom ash hopper before significant clinkers form. Also the use of underfire air is desirable to assist in preventing clinker formation and improve burnout of the refuse.

RDF is highly volatile, and as a result it will ignite earlier than coal. This may improve coal combustion and achieve more complete burnout within the flame envelope.

Co-firing RDF in a slagging furnace such as the cyclone boilers used by Baltimore Gas & Electric at their C. P. Crane Station may be a more effective method of incinerating refuse. A cyclone furnace is a water cooled cylinder wherein coal is fired at extremely high heat release rates and temperatures. Coal and combustion air are admitted tangentially to impart a centrifugal action to the coal particles. High combustion temperatures exceeding 3000°F cause the ash to melt into a molten slag that lines the cyclone walls. The injected fuel adheres to the molten slag and burns to completion. The molten ash is removed through a slag tap.

The cyclone furnace can burn a wider variety of fuels compared with a suspension-fired unit. Also, since RDF ash typically exhibits lower fusion temperatures than coal ash, it may be beneficial in developing a low viscosity slag. The tangential injection of the RDF into the cyclone must achieve complete mixing with the slag liner while minimizing the escape of lighter particles through the center of the vortex. If the RDF can be effectively incinerated in the slag liner, and the ash can be removed as molten slag, the use of a cyclone furnace may be less sensitive to the variability of RDF and result in less slagging and fouling.

The results of co-firing at the C. P. Crane Station have been favorable to date. Clinkering problems have apparently been corrected by adjustments to combustion air control, and recent inspections have indicated no abnormal corrosion or erosion [5].

TABLE 2. PRINCIPAL FACILITIES BURNING 100% RDF IN TRAVELING GRATE SPREADER STOKER BOILERS

Location	No. Boilers	Capacity (Each) 1000 PPH*	Project Status
Akron, OH	3	126	Temporarily closed while re-pairing damage from explosion in 12/84.
Albany, NY	2	100	Operating
Columbus, OH	6	165	Operating
Dade County, FL	4	200	Operating
Hempstead, NY	2	200	Shut down due to environmental problems.
Lawrence, MA	1	250	Started operation 9/84.
Niagara Falls, NY	2	300	Operating
Portsmouth, VA	4	200	Under construction. Operation scheduled for spring 1987.
Naval Shipyard			

* PPH = Pound per hour

Performance

Boiler performance is not significantly affected when 10-20% RDF (on a Btu basis) is co-fired with coal in suspension. Losses associated with increased excess air, higher moisture and additional unburned fuel may reduce boiler efficiency by 2-3%. The limit on the amount of RDF fired will most likely be determined by other concerns, such as slagging, fouling erosion or corrosion.

The decision to burn RDF as a supplemental fuel in a modern unit must be evaluated on a case-by-case basis. It is very likely, however, that units firing a marginally acceptable coal will experience the greatest combustion problems, unit derates and overall loss in performance.

100 PERCENT RDF FIRING IN EXISTING BOILERS

Converting a suspension-fired unit to burn 100% RDF is generally infeasible. The conversion of such a unit to dedicated RDF combustion would require that the lower furnace be removed and replaced with a traveling grate. This would be expensive and would cause a significant derate in boiler output because of the tight furnace design and the resurfacing required to reduce steam temperatures. In general, the modern, higher pressure and temperature, suspension-fired units are better suited to co-firing RDF.

A more practical approach to dedicated RDF combustion would be the conversion of existing boilers originally designed to burn coal on a traveling grate stoker. Stoker-fired boilers are capable of burning a wide variety of fuels, and as shown in Table 2 they are being used in a number of installations to burn 100% RDF. Although these boilers were specifically designed for dedicated RDF combustion, they are very similar to the spreader stokers that were heavily marketed to electric utilities in the 1940's and 1950's. If these units are still available, they may be well suited for conversion to dedicated RDF combustion since they exhibit a number of favorable characteristics.

1. Most units were built with steaming rates in the 100,000 to 300,000 lb/hr range, which is well matched to the production capacities commonly used in RDF processing facilities.

2. They are generally sited close to electrical load centers, which are usually near the waste-generation centers.
3. The design steam temperatures and pressures are comparatively low, offering protection against high-temperature chloride corrosion problems.
4. The conservative design of boilers of this vintage offer maximum generating capabilities which may exceed guaranteed or rated output with substantial margin. In addition, the erosion and corrosion that may have occurred in the pressure parts may not constitute a serious reduction in useful life, since liberal materials thickness allowances were typically used in the original design.
5. Most of these units are now operated at a low capacity factor for standby peaking service only. Conversion to base loaded refuse firing may offer more productive use of existing resources.

These characteristics may qualify an existing stoker-fired boiler for conversion to dedicated RDF firing; however, a number of considerations must be evaluated further to determine whether conversion is truly cost effective. Some of the major areas of concern are as follows:

Fuel Feed

The special properties of RDF and the special equipment required to convey it and meter it into the furnace make it impractical to use existing coal storage silos and feed equipment. Most of this equipment would likely require removal to make room for the new RDF feed system.

The larger particle sizes used for grate-fired RDF create some special material handling problems. Automated reclaim systems using augers and screws will be susceptible to wrapping and pluggage due to the increased amount of streamers and rags. Also, belt conveyors should be used in lieu of pneumatic conveying systems.

The boiler feed system must be able to spread a controlled amount of RDF uniformly over the grate. This is generally accomplished by volumetrically metering the RDF from small surge bins into air-swept spouts. The air supply is usually pulsed in order to achieve uniform distribution from the front to the rear of the furnace.

Several methods of feeding the air-swept spouts have been used with varying degrees of success. The ease with which RDF compacts and agglomerates makes it difficult to meter consistently and uniformly. Past experience has demonstrated a need for a feed system that keeps the material moving in constant agitation and does not require lengthy storage times in deep piles. Overrunning systems that top off shallow live bottom bins and swinging feed spouts have worked well.

Grate Systems

Successful burning of RDF on existing grates will depend on the type of grate used and its condition. The design and operation of underfeed stoker grate systems are unsuitable for RDF combustion because of pluggage and compaction problems. These types of grate systems will need to be replaced. The traveling grate, with front ash discharge commonly used in a spreader stoker, appears to be the best candidate for burning RDF. The spreader stoker is very flexible and can burn fuels with a wide range of burning characteristics. There are many successful installations of spreader stokers burning refuse, wood waste, bagasse and other varied materials.

