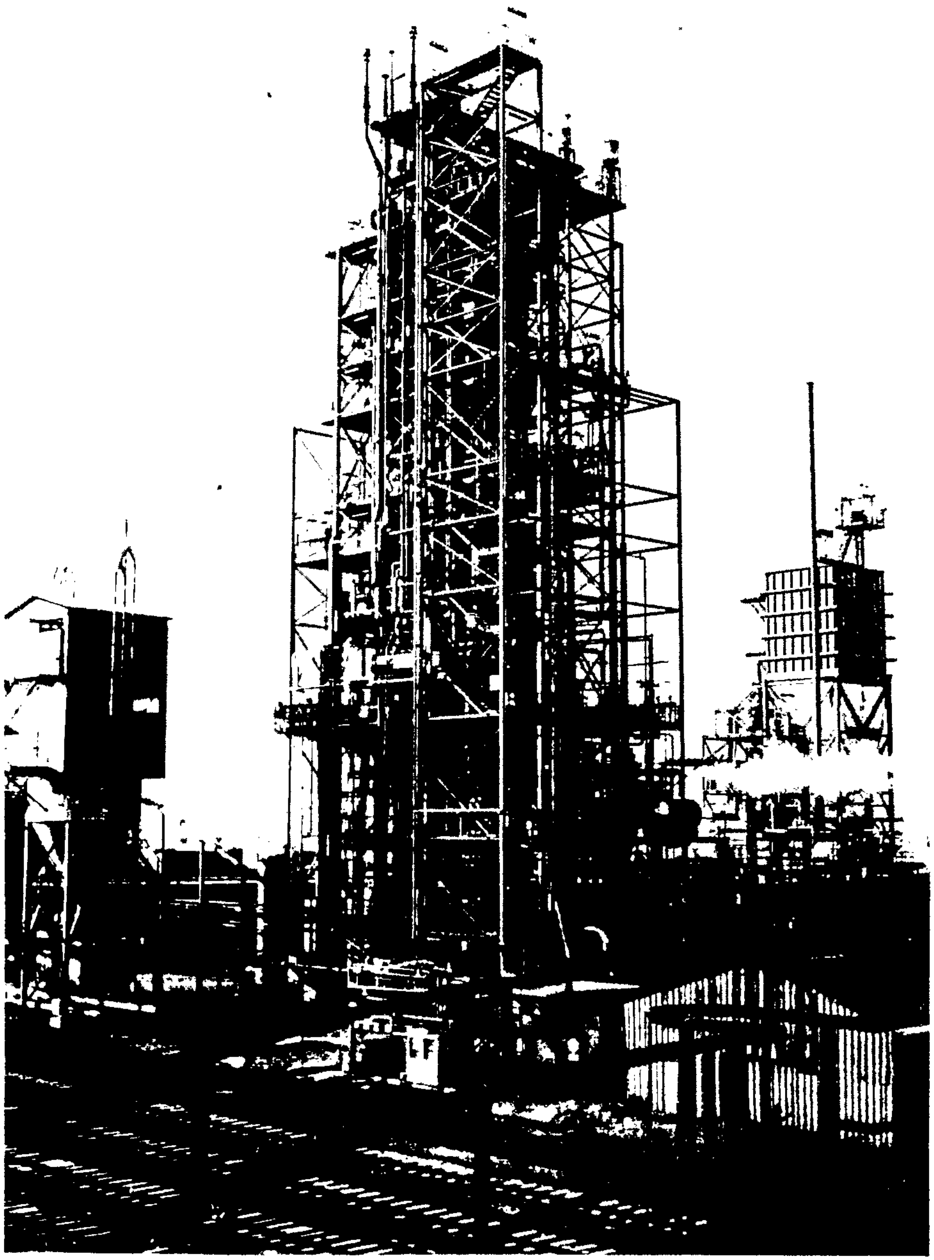


Fig. 1 Shell-Koppers coal gasification process 150 T/D pilot plant at Deutsche Shell's Harburg Refinery

SHELL-KOPPERS GASIFICATION PROCESS DEVELOPMENT

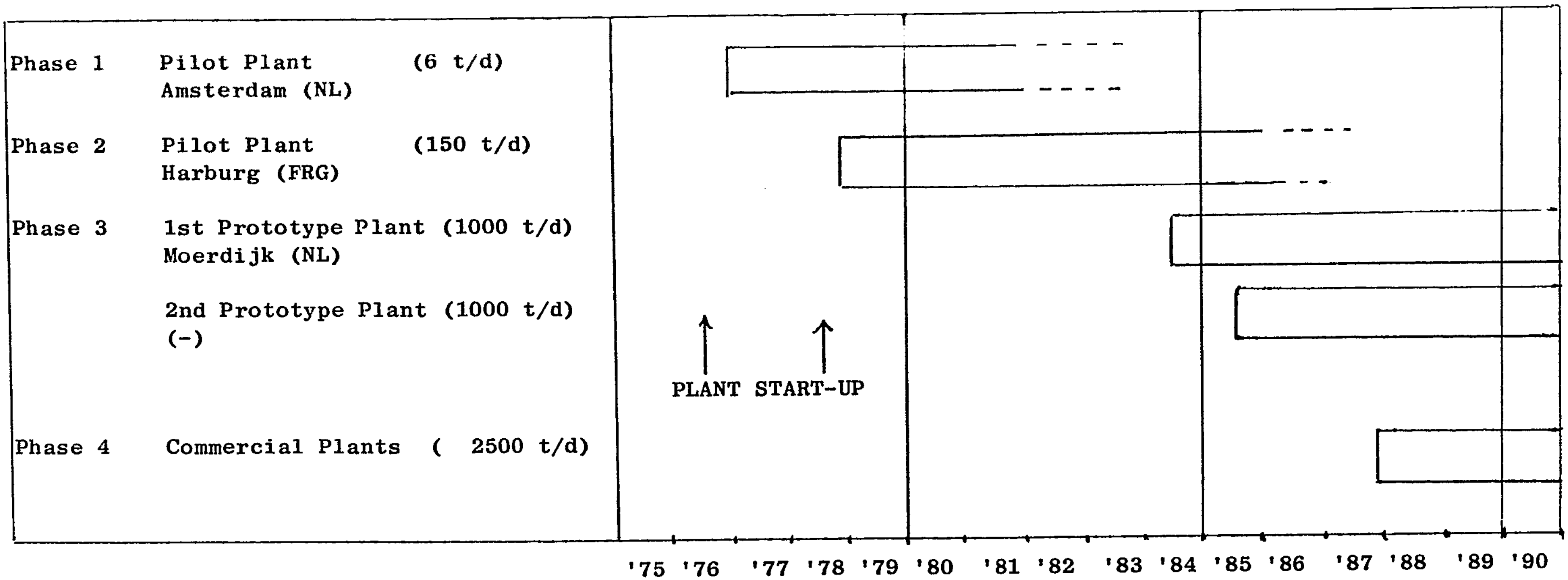
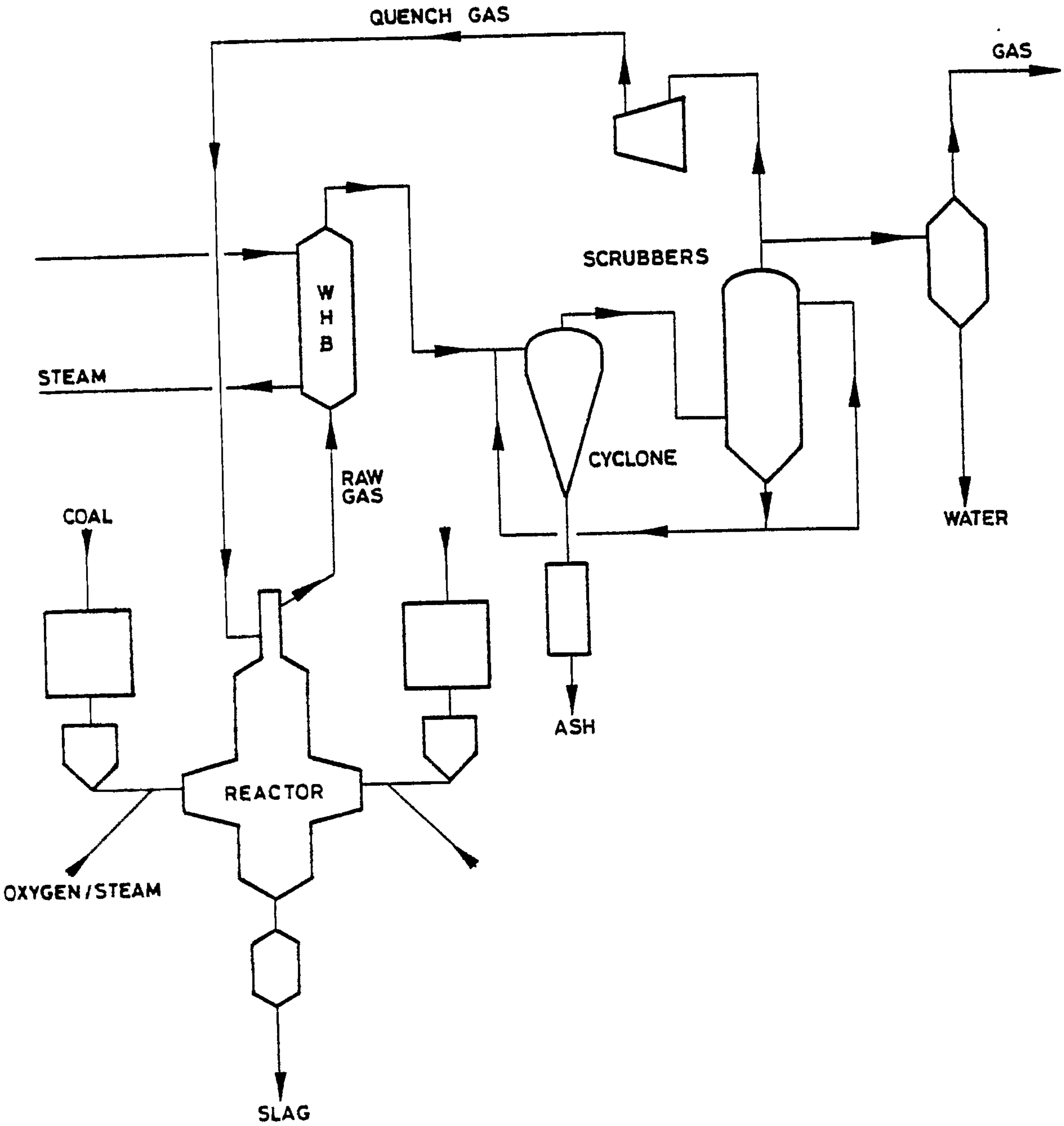