The existing grate size may be a limiting factor. The distance over which the RDF can be effectively spread

with an air-swept spout is limited to about 18 feet. In addition, the amount of RDF fired will be limited by the allowable grate release rate, usually no more than 750,000 Btu/hr-ft².

Furnace Requirements

A liberally sized furnace is required for the combustion of RDF. The furnace contains the combustion process and exposes the products of combustion to the water wall surface which cools the combustion gas to acceptable temperatures before passage into the convection section. The furnace must provide the necessary residence time for complete burnout of RDF particles and lower the gas temperature leaving the furnace to at least 1600°F. This corresponds to a volumetric heat release rate of less than 20,000 Btu/hr-ft³. Although many of the older stoker units were liberally designed, it is likely that conversion to RDF will result in some derate in boiler output. This might be remedied, however if the grate can be lowered and additional water wall surface extended to the lower grate elevation.

Combustion Air

Existing fan capacities and head requirements must be reviewed to determine if they can meet the higher excess air requirements and air distribution needs of RDF firing. Approximately 50-60% excess air will be required to incinerate the RDF. Additional under-fire air may be needed to increase velocity through the air passages and more vigorously agitate the burning solids on the grate. This will expose unburned materials more effectively.

Combustion of RDF on a spreader stoker results in a large part of the fuel burning in suspension above the grate. This type of semi-suspension firing extends combustion higher up in the furnace and can cause excessive furnace exit gas temperatures. Overfire air must, therefore, be injected above the bed in order to completely incinerate the rising particles and cool the gases before they leave the furnace.

The intertube gas velocity will be limited to 30 to 40 fps to minimize erosion. The combination of higher excess air requirements and lower allowable gas velocities may become the limiting constraint on boiler output.

Corrosion Protection

Refuse fuels exhibit high concentrations of chlorine, introduced from chlorinated plastics and from the salts of organic materials. Combustion of these materials forms hydrogen chloride which is very aggressive and can greatly increase corrosion rates in the boiler water walls and superheater tubes.

The lower furnace tube metal can be protected by the addition of refractory to a height extending well into a predominantly oxidizing zone. The superheater tubes can be protected by limiting steam temperatures to about 750°F. Below this temperature, the exposed metals are not severely affected. If the existing unit is designed to generate higher temperature steam, tube surface must be removed or rearranged. If higher temperatures must be maintained, the superheater section should be replaced with corrosion resistant materials.

Refractory coatings on tubes has not yet been demonstrated to be a cost effective and reliable method of protecting high temperature tubes.

Air Heaters

It is generally desirable to add an air heater if one does not already exist. Preheating the combustion air to about

300°F or 350°F helps to dry and ignite the RDF and also improves boiler efficiency. The absorption of heat from the gas stream may be limited, however, by gas side dew point corrosion temperatures. Combustion of refuse fuel creates potentially higher acid dew point temperatures and consequently requires that flue gas temperatures be kept relatively high, 350°F or higher. If this cannot be achieved, it may be necessary to remove economizer sections or preheat the air with separate steam coil air heater.

If an air heater is added, a tubular type of air heater is generally desired. The close passages of the regenerative heat exchanger furnace are susceptible to pluggage and, because of air leakage, can be a fire hazard. With either design, tubular or regenerative, air preheating with a steam coil air heater keeps metal temperatures above acid dew points and minimize cold end corrosion.

CONCLUSION

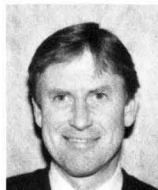
Utility experience in burning refuse as a supplementary fuel in modern fossil fueled boilers has not been viewed as a total success. There can be serious problems when existing combustion equipment designed for a specific fuel is used to burn less than design quality fuels. Recent improvements, however, in the processing technologies and a better understanding of required boiler modifications has minimized material handling and combustion problems associated with co-firing RDF.

Another approach is the conversion of existing boilers to burn 100% RDF. If available, traveling grate spreader stoker boilers originally designed to burn coal may be good candidates for conversion to dedicated RDF combustion.

Participation of electric utilities in the resource recovery industry offers benefits to both the utility and the local community. Benefits to the utility include the productive use of existing resources and the ability to retain customers that might otherwise be lost to competing energy producers. The community benefits by reducing the volume of waste that must be landfilled.

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Advanced IGCC Power Systems for the United States

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Technologies are being developed to increase efficiency, and reduce capital costs of IGCC systems by simplifying them. These include hot gas sulfur and particulate cleanup, desulfurization in the gasifier and improved performance of the power island.

INTRODUCTION

Electric utilities are businesses in transition. This transition comes from: an end to a period of rapid expansion; collapse of the former strategy for capacity additions; the emergence of acid rain as a national issue; a reversal of trends on fuel availability; and a change in Government policies including utility deregulation and mandated supply options such as the Public Utilities Regulatory Policy Act and the Fuel Use Act.

Market and technology events combined to start this transition. During the 1960's and early 1970's, electric utilities enjoyed a period of rapid expansion in pace with a rapidly growing economy. Improved generating technology fueled the expansion and provided low cost power. But the oil crisis in the early 1970's, the later shortage in gas deliverability, delays in nuclear plants, and the collapse of demand under slow economic growth completely changed the picture. Suddenly, electric utilities had uncompleted plants whose costs were escalating continuously. Loss of fuel flexibility followed as oil, gas and nuclear options faded due to cost, availability or public concerns. Utilities that had planned large capacity expansions could not complete and finance them. Coal came under attack as the cause of "acid rain," a controversial and technically unresolved issue. Utilities began considering generation expansion strategies that were unthinkable a few years ago: using oil and gas; abandoning economies of scale by planning small, incremental capacity additions to better match demand; and looking to new technologies that could use coal with superior environmental performance to handle possible tighter regulations in the future.

Two promising new technologies for power generation are fluidized-bed combustion and Integrated Gasification Combined-Cycle (IGCC) systems. Atmospheric fluidized-bed combustion for utility power generation is being pursued at the demonstration phase by the Tennessee Valley Authority. Pressurized fluidized-bed combustion is at an earlier stage of development. IGCC systems look especially attractive due to their potential for future improvements. But they carry more risks than the fluidized-bed combustion systems. IGCC systems offer large benefits in dealing with the acid rain issue through their use for repowering of existing plants while adding new capac-

ity [1]. This paper explores some of the promising developments that should make advanced IGCC systems a strong choice for utility generating plants in the mid-1990's and beyond.