Fig. 2

SHELL-KOPPERS COAL GASIFICATION



ELECTRICITY GENERATION VIA
GASIFICATION/COMBINED CYCLE

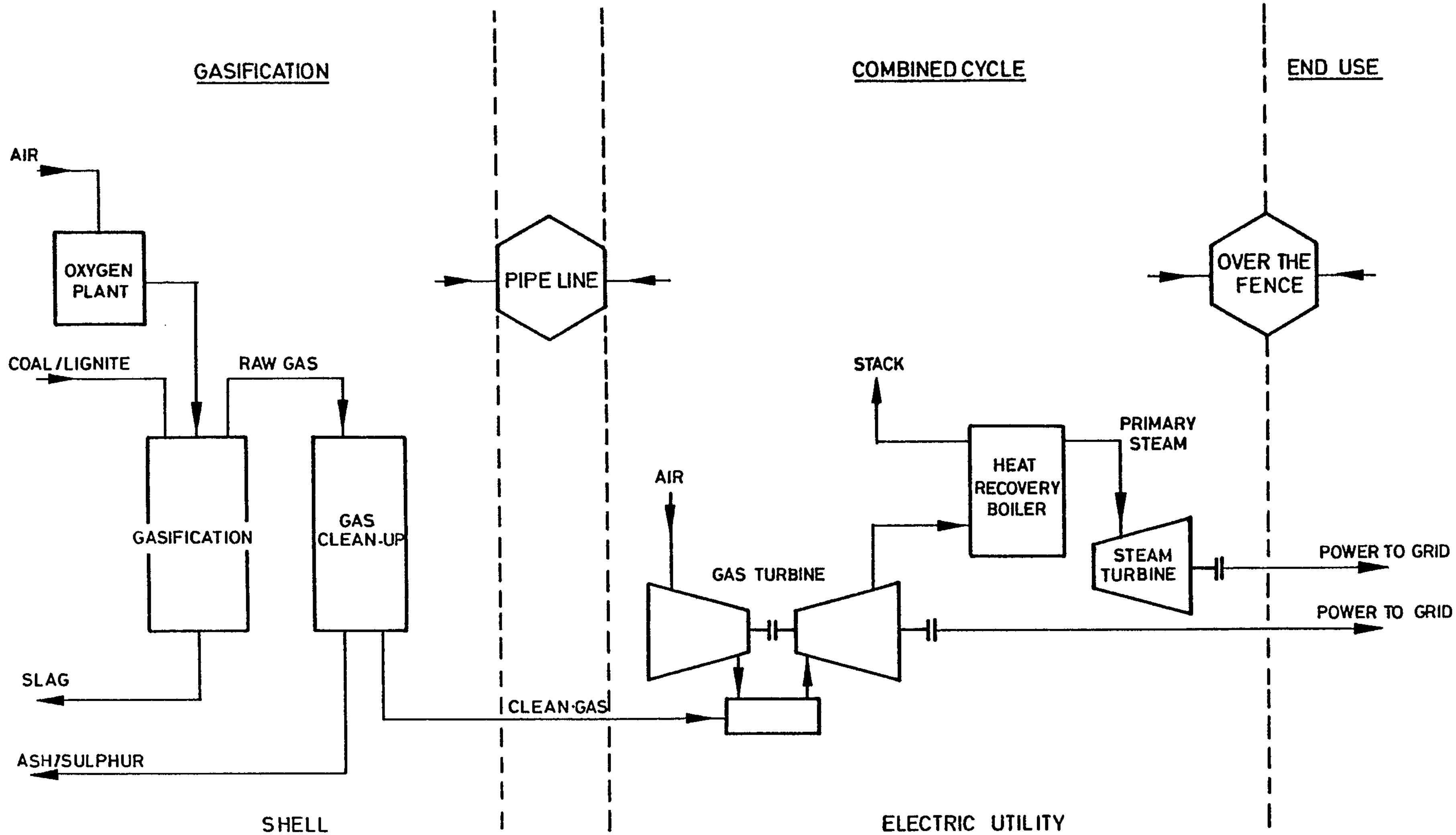


FIG. 4

SHELL - KOPPERS COAL GASIFICATION

BLOCK SCHEME FOR FUEL GAS, SYNGAS, OR H_2 PRODUCTION

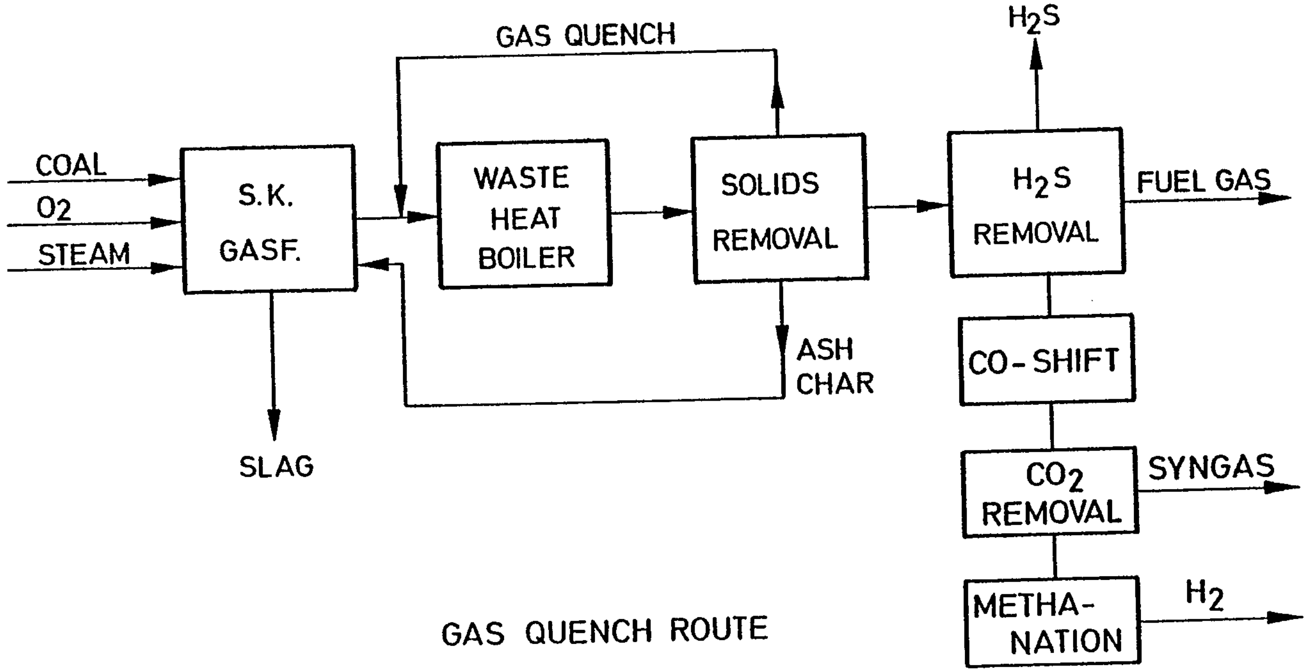


Fig. 5

Fig. 6

TYPICAL BLOCK SCHEME FOR HYDROGEN OR SYNGAS PRODUCTION
SHELL-KOPPERS COAL GASIFICATION

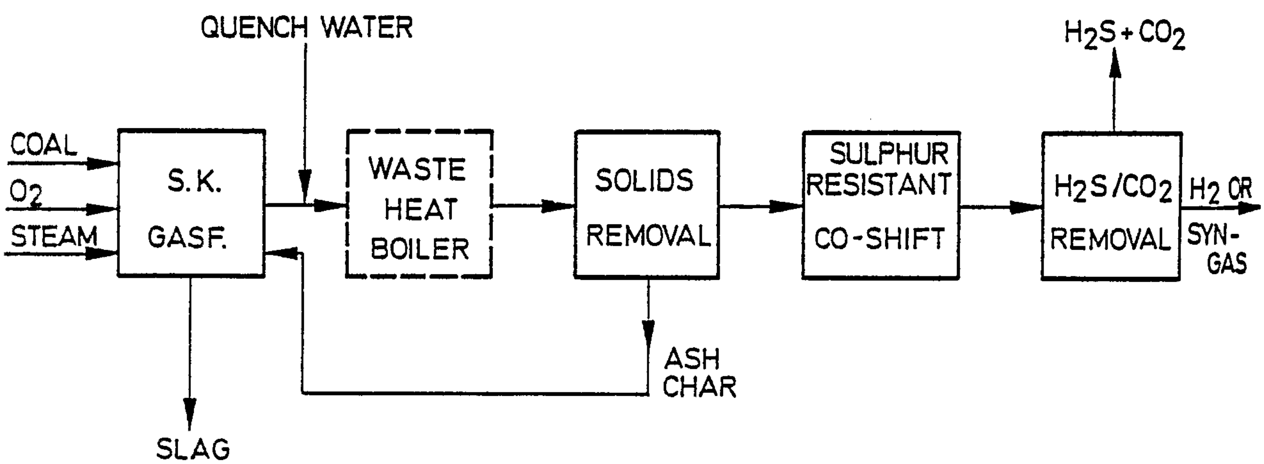
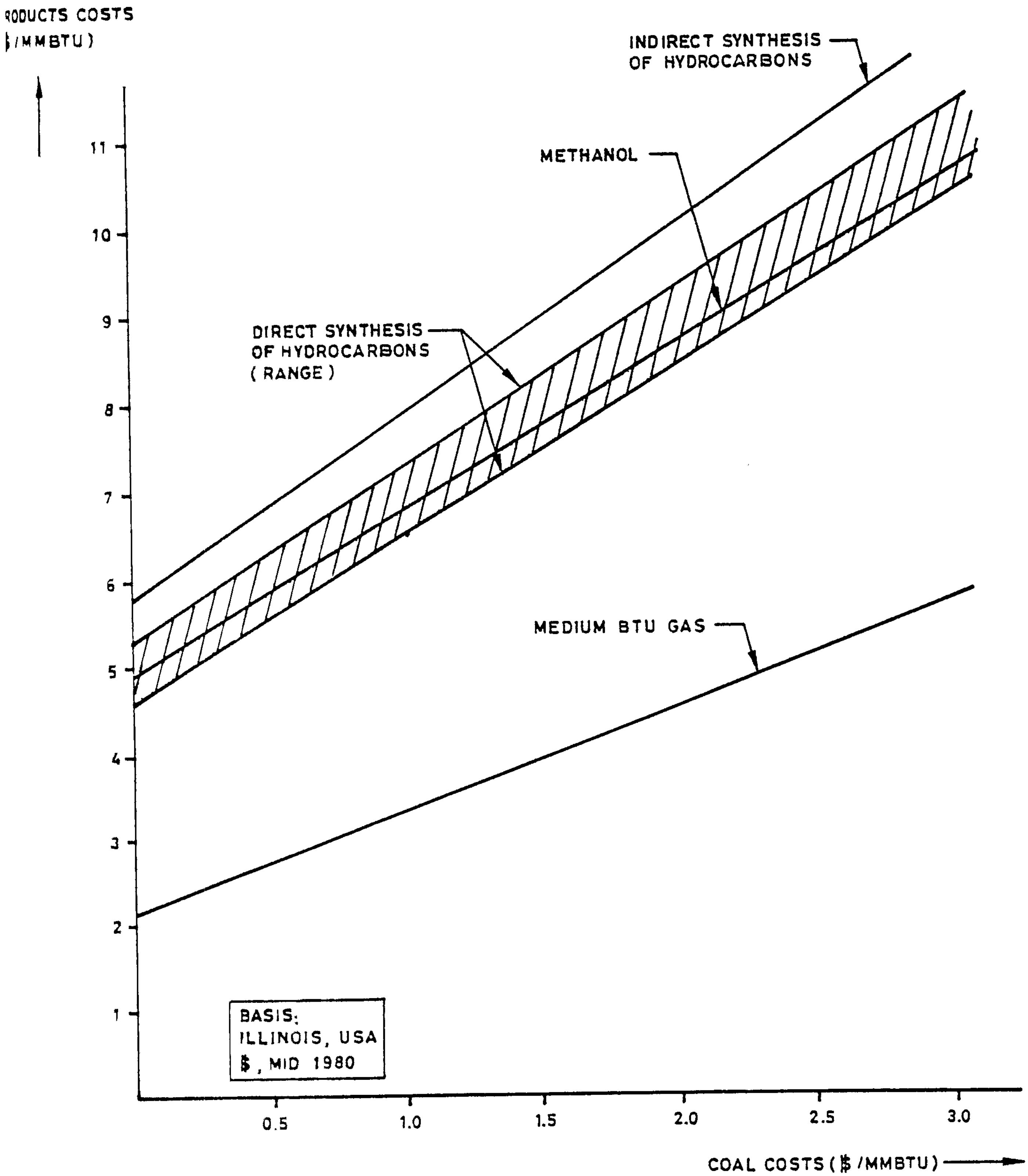


FIG. 7

COAL CONVERSION COSTS



SHELL-KOPPERS COAL GASIFICATION

ECONOMICS FOR VARIOUS APPLICATIONS

THE MEMPHIS DEMONSTRATION PLANT PROGRAM

J. G. Patel and R. W. Gray

THE MEMPHIS DEMONSTRATION PLANT PROGRAM

J. G. Patel
Institute of Gas Technology
R. W. Gray
Memphis Light, Gas and Water Division

SUMMARY

Memphis Light, Gas and Water (MLGW) and the U.S. Department of Energy (DOE) have a joint program to design, construct, and operate an industrial fuel gas demonstration plant in Memphis, Tennessee. The plant will use the Institute of Gas Technology's U-GAS® Process to produce fuel gas equivalent to 50 million cubic feet per day of natural gas from 3200 tons of Western Kentucky coal. During Phase I of the program, main activities have been design of the plant, operation of the U-GAS pilot plant, the compilation of an enviornmental report and a definitive plant cost estimate. Phases II and III negotiations with DOE are being finalized. The total cost of Phases II and III is to be cost-shared by DOE and MLGW. The demonstration plant program's goal is to test the feasibility of a multiple-user coal gasification plant located in an urban area.

INTRODUCTION

The Memphis Light, Gas and Water Division of the City of Memphis (MLGW) supplies electric, gas, and water utilities to its customers. In 1970, Memphis residences and industries used 93.4 billion cubic feet of gas. Curtailment from the supplier decreased the supply to 62 billion cubic feet by 1977. With the goal of assuring an adequate industrial fuel supply, MLGW became a partner with the U.S. Department of Energy (DOE) in an Industrial Fuel Gas Demonstration Plant Program.

The plant will convert high-sulfur bituminous coal into medium-Btu gas using the Institute of Gas Technology's (IGT's) U-GAS Process. It will provide for industrial users the equivalent of 50 million cubic feet of natural gas per day from about 3200 tons of coal. The gas, consisting mainly of carbon monoxide and hydrogen with a heating value of about 300 Btu/SCF, will be suitable for industrial use.

The overall project is being conducted in three phases over an 8-year period. Phase I has been completed. MLGW and DOE are completing final negotiations for Phases II and III. The demonstration plant program is aimed to test the feasibility of a multiple-user coal gasification plant located in an urban area.

THE PROGRAM

DOE executed a contract with MLGW that requires MLGW to perform process analysis, design, procurement, construction, testing, operation, and evaluation of a plant to demonstrate the feasibility of converting high-sulfur bituminous coal to industrial fuel gas with a heating value of about 300 Btu/SCF The demonstration plant will be based on the U-GAS Process, and its product gas is to be used in commercial applications in Memphis, Tennessee. The plant, to be located on the Mississippi River, will be designed to produce the fuel gas from 3200 tons per day of Western Kentucky coal. The fuel gas will be pipelined to industrial customers in the Memphis area to replace their present energy source of natural gas or fuel oil and will be used mainly for raising steam or for process heat.

To carry out the program, MLGW has established an industrial project team as shown in Figure 1. The role of each team member is as follows:

- MLGW, the prime contractor, has the responsibility of executing the demonstration plant program. It will also be the owner of the plant and the distributor of the industrial fuel gas

- Foster Wheeler Energy Corporation (FWEC) is the architect-engineer and construction manager. FWEC has broad experience in coal handling and processing and the capability to execute large engineering projects

- IGT is the developer of the U-GAS Process and operator of the pilot plant for obtaining design data.

- Delta Refining Company (DRC) is an oil refining company that provides operating experience in design and operation of the plant

- Various other subcontractors, the largest being Energy Impact Associates (EIA), which is assisting in preparing the environmental report, provide services on the program.

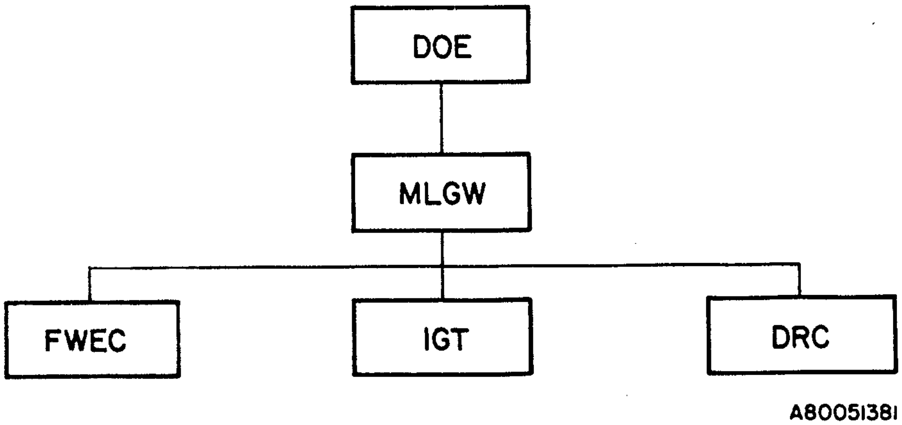


Figure 1. ORGANIZATION CHART

The overall program is being conducted in three phases over an 8-year period. Table 1 shows the schedule and major milestones. Phase I, completed in December 1979, consisted of Program Development and Conceptual Design; Phase II, estimated to last 48 months, involves detailed design and construction; Phase III, which covers a 20-month period, consists of operation of the demonstration plant. At the completion of the program, the plant will be operated by MLGW as a commercial venture. The financing arrangements for the three phases are shown in Table 2. Phase I was completely funded by DOE; Phase II and III costs are to be shared. MLGW will finance its share with municipal revenue bonds. The cost-sharing agreements with the Government require eventual payback of all the costs incurred by the Government for the revenue of the plant during commercial operation. The main aim of the cost-shared project funding is, therefore, to reduce MLGW's financial risk in this first-of-a-kind plant.

Phase I of the program was completed in December 1979. The main activities of Phase I are shown in Table 3. The conceptual design for the commercial plant, the demonstration plant design, and the plant cost-estimate have been completed and have been submitted to DOE. Successful tests have been conducted at the U-GAS pilot plant with the candidate coal, and adequate data have been obtained for the demonstration plant design. All environmental information from the field, the plant design, and all assessments have been assembled in the form of an environmental report. This has been submitted to DOE so they can prepare an Environmental Impact Statement in compliance with the National Environmental Policy Act. The present schedule indicates the final completed Environmental Impact Statement with a record of decision will be made in October 1980. The plant is designed to meet all applicable environmental regulations. The plant site is located on the Mississippi River within the existing floodplains. A site evaluation and selection report on the present and other feasible alternate sites has been completed.

Table 1. PROGRAM SCHEDULE

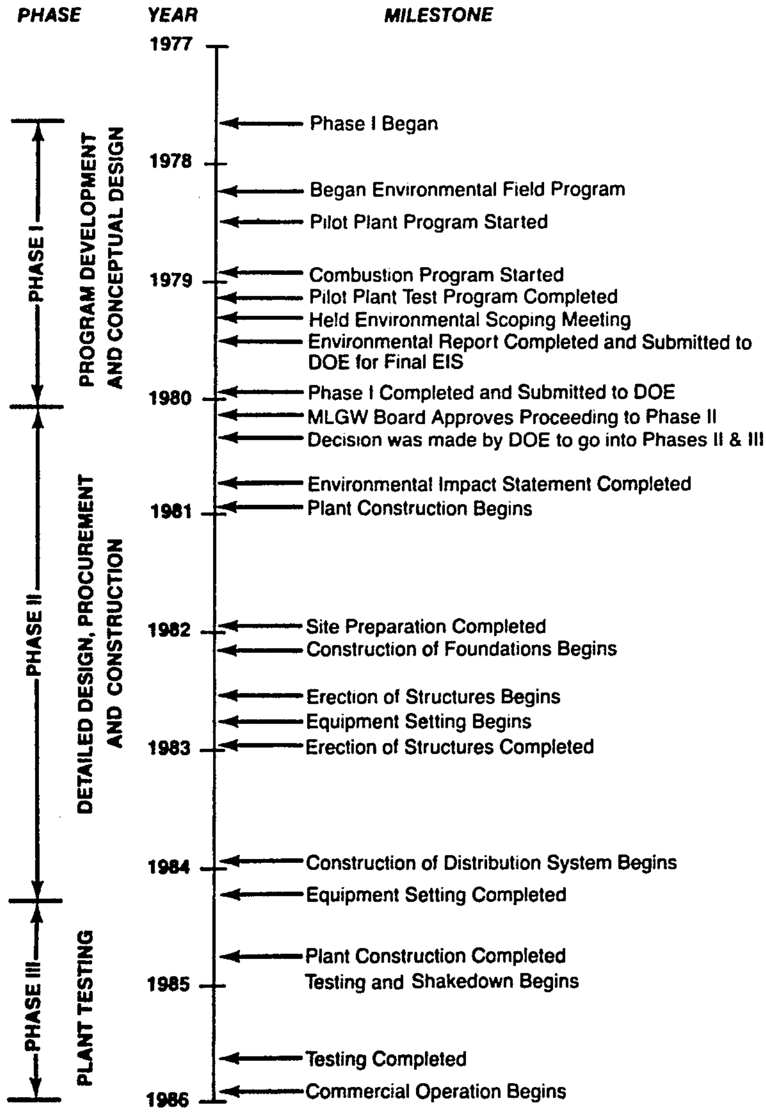


Table 2. PROJECT FUNDING BY PHASE

Phase	Cost, Million $	Govt. Funding	MLGW Funding	Duration, months
I	11	11	0	26
II	350*	Cost Shared**		48
III	80*	Cost Shared**		20

* Estimated costs.
** Funding levels between MLGW and DOE are still to be determined.

Table 3. PHASE I ACTIVITIES

Conceptual Design of Commercial Plant
Demonstration Plant Design and Cost Estimate
Demonstration Plant Environmental Analysis
Technical Support - Pilot Plant Operations

After evaluating the offers from the prime contractors of the two competitive programs carried through Phase I, DOE selected MLGW to start negotiations for Phase II and Phase III in late February. The main negotiating point has been the cost-sharing arrangement of the remaining project costs. Presently, the negotiations are being finalized; the final contract is expected to be signed by the time this paper is presented.

The remainder of this paper will discuss the results of the pilot plant tests conducted with the candidate coal to obtain the design basis for the demonstration plant, the details of the demonstration plant design, and some economics of the commercial plant.

TECHNICAL SUPPORT PROGRAM

U-GAS PROCESS

The U-GAS Process has been selected by MLGW and DOE for the demonstration plant program. The U-GAS Process has been developed by IGT to produce a medium-Btu (300 Btu/SCF) fuel gas from coal in an environmentally acceptable manner.

The process shown in Figure 2 accomplishes four important functions in a single-stage, fluidized-bed gasifier. It decakes coal, devolatilizes coal, gasifies coal, and agglomerates and separates ash from char.