THE IGCC SYSTEM

The key elements or subsystems that make up an IGCC system are the gasifier, heat exchanger, gas cleanup, and power generation block. These subsystems are integrated to carefully manage thermal energy. Successful integration minimizes both waste heat and the use of the salable product (electricity) within the plant. Such choices can be trade-offs against plant complexity and cost.

The gasifier subsystem converts coal into a combustible gas by reacting it with steam and oxygen or air at temperatures up to 3,000°F (1,920 K) and pressures from near-atmospheric to over 80 atmospheres (8,100 kPa). There are three main types of gasifiers: entrained-, fluidized-, and fixed-bed reactors, which respectively deliver raw gas at temperatures of about 2,300°, 1,800°, and 1,000°F (1,533, 1,255, and 810 K). The entrained-bed gasifiers provide hydrogen and carbon monoxide as the major combustible products. The fixed-bed gasifier delivers these gases plus some methane, small amounts of higher hydrocarbons and combustible coal tars. Fixed-bed gasifiers deliver the highest chemical energy per pound of coal input. Fluidized-bed gasifiers are a unique compromise of the two other types. Reference [2] is an excellent source for the basic gasifier systems and their applications.

The power subsystem lies at the other end of the IGCC system. Combined-cycle systems include a combustion gas turbine to burn the fuel and expand the hot, high-pressure gases to generate electricity. The thermal energy remaining in the combustion exhaust is recovered in a boiler to make steam for a separate steam turbine to generate additional electrical power.

The gas turbine cannot use the raw gas from a gasifier directly due to contaminants from the coal. Critical contaminants in the fuel gas are sulfur compounds, nitrogen compounds, particulates and alkali. The last two contaminants can shorten turbine life. Fluid-bed and entrained-bed gasifiers provide fuels low in organic nitrogen com-

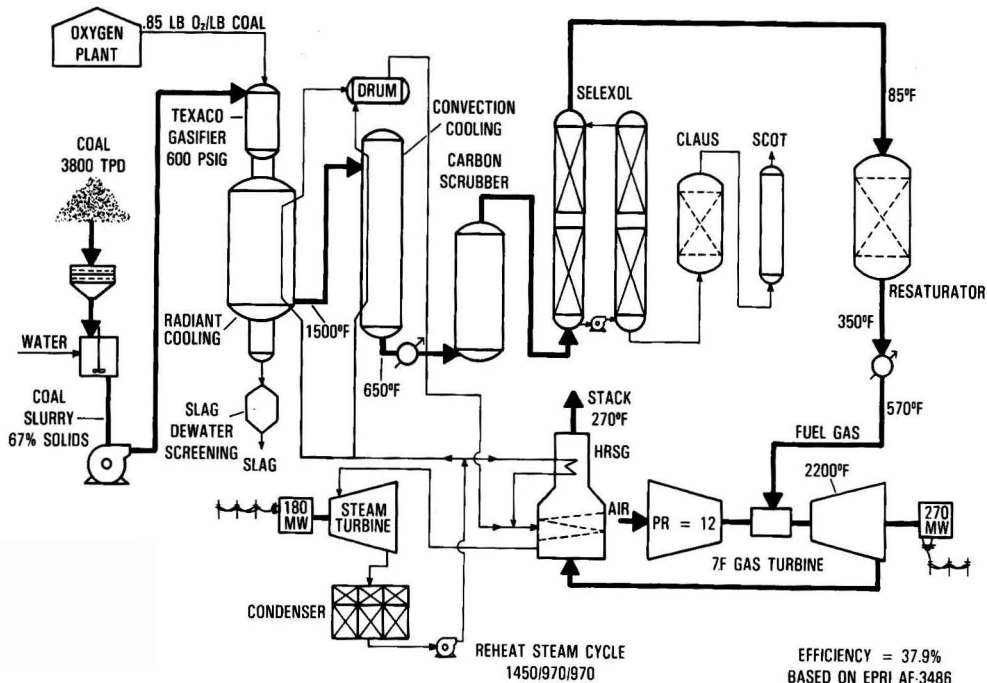


Figure 1. Process diagram for a commercial "cool water" power plant.

pounds, and the turbine combustor can be designed to minimize NO_x emissions. Sulfur compounds, particulates and alkali must, however, be removed by other equipment. Sulfur and nitrogen oxide and particulate emissions must at least meet New Source Performance Standards for coal-fired utility plants. Any tars in the fuel gas would remain in the vapor state and burn in the combustor if the fuel gas is at $1,000^\circ\text{F}$ (810 K), but there is no extended experience with the combustor fuel control valve using this fuel. Combustor inlet fuel temperatures of $1,000^\circ\text{F}$ (810 K) are state-of-the-art, and $1,200^\circ$ to $1,400^\circ\text{F}$ (920 to 1,035 K) seem feasible in the future. Additional fuel gas specifications include fuel gas heating values of greater than 120 Btu/scf (4,470 kJ/m³) and alkali concentrations of less than 20 ppb at the expander inlet.

The cleanup subsystem processes the raw fuel gas to a form suitable to the combustion turbine and is the key to a successful system. Commercial desulfurization processes require inlet temperatures of about 300°F (420 K). The product gas from all types of gasifiers is too hot for these "cool" systems and gas cooling is thus required. A variety of "cool" gas stream cleanup processes is available commercially. They are efficient and remove all four harmful contaminants mentioned earlier to levels well below environmental and combustion turbine limits. They also remove tars and even uncontrolled trace elements. But they add cost, are complex, and require heat exchange equipment to cool the gas for cleanup and later to raise the temperature to recover the heat. The replacement of "cool" cleanup subsystems with "hot" subsystems which do not have these drawbacks is a major aspect of advanced IGCC systems.