In the process, washed coal (1/4-inch X 0) is dried only to the extent required for handling purposes. It is pneumatically injected into the gasifier through a lockhopper system. Within the fluidized bed, coal reacts with steam and oxygen at a temperature range of 1750° to $1900^{\circ}F$. The temperature of the bed depends on the type of coal feed and is controlled to maintain nonslagging conditions for ash. The opeating pressure of the process depends on the ultimate use of product gas and may vary between 50 and 350 psi. The pressure must be optimized for a particular system. At the specified conditions, coal is gasified rapidly, producing a gas mixture of hydrogen, carbon monoxide, carbon dioxide, and a smaller percent of methane. Because reducing conditions are always maintained in the bed, nearly all of the sulfur present in the coal is converted to hydrogen sulfide.

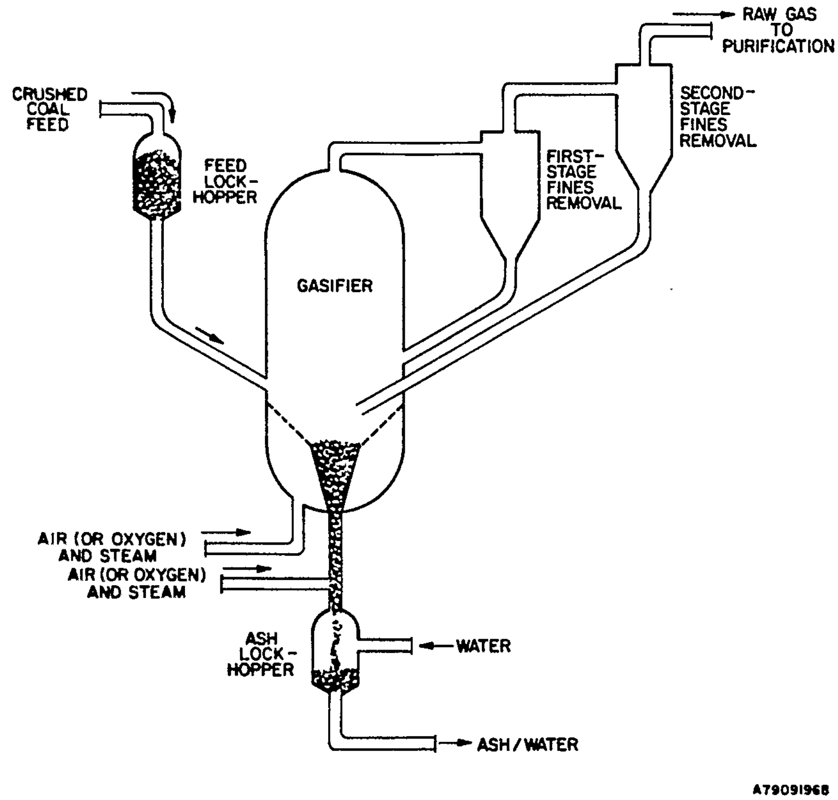


Figure 2. SCHEMATIC OF THE U-GAS GASIFIER

Simultaneously with coal gasification, the ash is agglomerated into spherical particles and separated from the bed. Part of the fluidizing gas enters the gasifier through a sloping grid. The remaining gas flows upward at a high velocity through the ash agglomerating device and forms a hot zone within the fluidized bed. The temperature within the hot zone is greater than at other locations in the bed. High ash-content particles agglomerate under these conditions and grow into larger and heavier particles. Agglomerates grow in size until they can be selectively separated and discharged from the bed into water-filled ash hoppers where they are withdrawn as a slurry. In this manner, the fluidized bed achieves the same low level of carbon losses in the discharge ash that is generally associated with the ash-slagging type of gasifiers.

Coal fines elutriated from the fluidized bed are collected in two external cyclones. Fines from the first cyclone are returned to the bed, and fines from the second cyclone are returned to the ash agglomerating hot zone, where they are gasified, agglomerated with bed ash, and discharged with ash agglomerates. The raw product gas is significantly free of tar and oils, thus simplifying ensuing heat recovery and purification steps dictated by the end use of the product gas.

PILOT PLANT PROGRAM

Most of the U-GAS Process development work has been performed in the pilot plant, which was put into operation in 1974. It is located at IGT's test facilities in southwest Chicago; the same facilities also contain the HYGAS® pilot plant. The pilot plant consists of a gasifier and all required peripheral equipment, with utilities and other support services provided by the HYGAS pilot plant. The major equipment in the pilot plant is a drying and screening system, feed storage silos, a lockhopper system (weighed) for feeding a dry pulverized material at rates up to 3000 pounds per hour, a refractory-lined fluidized-bed reactor with a special agglomerate withdrawal system in its base, a product gas quench system, a cyclone system for removal and recycle of elutriated fines, a product gas scrubber, a product gas incinerator, and all necessary instrumentation and controls.

The U-GAS Process development work is divided into three separate parts: Part 1 during which the process feasibility was demonstrated using metallurgical coke and char as feed; Part 2 during which the pilot plant was modified to feed coals, and trial tests were made with coal; and Part 3 during which process feasibility was proved using coal as feed, and data were developed for scale-up of the process and design of the demonstration plant. The Part 3 operations were conducted with Western Kentucky No. 9 coal. The properties of the coals tested in the pilot plant are shown in Table 4. The objective was to provide mechanical, operating, environmental, and process data for the preliminary design of the demonstration plant using Western Kentucky No. 9 coal. A total of 16 test runs were conducted over the period of 15 months beginning with January 1978. A summary of the U-GAS pilot plant tests are shown in Table 5.

The highlights of test operations were as follows:

- The pilot plant tests firmly established process feasibility and provided a strong data base for completing the preliminary demonstration plant design.
- Four consecutive, extended-period tests of up to 200 hours were conducted during which good-quality raw-product gas (285 Btu/SCF) and high-ash-content (80 to 90 weight percent) ash agglomerates were produced from Western Kentucky coal.
- A technique of feeding caking coals directly into the gasifier without pretreatment was perfected. Over 400 tons of caking coal with a free-swelling index (FSI) of 4 to 7 were fed.
- Stable operability of the gasifier while recycling entrained coal fines back into the gasifier under continuous agglomerating conditions was demonstrated.

Table 4. PROPERTIES OF COALS TESTED IN PILOT PLANT
(Western Kentucky No. 9 Coal)

	Washed	Unwashed
Proximate		
Ash	12.0	18.9
Volatile	35.8	34.4
Fixed Carbon	49.1	45.1
Ultimate		
Carbon	72.2	64.3
Hydrogen	4.5	4.4
Oxygen	6.8	6.2
Nitrogen	1.2	1.1
Sulfur	3.1	4.6
Chlorine	0.13	0.19
Ash	12.1	19.9
Free Swelling Index (FSI)	4-7	5-6
Higher Heating Value	12,498 Btu/lb	11,570 Btu/lb

Table 5. SUMMARY OF PILOT PLANT TESTS

Test Run	Feed Material	Dates	Operating Period,* hr	Coal Feed, ton	Coal Conversion** Attained, %
124	Run-of-mine Western Kentucky bituminous coal	6/78	168	84	81
130	Washed Western Kentucky bituminous	11/78	106	88	76
131	Washed Western Kentucky bituminous coal	12/78	104	70	94
132	Washed Western Kentucky bituminous coal	1/79	74	47	89
133	Washed Western Kentucky bituminous coal	2/79	153	104	92

*Total hours of operation with coal during the run.

**Based on moisture, ash-free coal feed to the gasifier.

- Data related to environmental aspects of the U-GAS Process, particularly wastewater characteristics, which indicated the presence of only trace quantities of tar and oils, were provided.

- The pilot plant operated for more than 100 hours at pressures of up to 60 psia (gasifier design pressure is 65 psia), providing the applicability of the ash agglomeration technique and the ash agglomeration discharge mechanism at moderate elevated pressures.

- A broad operating window for the major operating variables of temperature, superficial velocity, and bed ash-content was established.

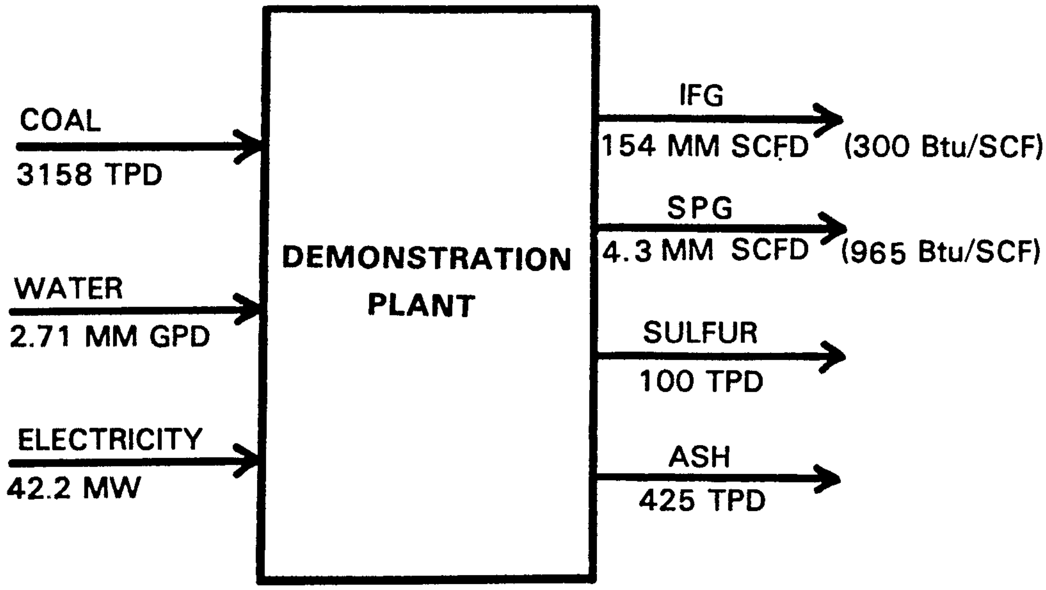
- Suitable materials of construction were tested, and the design for the internal components of the gasifier was established.

In addition to the pilot plant activities, support studies have been conducted on 1) ash chemistry to better understand the principle of ash agglomeration, 2) bench-scale tests to determine the main operating variables affecting the formation of ash agglomerates, 3) cold-flow model tests to define the mechanism of selective separation of agglomerates and obtain scale-up information, 4) computer modeling to predict the performance of the gasifier, and 5) combustion experiments to determine utilization characteristics of IFG.

DEMONSTRATION PLANT DESIGN

The Industrial Fuel Gas Demonstration Plant produces a nominal 50 billion Btu/day of product gas, which is equivalent in energy output to approximately that of a 10,000 barrel/day oil refinery. The overall plant balance is shown in Figure 3. The coal feed rate to the plant is 3158 ton/day of Western Kentucky No. 9 coal. The properties of the design coal are shown in Table 6. The product gas has a heating value of 300 ±30 Btu/SCF. Gas in the amount of 45 billion Btu/day is available as send-out gas to IFG customers. The remaining 5 billion Btu/day of this gas is further processed to pipeline quality (950 Btu/SCF) and deposited in the Memphis natural-gas distribution system to generate Btu credit. The Btu credit can be withdrawn and used to satisfy IFG customer demand when the U-GAS production facility is totally or partially down for maintenance. By the use of the credit generation system, the demand of IFG customers can thus be assured. The demonstration plant design has been prepared by FWEC.

Figure 4 is the plant block flow diagram showing the process sequence and process-related support facilities of this demonstration plant. Each process unit as well as each process-related support facility is described briefly in the following summary.



THERMAL EFFICIENCY = 69.3%

Figure 3. DEMONSTRATION PLANT OVERALL BALANCE

Table 6. PROPERTIES OF WESTERN KENTUCKY DESIGN COAL

Ultimate	As Received 2-inch X 0, Wt %
Moisture	11.0
Ash	12.0
Carbon	61.1
Hydrogen	4.3
Nitrogen	1.0
Chlorine	0.2
Sulfur	3.5
Oxygen	6.9
	100.0
Proximate	
Moisture	11.0
Ash	12.0
Volatiles	35.4
Fixed Carbon	41.6
	100.0
HCV Btu/lb	11,157
FSI	4-6

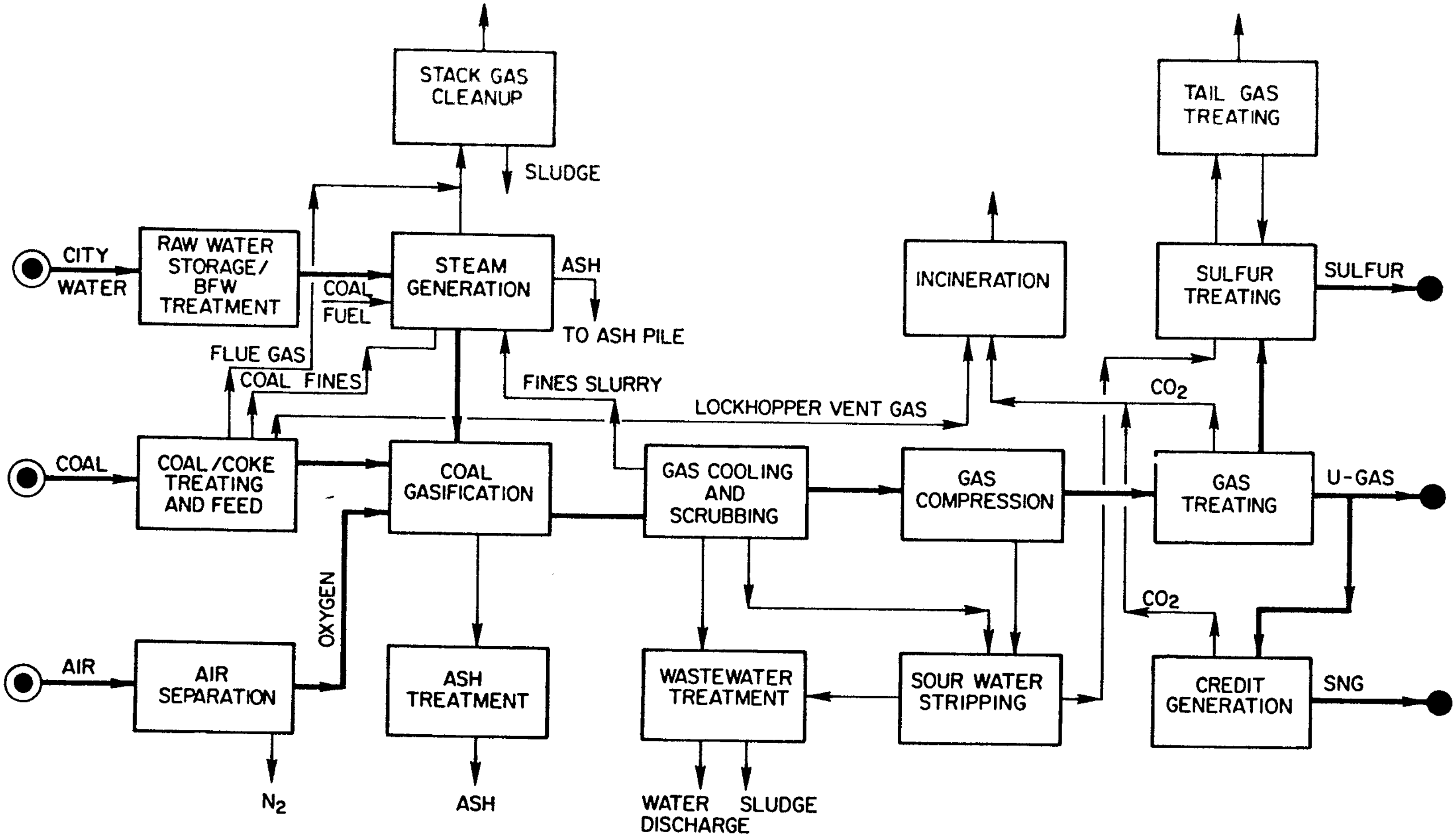


Figure 4. DEMONSTRATION PLANT FLOW DIAGRAM

- Air Separation Plant
  Compresses intake air and separates it into oxygen and nitrogen. The oxygen is compressed and sent to the gasifiers. A small portion of the nitrogen is returned for plant use. Liquid oxygen and nitrogen can also be produced to keep their respective storage tanks filled and thereby provide the necessary reserve for an outage of the air separation plant.

- Coal/Coke Treating and Feed
  Coal is crushed from 2 to 1/4-inch X 0 and dried to 2.5% moisture in a dryer mill. The dried, sized coal is stored in a coal silo. Sized coke received by the plant is also dried by a separate dryer and stored in a coke silo. Coal or coke is conveyed to the gasifier feeding systems from either the coal or coke silo. Dual conveying systems are provided to fill the gasifier feeding systems; one serves as a spare. Each gasifier has its own feeding system. The gasifier feeding system is a multi-feed hopper system, each consisting of a receiving hopper, two lockhoppers and two injection hoppers. Each injection hopper feeds into three pneumatic injection lines that transport coal or coke into the gasifier.

- Coal Gasification
  Contains the coal gasifiers where steam and oxygen react with the coal in a fluidized bed at about 1875°F and 75 psig to produce hot, raw gas (CO, CO_2 and H_2). Within the reaction zone of the fluidized bed is an ash-agglomerating zone. The ash agglomerates drop into a water quench. Fines carried over with the hot, raw gas are returned to the gasifier through external cyclones.

- Coal/Coke Handling
  Receives the incoming washed coal (2-inch X 0) from barges and transports it to a 14-day live-coal storage pile. From there coal is transported to Coal/Coke Treating and Feed.

- Ash Treatment
  Receives the agglomerated quenched ash slurry from the gasifiers and conveys it hydraulically to the dewatering bins. The dewatered ash is then discharged into trucks for disposal to the ash pile. The water from the dewatering bins is collected in the clarifier where clean water overflows into a sump tank while the underflow is pumped back to the dewatering bins. The clean water is then recycled to the gasifiers. A start-up pump is provided for initial transport of slurry to the dewatering bins when the gasifier pressure is too low for conveying.

- Gas Cooling and Scrubbing
  Cools the gas from 1875° to 450°F. For purposes of heat recovery, the gas passes in sequence through a high-pressure steam generator, high-pressure steam superheater, another high-pressure steam generator, and a boiler feed-water

preheater. After heat recovery the raw gas is quenched to saturation and passes through scrubbers. In the scrubbers particulate matter is removed by scrubbing with water. This section and Coal Gasification are four parallel trains and the balance of the plant is one train.

- Gas Compression
  Scrubbed gas is cooled, compressed to sufficiently high pressure and cooled again to go through gas treating and deliver the gas at 150 psig to the industrial fuel-gas distribution header.

- Gas Treating
  Receives the cooled gas from the Gas Compression Section. It then passes to a Selexol unit where H_2S and COS are removed to meet the product gas sulfur specification, and enough CO_2 is removed to obtain a constant-heating-value product gas. The product is then odorized and metered before being discharged to the industrial fuel gas distribution system.

- Sour Water Stripping
  Receives sour water mainly from the Gas Cooling and Scrubbing Section. The major portions of ammonia and hydrogen sulfide are removed by means of steam stripping.

- Sulfur Recovery
  Handles both sour gas and acid gas. It converts the sulfur compound in three catalytic stages of Claus-type sulfur recovery unit to achieve 96% recovery. Sulfur goes through condensers, seal pit and rundown pit, and storage tank before being loaded into tank trucks.

- Tail Gas Treating
  Receives the tail gas from Sulfur Recovery. It then goes to a Beavon unit package where remaining sulfur is converted to H_2S and then removed to a Stretford Unit. The tail gas is reheated to achieve satisfactory buoyancy and discharged to the atmosphere.

- Credit Generation
  Treats from 10% to 30% of the product gas to product pipeline-quality gas that will be deposited into the Memphis pipeline gas distribution system to generate a reserve of credit. This reserve can be withdrawn during U-GAS plant outage. Pipeline gas withdrawn from the Memphis pipeline gas distribution system will be adjusted to the U-GAS heating value prior to its distribution to the U-GAS customer.

The non-process sections to support the process and to provide utilities to the process include the following functions.

- Utility Area, which includes:
  Steam Generation
  Raw Water Storage
  BFW Treatment

- Waste Water Treatment

- Cooling Tower

- Flare

- General Facilities, which include:
  Long-Term Coal Storage for 90 days
  Long-Term Ash and Solid Waste Storage
  Interconnecting Piping
  Roads and Fences
  Firewater System
  Power, Lighting, and Communication
  Sewers

MARKETING

RELIABILITY

The main selling point of the Fuel Gas Demonstration Plant is the reliability and the assurance of supply. To increase the attractiveness of this fuel gas to potential industrial customers, the reliability of supply must be insured, even during periods of plant shutdown or repair and maintenance. The plant is designed to enhance reliability by its use of modular gasifier trains and several backup systems. For the present, reliability is of special concern because only one plant (rather than several independent plants, as would be the case for an already developed system) will be available to produce gas for customers.

Additional reliability will be obtained by using the existing natural gas system as backup and establishing a credit system. During normal operation, up to 10% of the product gas from the plant will be methanated to natural-gas quality or to synthetic pipeline gas and introduced into the existing Memphis natural gas system, thereby accruing "credit" against periods of time when the plant is not operating. During these periods the "credited natural gas" will be withdrawn, diluted with air to the proper medium-Btu heating value, and distributed to the industrial customer. In addition, when the plant output is in excess of demand, the difference will be injected into the credit generation unit and the excess methane sold to generate revenue. The credit system is sized to handle up to 30% of the plant output. Thus, besides providing reliability, the credit system also allows the IFG plant to operate at a steady-state full-load manner (high operating factor).

As part of the marketing effort, surveys on potential customers' burner systems have been conducted to evaluate their suitability for using the fuel gas. Based on these studies, 25 customers have been identified as potential users of the fuel gas. These customers have indicated interest in the purchase of gas at competitive prices. Also, two large industrial parks are now being planned in the vicinity of the plant site. Industries will be attracted to these parks because of the assured supply of fuel and will be potential users of the gas.

The cost of fuel gas produced in the demonstration plant is expected to be competitive with fuel oil and other alternate forms of energy.

COMMERCIAL PLANT ECONOMICS

During Phase I, a commercial plant conceptual design and a cost estimate were prepared by FWEC. The commercial plant is defined as a plant built using the experience gained from the construction and operation of the demonstration plant. Therefore, there are quite a few differences between the demonstration plant and the commercial plant design.

The commercial plant produces 50 billion Btu/day of industrial fuel gas from a total coal feed of 2792 tons/day of Western Kentucky No. 9 coal. Approximately 175 million SCF/day of product gas with a heating value of 330 $\pm$30 Btu/day is produced. Unlike the demonstration plant, the commercial plant does not have a credit generation system to produce pipeline gas, a separate pipeline, or a site requiring special preparations. Other major differences are use of product gas as boiler fuel, catalytic hydrolysis of carbonyl sulfide, sparing and backups philosophy, and gasifier carbon conversion efficiency in the commercial plant design.

Using the commercial plant conceptaul design, erected plant cost-estimates were prepared on a process unit basis. Costs were obtained both from process licensors and vendors whenever possible. Other costs were based on FWEC in-house information. The economic analysis and calculation of gas costs presented here are based on C. F. Braun Utility Financing Method. The total capital replacement is estimated to be $197.4 million, expressed in Fourth Quarter, 1979 dollars. The breakdown is shown in Table 6. The annual operating cost based on a 20-year plant life and 90% stream factor is $45.29 million. Table 7 shows the itemized operating costs. Using the utility financing method, the average cost of gas is $4.25 per million Btu.