COOL WATER, AN IGCC BASE LINE

The Cool Water Coal Gasification Program introduced the first IGCC system to the United States and to the world in 1984 to 1985. Early reports of its performance

have been exceptionally good [3]. The technical success of the Cool Water IGCC plant has opened the door to new technologies that can be the power options for the 1990's. The system uses an oxygen-blown, Texaco entrained-bed gasifier. Radiant and convective heat exchangers cool the gas for "cool" cleanup and raise saturated steam for superheat in a heat recovery steam generator (HRSG). Particulates are scrubbed, and the gas is cooled further for sulfur removal by the Selexol process. The gas is then reheated for combustion and expansion in a 65-megawatt (MW) General Electric Frame 7F combustion gas turbine. Remaining energy is recovered in a HRSG to superheat steam for a 55-MW steam turbine. The medium-Btu gas from the gasifier has a heating value of 260 to 280 Btu/scf (9,700 to 10,400 kJ/m³) with 38% hydrogen. Environmental performance has exceeded all targets. Heat rates have met design targets of about 10,500 to 11,500 Btu/kWh (11,080 to 12,130 kJ/kWh), depending on the coal being gasified.

Figure 1 is a process schematic for a mature Cool Water type plant with many of the redundancies removed from the first-of-a-kind design. The calculated efficiency is 37.9%, giving a heat rate of 9,000 Btu/kWh (9,500 kJ/kWh). Other configurations of this basic system have been developed [4]. These include cooling the gasifier product gas by using a radiant-only cooler or substituting a quench for the radiant and convective coolers. These simplifications reduce capital requirements at the expense of increased heat rate.

MODULARITY, STAGED CAPACITY ADDITIONS AND REPOWERING

The strategies used by utilities to add generating capacity are important considerations for new technologies. These "business aspects" can sometimes far outweigh technical considerations in an economic analysis. A basic premise of this paper is that simplified IGCC systems

lend themselves exceptionally well to modular design, and modular design is the ideal approach to staging addition of new capacity and repowering.

The lesson learned during the 1970's and 1980's was that adding large increments of capacity may not be an optimum strategy despite the economy of scale of large systems. It leads to large financial exposure and risk that can be reduced if capacity more closely follows demand through phased capacity additions. A number of papers have analyzed this strategy in detail [5].

The concept of phased additions for IGCC systems is illustrated in Figure 2. The upper part of the figure depicts capacity and capital expenditures with one large, unphased conventional add-on. Both capital-at-risk and excess capacity are quite large. The lower part of the figure depicts phased capacity add-ons, in this case, two gas-fired combined-cycle modules which are eventually converted to coal-gas firing in the third phase. Capital-at-risk is much lower for phased addition to achieve the same new system capacity. Excess capacity that is not producing revenue is also much lower. This strategy can also apply to repowering an old plant, starting with the existing steam turbine.

The ability to add smaller power increments is key to phased additions: shop fabrication of small modules replicating a standard design lends itself to low cost production and minimal on-site construction, contrary to the traditional scaling law [5]. The standard factory design can also significantly reduce or eliminate customized architect-engineering designs which have cost as high as \$300/kW for conventional coal-fired plants. Coupled with costs due to long field construction, these "indirects" can dominate plant costs. Figure 3 compares the relative costs of a natural gas-fired, combined-cycle power plant that is very modular and requires "simple" field installation,

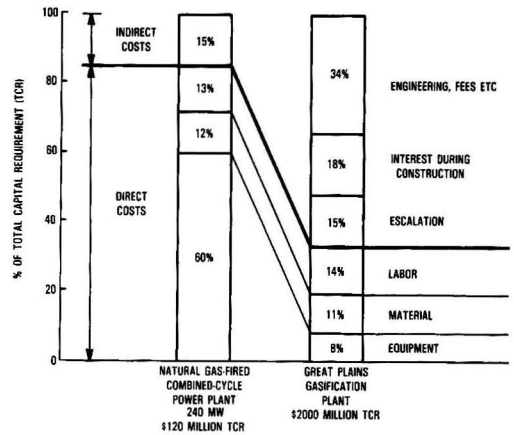


Figure 3. Capital costs distribution for modular and non-modular plants.

with the Great Plains gasification plant that is essentially custom-built. Indirect costs comprise only 15% of the gas-fired plant but comprised two-thirds of the Great Plains plant.

Technical developments are under way that can simplify IGCC systems to make the strategy described above achievable. These include development of a hot gas cleanup subsystem, in-bed desulfurization during gasification, simple tail gas handling and aircraft-derivative combustion turbines in advanced cycles. The remainder of the paper discusses these developments and shows their impact on future IGCC systems.

HOT GAS DESULFURIZATION

The development of a hot gas cleanup subsystem is key to simplification and integration of the IGCC system. The Department of Energy's Morgantown Energy Technology Center (DOE/METC) has been investigating hot gas desulfurization since the 1970's. Iron oxide was the first sorbent chosen because it sulfidized and regenerated easily. However, its equilibrium partial pressure of hydrogen sulfide was higher than desired for highly effective desulfurization, and other sorbents were sought.

Zinc ferrite ($ZnFe_2O_4$, magnetite) was found to be a most suitable replacement for iron oxide. It combines the high-sulfur affinity of zinc oxide and the easy regenerability of iron oxide, allowing desulfurization to below 10 parts per million (ppm) of hydrogen sulfide in the exit gas. Furthermore, it sulfidizes best around 1,000° to 1,200°F (810 to 920 K), a temperature range closely matching that of the product gas from fixed-bed gasifiers. The sulfidation reactions produce ferrous sulfide and zinc ferrite. Regeneration is done by roasting the spent sorbent with air, releasing sulfur dioxide and restoring the zinc ferrite. The successful integration of this type of gasifier and zinc ferrite cleanup eliminates completely the costly heat exchanger system shown in Figure 1 for "cool" cleanup plants.

The zinc ferrite sorbent has been tested in laboratory-scale reactors using simulated coal-gas mixtures and in a larger bench-scale reactor, measuring 6 inches in diameter and 6 feet in length (0.15 m by 2 m long), desulfurizing a slipstream from the METC 24-ton per day (22,000 kg/day) fixed-bed gasifier. Figure 4 shows H_2S concentrations at the inlet and exit of the zinc ferrite reactor while desulfurizing gas from the METC gasifier [6]. Sulfur levels in the product gas were typically reduced by three orders of magnitude or better for all of the high-sulfur coals tested.