MM/mdc/ph

Table 6. COMMERCIAL PLANT CAPITAL REQUIREMENT
(Fourth Quarter, 1979 Dollars)

	$ Million
Erected Plant Cost	129.2
Contractor's Charges	21.3
Start-Up Costs	9.1
Working Capital	8.4
Interest during Construction	29.4
Total Capital Requirement	197.4

Table 7. ANNUAL OPERATING COSTS
(90% Stream Factor)

	$1000/yr
Coal at $26/ton ($1.18/MM Btu)	$ 23,863
Catalyst and Chemicals	225
Water ($.25/100 ft^3)	194
Electricity (2.5¢/kWh)	8,560
Operating Labor and Supplies	5,847
Maintenance, Labor and Supplies	4,977
Insurance and Taxes	3,900
Gross Operating Costs	$ 47,566
Less By-Products	2,279
Net Operating Costs	$ 45,287
Annual Gas Production, 10^{12} Btu/yr	15.93
Average Gas Cost*, $/million Btu	4.25

*Based on C. F. Braun Utility Financing Method

WINKLER FLUID BED COAL GASIFICATION EXPERIENCE WITH LOW GRADE COALS

Dr. F. Boegner

T. J. Kendron

T. K. Subramanian

WINKLER FLUID BED COAL GASIFICATION EXPERIENCE WITH LOW GRADE COALS

Dr. F. Boegner
Davy International AG
Cologne, West Germany

T. J. Kendron
T. K. Subramanian
Davy McKee Corporation
Cleveland, Ohio USA

Abstract

One of the proven commercial systems in the gasification of carbonaceous material is the Winkler fluid bed process.[1] This process has been commercially used in the following production of:

- Low and medium Btu industrial gas
- Synthesis gas for ammonia, methanol and oxo-alcohols
- Fisher-Tropsch synthesis gas
- Reducing gas
- Hydrogen

The Winkler fluid bed process is well suited for more reactive, young carbonaceous materials like peat, lignite and subbituminous coals which are available in abundance around the world.

Low grade coals with a high percentage of fines in the range of 0-3/8", high ash content up to 50%, and high sulfur content have been successfully gasified in the commercial Winkler gasifiers. These gasifiers have performed with high efficiency, especially when the ash fusion temperature is in the range of 2000-3000^{o}F as found in the majority of U.S. coals.

For commercial plants it should be remembered that, though the economy of total production depends upon gasification efficiency which varies from process to process only in a small range, it is very important to have high performance and reliability in the gasification process and related equipment.

The paper describes how the Winkler fluid bed gasification has been improved and simplified in both operation and maintenance for commercial plants.

Winkler Fluid Bed

This presentation deals with one of the most reliable, large scale, commercially proven gasifiers - the Winkler fluid bed gasifier. Table 1 shows the list of commercial Winkler plants that have been successfully operated outside the U.S.

Proven Technology With High Stream Factor

A proven gasification process must be defined as the one that has maintained many years of successful operation. Such a process is Winkler fluid bed gasification. One of the several Winkler plants now in operation has in its 20 years of production successfully operated on a wide range of coal qualities and with ash in raw coal as high as 50%. This Winkler plant was designed for 900,000 scfh (24,000 Nm^3/Hr) total capacity, supplying synthesis gas for the Azot fertilizer complex in Kutahya, Turkey. This plant has an extremely high on-stream factor and reliability in the gasification section. In general, the economy of such a plant depends not merely on the gasification efficiency. The gasifier performance and its on-stream factor play an important role in the overall economy of the plant. Fig. 1 shows the effect of lower annual on-stream factor on the total plant production cost. As can be seen, the production costs increase quite rapidly when the plant is below 90% utilization.

SEE PAGE 227 FOR FIGURE 1.

The Kutahya Plant has a high on-stream factor without a spare unit. It has also produced, when downstream requirements dictated, as high as 130% of original design capacity without substantial loss of overall conversion efficiency.

The reliability of this plant is due to the continued development of technological ideas, engineering capabilities, and skilled manufacturing and operating experience.

Types of Feedstocks to the Winkler Gasifier

In principle, any carbonaceous material from wood to graphite can be gasified. Geologically young feedstocks such as wood, peat, lignite with the higher reactivities indicated by higher oxygen and lower carbon content, are in general more appropriate to Winkler gasification than elder bituminous and anthracite coals.

The first indication for successful gasification is the coal classification triaxial (O-H-C) diagram of Grout-Apfelbeck, Fig. 2.

Less reactivity can be compensated for by modifying certain operating conditions, mainly higher temperatures. The operating temperatures are influenced however by ash in the coal. The ash in the "low grade coals" can be as high as 50% and if the ash is intimately bound up with the carbon, a most effective and economic way of gasification is in a fluid bed without wasting energy for superfluous melting of half of the feedstock. The fluid bed operation allows the best mass and heat transfer between the reactants.

The molten fly ash particles are a serious problem in most gasification processes. In the Winkler process, by the application of a radiant boiler/cooler in the upper part of the gasifier, it is possible to maintain the temperature in this zone about 200-400°F higher than in the fluid bed. This prevents molten ash leaving the gasifier. The effect of the elevated temperature in the upper part of the gasifier also results in an increased carbon conversion efficiency of 5-10%.

Simplified Operation and Maintenance

The capability of the fluidized bed Winkler process for gasifying a wide size range of coal directly is very significant as more and more fines are being produced by the highly mechanical mining methods. Normally run-of-mine coal only requires crushing before being fed to the Winkler gasifier. In the early stages of design, the fluidization was carried out above a supporting slotted grate. The granular part of the gasification residue which fell to the grate was discharged by a stirring arm. The first major improvement in the Winkler gasifier was the elimination of such internal moving parts. The water-cooled, mechanically-swept grate was replaced by a refractory-lined cone, and the fluidizing media are injected by means of nozzles mounted around the circumference of the cone at various levels.[5, 6] This dramatically increased the operational reliability and simplicity of the Winkler gasifier. The conical configuration of the fluidized bed ensures uniform fluidization of the whole bed independent of the size of the gasifier. Fluidization starts at the bottom of the cone and additional amounts of fluidizing and gasifying media are injected successively along the height of the cone. Following this improvement in the design, the problems of slagging, gas channelling, and wide temperature

Advantages of Winkler Fluid Bed Gasifier

The main areas where the Winkler process shows its advantages are:

- o Environmental impact
- o Coal feed preparation
- o Ease and flexibility of operation
- o Gas quality

Environmental Impact In the gasification of coal on a commercial scale, the impact on the local environment is of major concern. In this respect, the Winkler process has much to offer:

- Instantaneous gasification of coal produces a raw gas free of tars, oils and higher hydrocarbons.

- Raw gas contains only parts per million of contaminants such as NH_3 and HCN.

- Liquid waste stream can be processed at the site with moderate costs using existing technology.

Coal Feed Preparation Run-of-mine coal received at the plant site requires minimal treatment before gasification.

- Drying of the coal is normally not necessary. Drying may be recommended when the surface moisture in coal is high.

- Pulverizing of the coal is not required. Run-of-mine coal in the size of 0-10 mm including all the fines is a suitable feedstock. All of the received coal can be utilized in the gasifier. No specific size classifying is required.

Ease and Flexibility of Operation The simplicity of the Winkler gasifier has the inherent quality of trouble free, reliable operation.

- The gasifier is a vertical, cylindrically shaped, refractory lined empty vessel without internal parts to initiate plugging.

- Fluctuations in the ash content of the coal do not upset the operation.

- The gasifier is capable of operation with a turndown ratio of 1:4. For example, an atmospheric, oxygen blown Winkler gasifier with a cross-sectional area of 270 sq. ft. (25 sq. meter) designed for 2,240,000 scfh ($60,000\ Nm^3/Hr$) can be turned down to 675,000 scfh ($18,000\ Nm^3/Hr$) or overloaded to 2,800,000 scfh ($75,000\ Nm^3/Hr$) without a considerable effect on the economics of gasification. The capacity is limited at the lower end by the minimum flow required for fluidization and at the upper end by the minimum residence time for substantial gasification of residues. This great flexibility can provide necessary reserves in case of a shutdown of one unit in a multitrain facility.

The normal capacity of a 270 sq. ft. cross-sectional gasifier is 2,240,000 scfh ($60,000\ Nm^3/Hr$) when blown with oxygen; and 3,730,000 scfh ($100,000\ Nm^3/Hr$) when blown with air. This high capacity per unit can still be increased at elevated pressure.[2] Making use of proven pressurized feeding systems, Davy McKee guarantees, at present, the gasifier operation up to 4 bar.[3]

- Long residence time promotes safety by preventing oxygen breakthroughs. The gasifier has a relatively large inventory in the fluidized state that allows safety in the event of interruption of coal feed.

- Reliable operation; better than 90% on-stream time over 20 years at the presently operating Kutahya plant.

- It is easy to start-up and shutdown.

- Oxygen or air can be used as the gasification oxidant.

Gas Quality The Winkler gasifier produces mainly CO, CO_2, H_2, CH_4 and sulfides. No higher hydrocarbons are generated. The gasifier produces favorable gas quality for the synthesis gas generation for chemicals. Unlike other commercially proven gasification processes, the oxygen blown Winkler gasifiers have had generated 82% CO + H_2 and 14% CO_2. The produced raw gas has $CO:H_2$ ratio of about 1:1. This indicates that a major part of CO-shift conversion takes place in the gasifier itself and subsequent $CO:H_2$ ratio adjustment for synthesis gas generation is lower than other processes.[2]

Process Description

A typical process scheme of a Winkler fluid bed coal gasification plant is shown in Fig. 3.

Coal preparation is by the simplest methods with regard to the requirements of Winkler fluid bed operation. Run-of-mine coal in the size of 0-10 mm including all the fines is the best suitable feedstock.

Drying is only required to insure adequate coal flowability and economic oxidant utilization. For certain moisture levels, drying coal in the gasifier with oxygen is more expensive than an external drying system.

Coal to be gasified is fed to the feed hopper, 1, from which a screw conveyor, 2, charges it into the lower part of the generator, 3. This section is a brick-lined cylindrical shaft with a conical bottom, in which fluidization of the coal is achieved. This area is equipped with nozzles, 4, for the injection of preheated air or oxygen and steam, which act as a medium for fluidizing the fine granular coal, subjecting the whole fuel bed to turbulent motion. Due to the high turbulence, which can be controlled by the velocity of injection, the particles of freshly fed coal are immediately mixed with the bulk of devolatilized, high ash inventory of the fluidized bed.[7] This also makes it possible to gasify moderately caking coals. The rapid mixing of fresh coal with the contents of gasifier due to turbulent motion of the fluidized bed corresponds to a diluent effect weakening the caking properties. The tendency of particles to cake together or of ash to sinter on the walls or to form clinkers can be overcome by increasing the intensity of fluidization.

Coals with a DAM index up to 10 have been handled without any problem. In the case of coals with more caking characteristics, other feeding systems (such as pneumatic) could be applied.

Air or oxygen mixture is also injected into the space above the fluidized bed. This allows more complete gasification of coal particles entrained in the upward flow of gaseous products using a higher temperature and the effect of the radiant boiler/cooler, as described previously. At these temperatures reforming of volatile matter in the fuel is complete, and no tar or liquid by-products, which would cause environmental problems, are produced.

Larger ash particles fall to the bottom of the generator and are removed by a cooled discharge screw, 5.

The considerable amount of sensible heat in the gas, which contains carryover fly ash from the feedstock, is recovered in a train of waste heat boilers, 6, including radiant boiler/cooler tubes along the wall of the generator, with steam superheater and boiler feed water heater.

In addition to steam for internal process consumption, an excess of steam is produced, suitable for steam turbine drives.

Following heat recovery the gas passes through a cyclone, 7, in which most of the solid particles, in the form of a fine, granular ash, are separated out and discharged by a rotary air lock to a screw conveyor.

Trials to recycle the fly ash or to gasify the fly ash as well as the bottom ash in a separate, second gasifier did improve the carbon conversion by a few percent, but the performance of the gasifier was lowered mainly by the operability of the hot cyclone.[8] Therefore this solution was substituted by a simplified method which improves the economy of the total plant and useful carbon conversion by approximately 5-10 percent. The discharged fly ash which contains combustible material and possesses generally sufficient ignition and firing characteristics has been used as a supplementary fuel for utility steam boilers.

Residual ash removal after the cyclone is accomplished in a scrubber, 8, the water from which is recycled via a settling vessel, 9. Even this small portion of fine fly ash which contains some combustible material may be used in an auxiliary boiler.

Elevated pressure operation allows the use of a venturi scrubber system, which is more effective than the former Theisen disintegrator system.

The sulfur contained in the fuel is converted mainly to hydrogen sulfide, but small amounts of COS are also present. The gas contains the same amounts of nitrogen and argon as are carried over by the air/oxygen. As formation of NO_x depends on the operating temperature, it is lower than the processes which are operated at temperatures above the ash fusion temperature.

Operating Data of Kutahya Plant

The Winkler fluid bed gasifiers at Kutahya, Turkey, in the AZOT fertilizer plant, have been successfully operated for more than 20 years with high reliability and low maintenance. Total design capacity is 900,000 scfh (24,000 Nm^3/Hr) of raw gas produced by two (2) Winkler gasifiers. Without spare gasification equipment, the plant has operated at 130% of its designed capacity with feedstocks containing up to 50% ash. Typical product data of this plant are shown in Table 2.

The lowest capacity the gasifiers are normally operated is 12,000 Nm^3/hr which is a turndown of approximately 50%. This turndown is limited by minimum flow to the compression unit but turndown in other plants has been much greater.

With such a range of gas production in this commercial plant, the Winkler gasifier has proven the versatility in capacity without any appreciable loss in efficiency.

Maintenance requirements in this plant have been confirmed to be low. The original gasifier refractory lining and waste heat boilers are still in operation. The ammonia plant has not been shutdown due to loss of the gasification plant.

For reference, the Winkler fluid bed gasification plant at Kutahya has averaged a maintenance charge of 2.25% per year on operating capital for material and labor. The operating people feel that the Winkler gasification process is simple to operate and it is reliable when compared to the other syngas systems available to them.

Summary of Winkler Coal Gasification Applications

The Winkler fluid bed coal gasification process has been commercially applied in more than 60 units and is suitable to produce:

- Fuel gas, low and medium Btu gases for industrial uses[9]
- Synthesis gas for chemical plants, Fischer-Tropsch synthesis and hydrogen plants[4]
- Reducing gas[10]

The advantage the Winkler process offers for utilization in the abovementioned applications is a raw gas quality that is nearest the end product requirements. For low-Btu gas with a heating value of 120 Btu/scf, the plant investment costs are minimized as an air blown unit is used. The medium Btu gas or synthesis gas can be produced with the heating value of 280 Btu/scf when oxygen is used as the oxidizing medium.

The raw gas contains no tars, oils and other higher hydrocarbons. Hence the liquid waste stream can be processed at the plant site for moderate costs using existing technology. The quality of the gas made is such that the $CO:H_2$ ratio is about 1:1 and hence the subsequent synthesis gas preparation for CO and H_2 adjustment, using the CO-shift conversion, will result in the minimization of carbon loss as CO_2.

The plant by-product char contains combustible material and possesses sufficient ignition and firing characteristics such that it can be used in the offsite steam boilers as demonstrated in the Kutahya plant. Thus, the full utilization of carbon in the feedstock results in high overall efficiency. It has to be pointed out again, that an accurate process comparison should not stay within the gasification itself. One should consider the total process system.

Davy McKee Experience

Gasification of carbonaceous materials is a proven technology for over 100 years. Davy McKee is by its long tradition one of the most experienced engineering companies in the world.

In 1901 Powergas Corporation, now Davy McKee, was founded for the sole purpose of marketing a gas producer that operated on coke, various types of coal, and even wood and waste biomass material. By the acquisition of R. T. Mathews Co. Ltd., Davy McKee has enriched its experience with another moving or fixed bed system, a two-stage gas producer.

Pintsch KG and Bamag (forerunners to Bamag Chemietechnik) which is also part of Davy McKee, had established themselves in coal gasification on a similar scale by building several hundred gasifiers of different types; including Pintsch Hillebrand gasifiers, Leuna type slagging gasifiers, rotary grate water gas and Winkler fluid bed gasifiers.

Literature Cited

1) Anwer, J., Boegner, F., and Roth, H., "Industrial utilization of young fossil fuels by gasification in fluidized bed." International IMTG Symposium, Bad Zwischenahn, November, 1979.

2) Anwer, J. and Boegner, F., "Kohlevergasung im Fluidatbett (Winkler Vergasung) unter Druck - Gasification of Coal in a Fluidized Bed under Pressure-," Brennstoff-Warme-Kraft (2/1976).

3) Davy McKee Corporation brochure "Winkler fluidized bed coal gasification."

4) Anwer, J., Boegner, F., Bailey, E. and Subramaniam, T. K., "Winkler coal gasification process and its application in ammonia synthesis plants," Ammonia from Coal TVA Symposium, Muscle Shoals, May, 1979.

5) Sabel, F., "Die Winklersche Wirbelschichttechnik - the Winkler fluidized-bed technique-," Chem. Ing. Technik (2/1952).

6) Bios final report No. 199 (1946) "Ten years of oxygen gasification at Leuna."

7) Flesch, W. and Yagneswaran, S., "New experiences with Winkler coal gasification at Neyveli," Chemical Age of India (9/1976).

8) Portatius, B., "Neuere Entwicklung der Vergasungstechnik nach Winkler - New developments of Winkler gasification-," Freiberger Forschungshefte No. 69 (1957).

9) Bailey, E. and Goodman, M., "Synthetic medium Btu gas via Winkler process," the Fourth Annual International Conference on Coal Gasification, Liquefaction and Conversion to Electricity, Pittsburgh, August, 1977.

10) Anwer, J. and Boegner, F., "New application of coal gasification (Winkler process) for direct reduction of iron ores," The Indian Iron and Steel Annual, July, 1976.

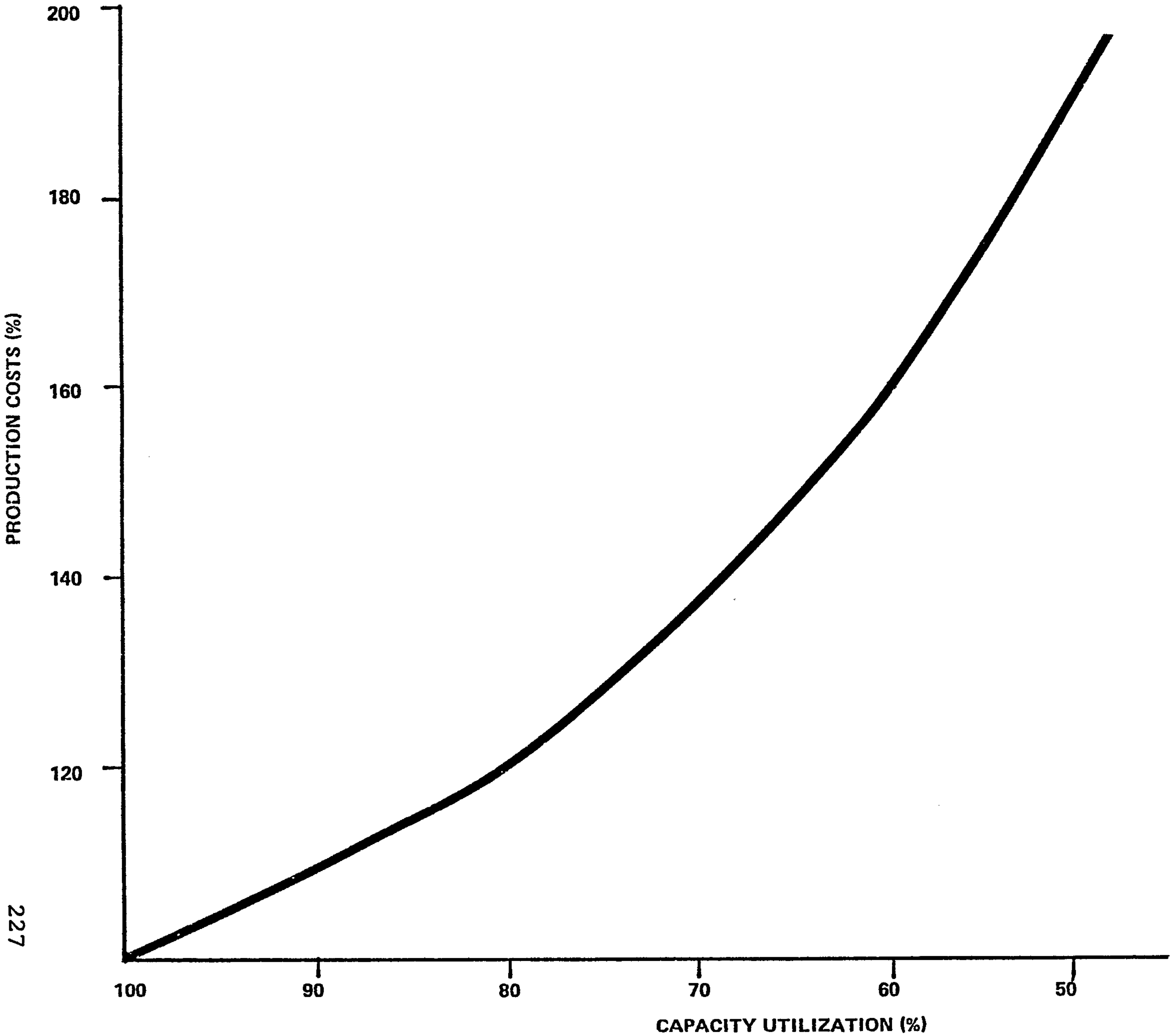
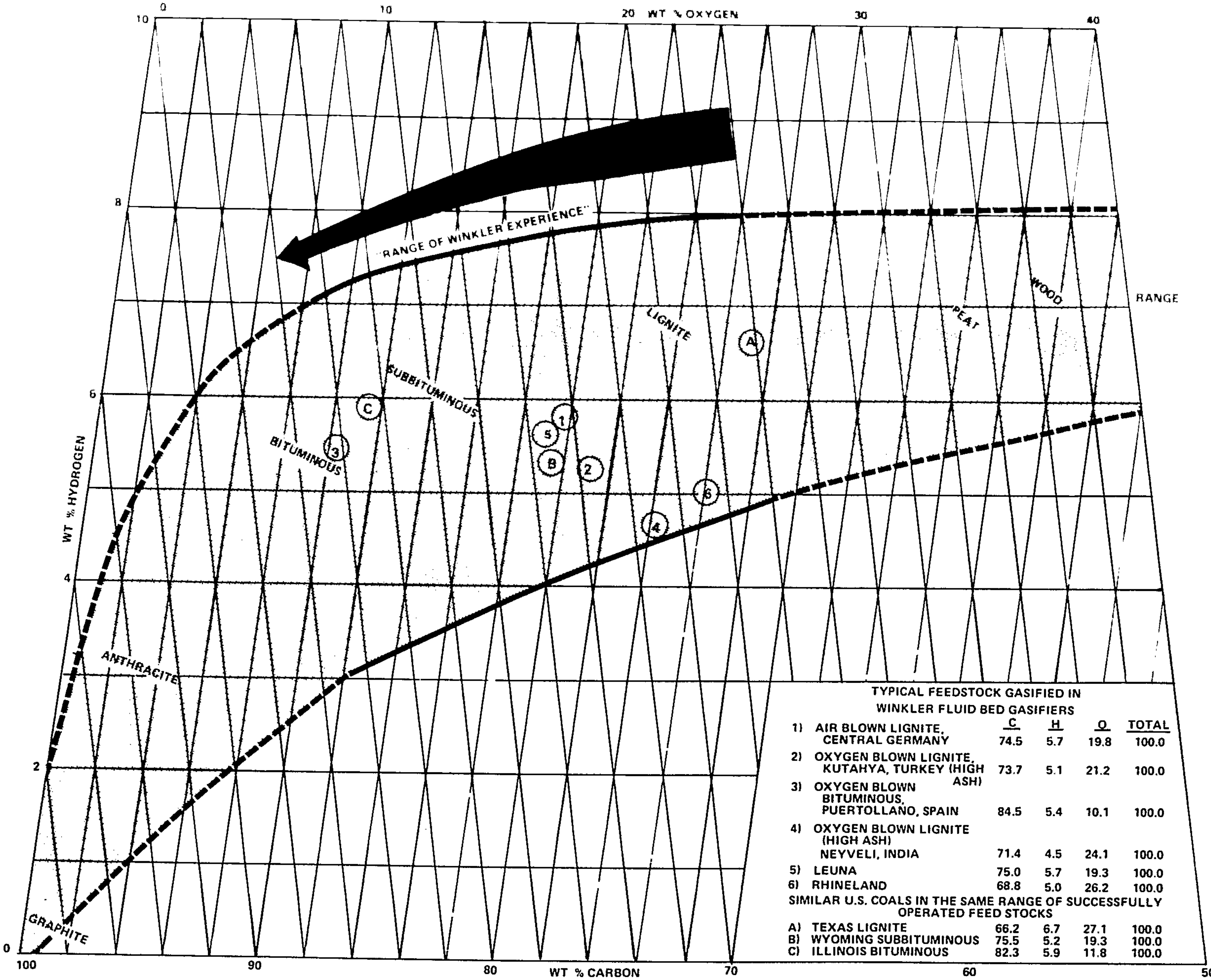