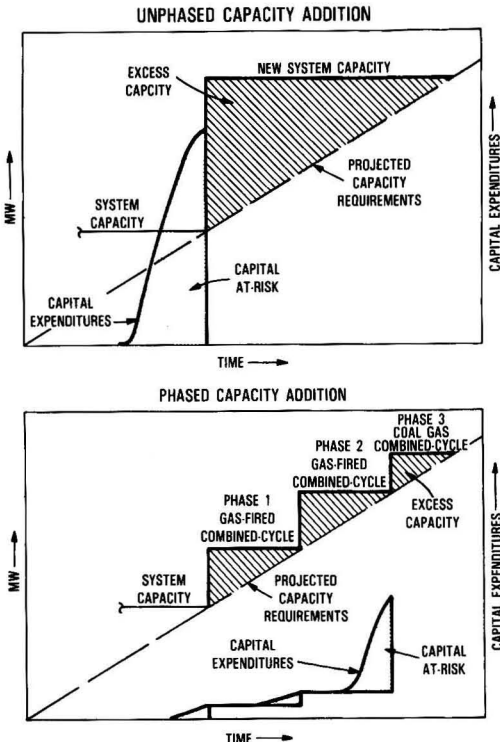


Figure 2. Economic benefits and reduced risks with phased construction.

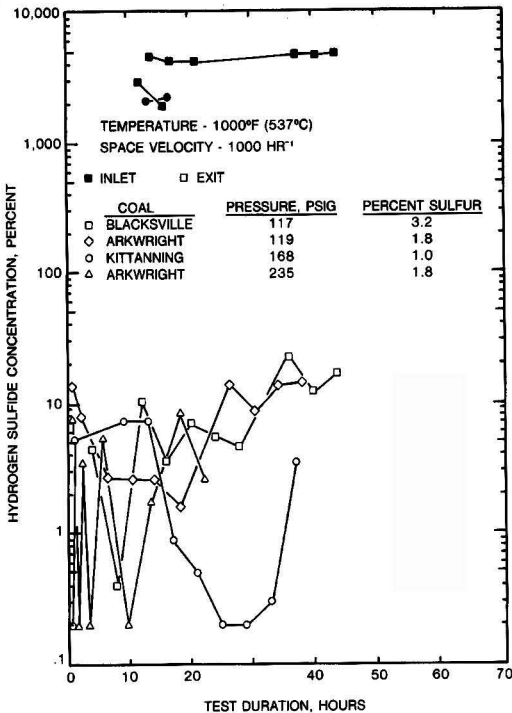


Figure 4. Desulfurization of a hot gasifier product gas by zinc ferrite.

There are many publications on zinc ferrite results [6]. The Proceedings from the Annual Gas Cleanup Contractors Meeting, sponsored by METC, provides the current status on hot gas cleanup research funded by the DOE. A recent compilation of test data shows that the zinc ferrite sulfidation and regeneration have been thoroughly examined over a range of operating conditions, such as temperature, pressure, and reactor space velocity [7].

A key performance parameter is durability of the sorbent — its ability to retain sulfur capacity and mechanical properties with repeated cycling between sulfidizing and regeneration. The theoretical sulfur capacity of the sorbent is 35 wt % sulfur. Capacities from 15 to 30% have been achieved; 25% is a commercial target but lower capacity is commercially workable. Sorbent formulations that increase mechanical strength usually do so at the price of lowering porosity and the rate of diffusion into the pellet, which results in larger reactors.

Tests of long-term sorbent chemical durability sponsored by METC have reported promising results [8]. Fig-

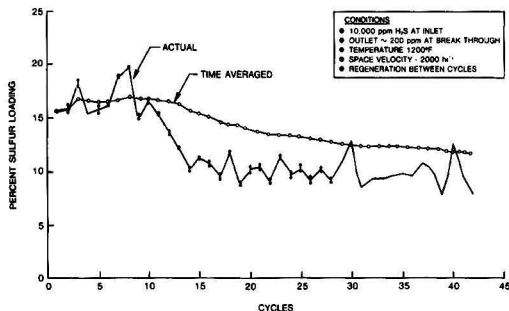


Figure 5. Durability test results for a zinc ferrite sorbent design.

ure 5 shows sulfur loading of a formulation tested over 42 cycles. Sulfur retention starts at about 17% of sorbent weight and decreases to about 10% at about 15 cycles with little further average degradation. The spherical pellets of this sorbent had double the crush strength of the original cylindrical sorbent tested by METC. Sorbent development on these lines is continuing.

Reactor designs for the hot gas desulfurization subsystem include fluidized, fixed and moving beds. Fixed-bed reactors have been used thus far in all zinc ferrite sorbent research. At least two beds would be required for a commercial operation so one bed could undergo regeneration as the other is sulfidizing. A moving-bed design offers some advantages. The sorbent would enter the top of the reactor and move slowly down its length for mechanical withdrawal at the bottom. The gas would flow countercurrently, exposing the exiting sorbent to the highest reactant concentration for best conversion. The sulfidized sorbent would be transferred to another vessel for regeneration and cleaning of any particulates. The moving-bed design could serve as a polishing granular-bed filter, allowing fine particulates that might escape an upstream cyclone trap to adhere to the zinc ferrite pellets with subsequent blowoff during regeneration.

The sorbent regenerates sulfur to SO₂ which must then be processed for ultimate disposal. Passing the SO₂-bearing gas through a hot fluidized bed of limestone or dolomite appears to be an attractive way of capturing the SO₂. If steam is used as a coolant and diluent during regeneration, it can be used to partly fulfill the steam requirements of the gasifier so the exit gas from the limestone fluidized bed could be injected into the gasifier. This suggests that the limestone bed could be integral to the gasifier, which in turn raises the possibility of in-bed desulfurization, to be discussed later.

An example of how the zinc ferrite process can be integrated into an attractive system is illustrated in Figure 6. This system uses an air-blown fixed-bed gasifier with a zinc ferrite desulfurization process. The plant size is 115 MW with a calculated efficiency of 36.8%. This plant is sized to a General Electric Frame 7 gas turbine. The process flow sheet is considerably simpler than the cool water diagram in Figure 1, and the gasification and power islands are well-integrated. But further improvements are possible.

REMOVAL OF PARTICULATES AND ALKALI COMPOUNDS

The other aspect of hot gas cleanup is particulate removal. Sufficient protection for the zinc ferrite reactor can likely be provided by cyclone separators for fixed- and fluid-bed gasifiers.

METC has been exploring several high-temperature, barrier-type filters for gasifier and fluidized-bed combustor applications. One of the most promising is the cross-flow filter, a ceramic device with a waffle-like design. Dirty gas enters one side of many porous surfaces formed by the monolith design and flows through the filter material, leaving the particulate matter behind on the surface. The filter elements are assembled in layers that can be hung on supporting manifolds in a pressure vessel. Periodic blowback clears the surface of accumulated particulate which falls to the bottom of the vessel for removal. The filter element has been tested at bench scale under fluidized-bed combustor and gasifier conditions, with temperatures up to 1,600°F (1,145 K) and pressures to 17 atm (1,725 Kpa). The filter efficiency has been 99.9+ % under high dust loadings. Its development is continuing.

Alkali was identified as a potential pollutant harmful to the turbine, but evidence is mounting that this may be a readily controllable problem in coal gas systems. As the

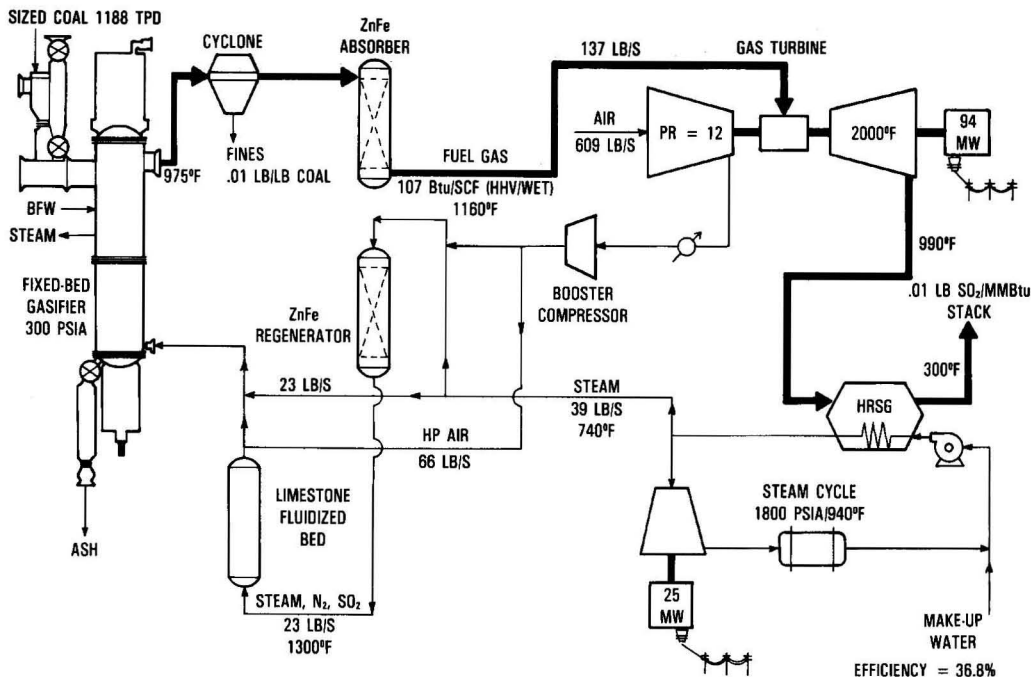


Figure 6. Fixed-bed gasifier/hot gas cleanup/combined cycle — 115 MW.

fuel gas is cooled below 1,400°F (1,035 K), alkali appears to condense on particulate that can subsequently be removed with the ash particles. Also, many of the natural alkali compounds in coal may not be released but remain combined with other ash components. If alkali vapor is found to be a problem, a disposable natural sorbent, emathalite, has reduced concentrations to a few ppm in research tests.

IN-BED DESULFURIZATION

The prospects of disposing of the SO₂ stream from regeneration using a limestone bed in the gasifier as well as achieving direct sulfur removal is an exciting subsystem simplification that can offer very substantial benefits. Equilibrium calculations predict that sulfur removal in excess of 50% can be achieved with calcium (limestone) sorbent, depending on gasifier temperature and weight percent steam present in the gas. Fluid-bed gasifiers operate around 1,800° (1,255 K) and with about 20% steam, implying that H₂S equilibrium partial pressures should be about 200 to 300 ppm, with sulfur removal efficiencies for high-sulfur coal in the 90+ % range. Fixed-bed gasifiers contain temperature zones ranging from 2,000°F (1,365 K) in the bottom combustion zone to 1,000°F (810 K) at the top of the gasification zone. Injection of SO₂ from zinc ferrite regeneration into the bottom of the gasifier should result in the capture of some portion of the SO₂ as sulfate and conversion of the remainder as H₂S for subsequent capture or recycle to the zinc ferrite system in the product gas.

In-bed desulfurization tests were carried out on the KRW fluid-bed gasifier at Waltz Mill, Pennsylvania. This is a 12 ton/day (10,890 Kg/day) pilot-scale gasifier, using a central jet to introduce reactants (air was the oxidant for these tests) and two stages of cyclones for recycling elutriated fines to the gasifier. Earlier tests using a 2.3% sulfur bituminous coal gave 70 to 80% sulfur capture. Later tests

used a 4.5% sulfur coal and both dolomite and limestone sorbents at calcium/sulfur mole ratios of 1.5 to 2.0. H₂S effluent concentrations at various operating conditions ranged from 484 to 686 ppm, and percent desulfurization from 83 (for the lowest-sulfur coal feed of 2%) to 91% averaged over steady-state periods of several hours. Shorter term sulfur removal reached 93%, comfortably above the NSPS requirement of 90%. Elutriation of carbon fines from the bed was reduced substantially in some cases, and the product gas quality increased from the normal average of 123 Btu/scf to 153 Btu/scf (4,580 to 5,700 KJ/m³). This is likely related to reduced heat losses and better carbon conversion efficiencies. SO₂ recycle will be tested in 1987.

These results for a fluid-bed gasifier support the option to use the gasifier as the primary device for sulfur removal and the hot gas cleanup system as a smaller polishing stage. All of the economic trends of such a system would be favorable.

Results of an early look at SO_x control economics compare zinc ferrite cleanup with some alternatives, shown in Figure 7. This plots the cost per ton of SO₂ removed from a 3.8% sulfur coal versus the number of sorbing/desorbing cycles available before sorbent replacement. Curves "a" and "b" are for current and target sorbent sulfur loadings with no in-gasifier capture. Curves "c" and "d" correspond to "a" and "b," respectively, but with 90% in-bed sulfur capture by limestone injection into the gasifier. Current testing places the technology near Curve "c," so an acceptable sorbent must last about 40 cycles to be cheaper than retrofit Limestone Injected Multiple Burner (LIMB) technologies. As Figure 5 showed earlier, that target has essentially been met.

ADVANCED POWER SYSTEMS

The last piece of the simpler IGCC system revolves around developments in the turbine industry. Industrial

— 3.8% Sulfur Coal —

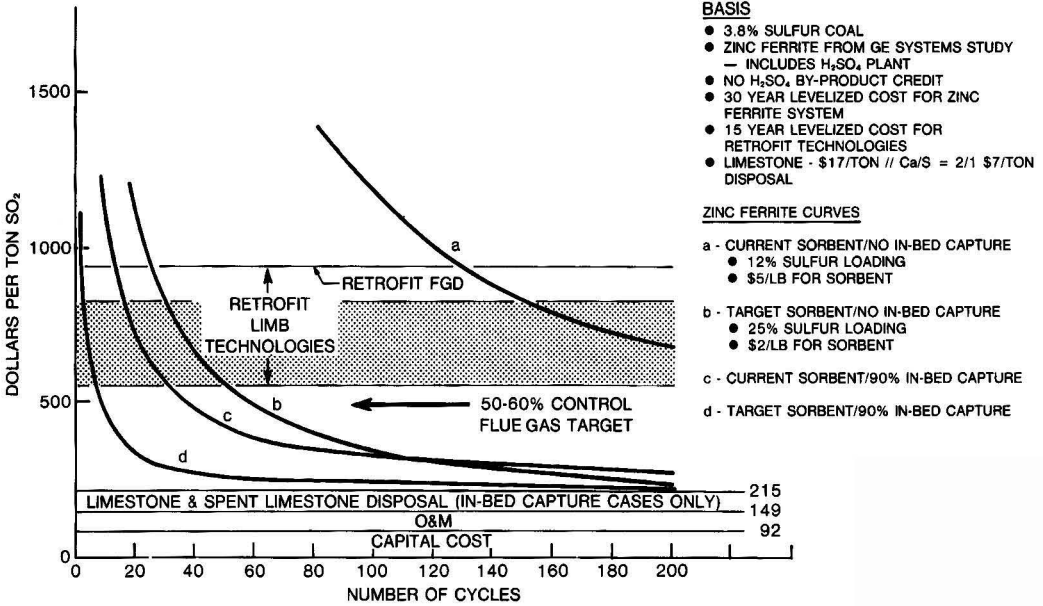


Figure 7. Sulfur removal costs versus sorbent durability.

power turbines were sized such that incremental additions would be at 200 MW or above. An extension of dual-shaft, aircraft-derivative turbine design allows smaller capacity additions plus increases in efficiency due to steam injection into the turbine and good integration with the gasification/cleanup subsystems. These machines do not operate as combined cycles. In the steam-injected gas turbine (STIG), steam from the heat recovery steam generator is injected into the combustor and between some expander stages to increase mass flow. The residual heat in the exhaust gas is captured in the HRSG. Such machines are already commercially available using natural gas as the fuel. In a coal gasification system, some of the steam is consumed in the gasifier. The air compressor is tapped for high-pressure air for the gasifier which is boosted via a separate steam-driven compressor to the higher gasifier pressure. A final development would be to intercool the STIG (an ISTIG) between the compressor stages. This raises the mass flow capacity and the efficiency for a machine of fixed size. These two options allow capacity increments, built on existing commercial machines, in the range of 60-120 MW.

A SUPERIOR SYSTEM

It is possible to design systems that presume success of all these ongoing developments and determine their projected efficiency and cost. Figure 8 plots the system efficiency versus capacity for METC calculations of various combinations of gasifiers [9]. Systems using fluid-bed gasifiers require some heat transfer or quenching equipment to bring the gas temperature down to the zinc ferrite operating range. The highest efficiency system is an air-blown, fixed-bed ISTIG with in-bed desulfurization and capture of SO₂ recycled from a zinc ferrite cleanup system. The module size is 133 MW, and cycle efficiency is 42.1%. Figure 9 shows the process diagram for the system. Its simplicity in contrast with Figure 1 is evident.

The power, gasifier and cleanup subsystems are fully integrated. Cost of electricity (COE) for the fixed-bed ISTIG system computes to 57 mills/kWh, 10th year current dollars levelized, with a capital cost of \$855/kW. The cycle efficiency of an air-blown fixed-bed system with a STIG was calculated to be 39.8% with a module size of 118 MW, COE of 61 mills/kWh and a capital cost of \$810/kW. Use of a fluidized-bed gasifier with the ISTIG resulted in a system of slightly higher COE and capital costs. These calculations presumed as much factory assembly as possible with minimal field erection and standardization of design to minimize engineering costs.

WHAT REMAINS TO BE DONE

The scenario depicted above shows the potential, not the reality of the technologies. These possibilities are be-

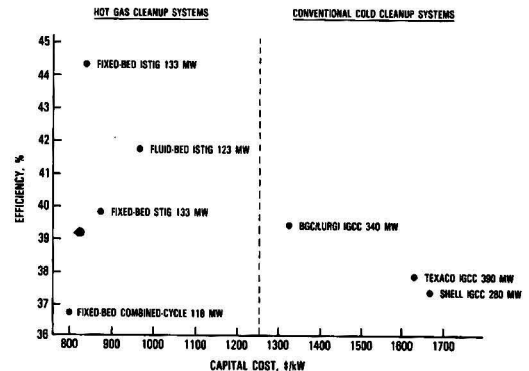


Figure 8. Calculated efficiencies for several IGCC system designs.

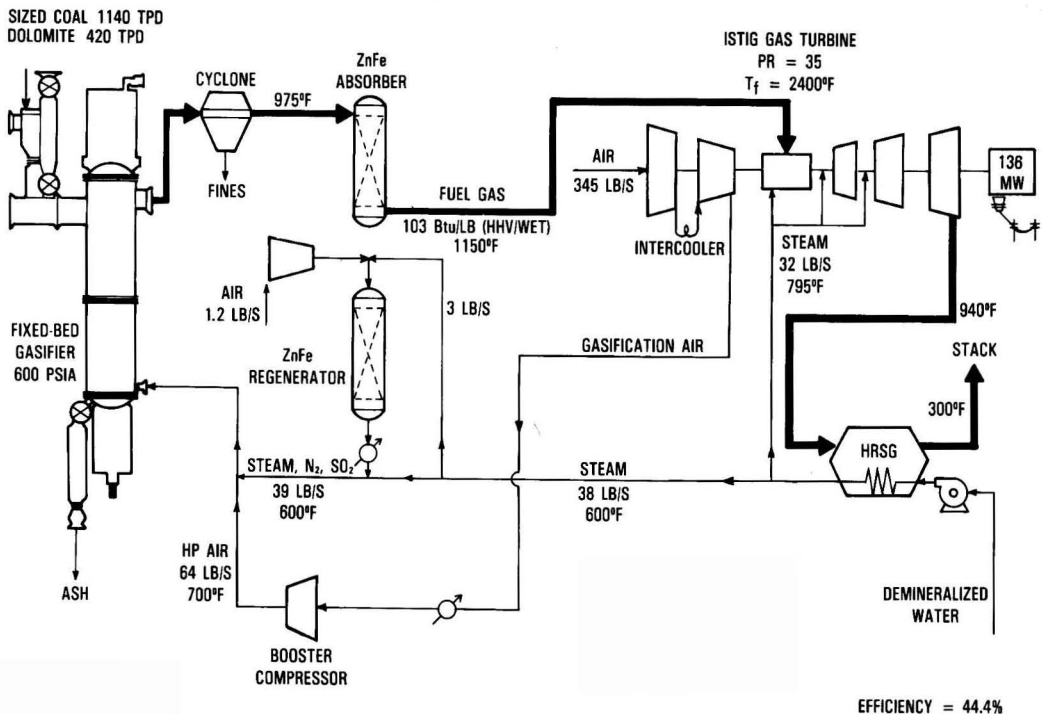


Figure 9. Process diagram for an ultra-efficient IGCC system design.

ing explored by the Government and industry through research. A scale-up of hot gas cleanup with the KRW fluid-bed, pilot-scale gasifier is being tested. The zinc ferrite vessel is a fixed-bed, 54 inches in diameter (1.4 m) with bed depths of 4 and 8 feet (1.2 and 2.4 m). This is being operated with continued in-bed desulfurization and eventually with SO_2 reinjection into the gasifier, providing the first system integration. METC has begun corresponding experiments with its 24 ton/day (21,780 kg/day) fixed-bed gasifier. METC is also continuing development of advanced sorbent systems that operate at higher temperatures and may regenerate to elemental sulfur instead of SO_2 . Support studies of alkali control (if needed) and suppression of nitrogen oxide emissions will also continue. Hot gas cleanup, air blowing and in-reactor desulfurization are all being considered by industry for entrained-bed gasifiers. A gas-fired STIG is now offered commercially, and intercooling, although a large development effort, is possible within the basic design of the aircraft-derivative combustion turbine.

Development of complete IGCC systems using hot gas cleanup has also begun. The Japanese have a 40 ton/day (36,300 kg/day) pilot plant at Yubari in northern Japan that uses a fluidized-bed gasifier, a fluidized-bed iron oxide sulfur sorbent, a screened moving-bed granular filter for particulate removal and a combustion turbine. In the United States, two of the clean coal technology demonstration projects selected by the Department of Energy for negotiation are IGCC systems. Both use the METC zinc ferrite cleanup system. One uses a fluid-bed gasifier with combined-cycle turbines, and one uses a fixed-bed gasifier with a STIG.

Obviously, the technology has far to go and perhaps not all goals will be achieved. But results continue to favor an optimistic view. If so, utilities will have many attractive options for power generation in the 1990's and the consumer can look forward to clean, economic power from coal as a result.

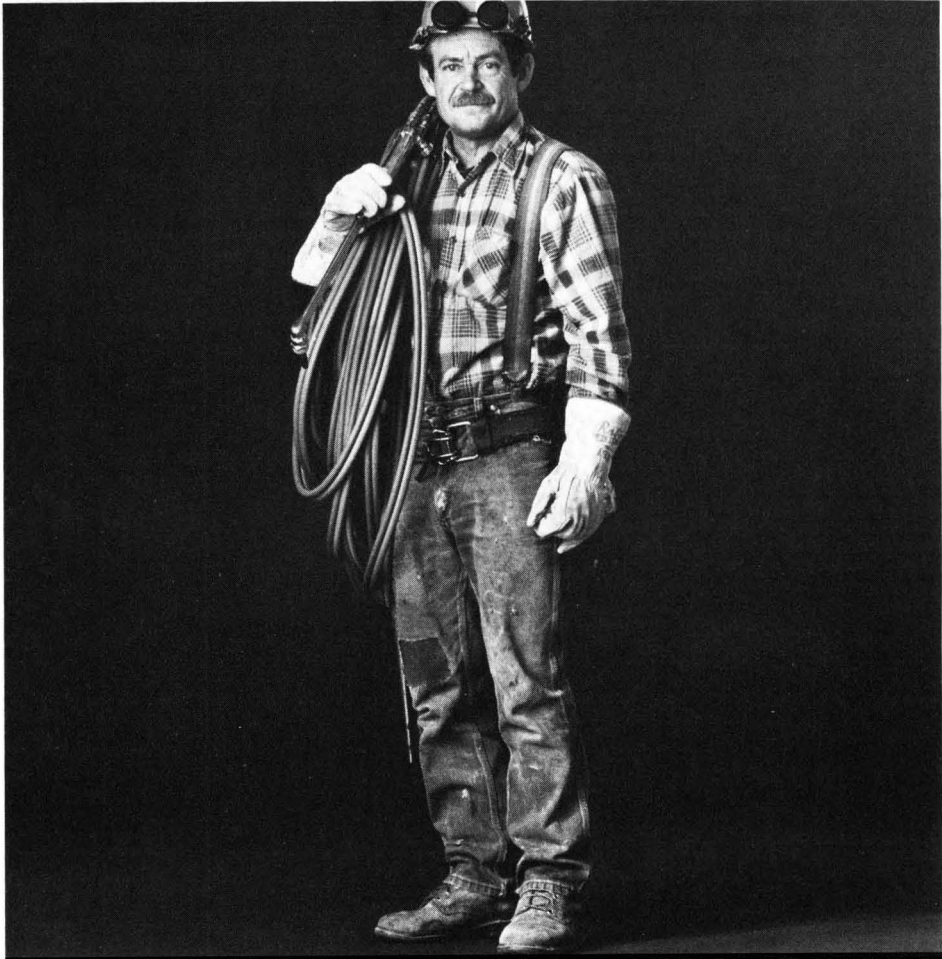
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