FIGURE 1: RELATIONSHIP OF OPERATING TIME ON PRODUCTION COSTS



TYPICAL FEEDSTOCK GASIFIED IN WINKLER FLUID BED GASIFIERS

	C	H	O	TOTAL
1) AIR BLOWN LIGNITE, CENTRAL GERMANY	74.5	5.7	19.8	100.0
2) OXYGEN BLOWN LIGNITE, KUTAHYA, TURKEY (HIGH ASH)	73.7	5.1	21.2	100.0
3) OXYGEN BLOWN BITUMINOUS, PUERTOLLANO, SPAIN	84.5	5.4	10.1	100.0
4) OXYGEN BLOWN LIGNITE (HIGH ASH) NEYVELI, INDIA	71.4	4.5	24.1	100.0
5) LEUNA	75.0	5.7	19.3	100.0
6) RHINELAND	68.8	5.0	26.2	100.0
SIMILAR U.S. COALS IN THE SAME RANGE OF SUCCESSFULLY OPERATED FEED STOCKS				
A) TEXAS LIGNITE	66.2	6.7	27.1	100.0
B) WYOMING SUBBITUMINOUS	75.5	5.2	19.3	100.0
C) ILLINOIS BITUMINOUS	82.3	5.9	11.8	100.0

FIG. 2 RANGE OF WINKLER GASIFIER EXPERIENCE
GROUT APFELBECK TRIAXIAL DIAGRAM

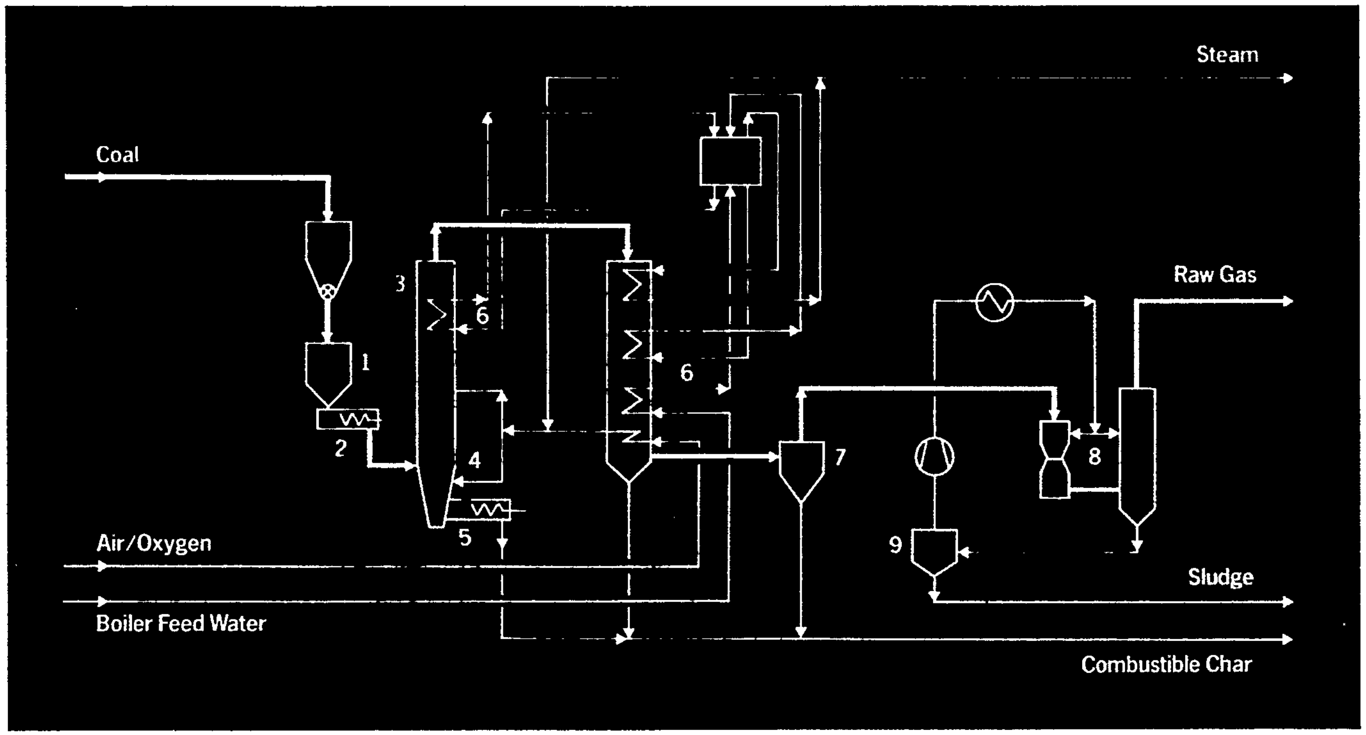


FIG. 3 TYPICAL WINKLER PROCESS SCHEME

X– TABLE 1: PLANT LIST OF WINKLER

ANT NO.	PLANT	IN OPERATION	FEEDSTOCK	FINAL PRODUCT
1	BASF, LUDWIGSHAVEN, GERMANY	1925–1958	LIGNITES/ BIT. COAL	TEST PLANT
2	LEUNA–WERKE MERSEBURG, GERMANY	1928–1970	BROWN COAL, SALT COAL, COKE	AMMONIA METHANOL, LOW BTU GAS
3	BRABAG, BOHLEN, GERMANY	1938– PRESENT	LOW TEMP COKE	HYDROGEN FOR SYN. FUEL
4	BRABAG, MAGDEBURG, GERMANY	1938–1945	LOW TEMP COKE	HYDROGEN FOR SYN. FUEL
5	YAHAGI, JAPAN	1937–1960	SEMI–COKE	AMMONIA
6	BRABAG, ZEITZ, GERMANY	1941– PRESENT	LOW TEMP COKE	HYDROGEN FOR TTH–FUEL
7	DAI–NIHONYINZO–HIRYO, JAPAN	1937–1959	CAKING BIT COAL	AMMONIA
8	NIPPON TAR, JAPAN	1937–1960	CAKING BIT COAL	AMMONIA
9	TOYO–KOATSU, JAPAN	1938–1969	CAKING BIT COAL	AMMONIA
10*	FUSHUN, MANDSCHUKUO (MANCHURIA)	1939–?	CAKING BIT COAL	SYN GAS FOR FISCHER TROPSCH
11	TREIBSTOFFWERKE, BRUX, CZECHOSLOVAKIA	1943–1972 1954–1973	LOW TEMP COKE OF LIGNITE	HYDROGEN FOR FUEL
12*	SALAWAD, USSR	1950– PRESENT	LIGNITE	WATER GAS
13*	BASCHKIRIEN, USSR	1950– PRESENT	LIGNITE	WATER GAS
14*	KOMBINAT SCHWARZE PUMPE	1950– PRESENT	LIGNITE	LBTU GAS
5*	DIMITROFFGRAD, BULGARIA	1950– PRESENT	LIGNITE	WATER GAS
6*	STARA ZAGORA, BULGARIA	1962– PRESENT	LIGNITE	WATER GAS
7	FABRIKA AZOTNIH, GORAZDE, YUGOSLAVIA	1953– PRESENT	LIGNITE	AMMONIA
8	CALVO SOTELO I PUEROTOLLANO, SPAIN CALVO SATELO II	1956–1970 1959–1970	CAKING BIT. COAL	AMMONIA
9	UK -WESSELING I, GERMANY UK--WESSELING II	1958–1967 1962–1967	LIGNITE LIGNITE	AMMONIA/ METHANOL
0	AZOT SANAYII TAS, KUTAHYA, TURKEY	1959– PRESENT	HIGH ASH LIGNITE	AMMONIA
1	NEYVELI LIGNITE CORP, INDIA	1965–1978	LIGNITE	AMMONIA

E: All Winkler gasifiers operated at atmospheric pressure
built by Bamag Chemitechnic (Presently Davy-McKee)
ilt by others.

FLUID BED GASIFIERS

X+1-TABLE 1

CAPACITY PER GENERATOR NORMAL 1000 nM^3/HR	1000 SCFH	MAXIMUM 1000 nM^3/HR	1000 SCFH	NUMBER OF GASIFIERS
2	75	–	–	1
60 100	2240 3730	100 –	3730 –	5
27.6	1030	30	1230	3
27.6	1030	33	1230	3
9	330	–	–	1
23	860	–	–	3
14	520	–	–	2
14	520	–	–	2
15	560	20	750	2
15	560	20	750	4
26 32	1000 1200	30 –	1120 –	5 2
23	860	–	–	7
23	860	–	–	4
65	2400	–	–	6
18	670	–	–	4
30	1120	–	–	5
5	190	7	260	1
9.5 9.5	350 350	– –	– –	1 1
12 12	450 450	17 17	630 630	1 1
12	450	18	670	2
17	630	21	785	3

TOTAL GASIFIERS

TABLE 2
TYPICAL KUTAHYA PLANT WINKLER GASIFICATION SECTION

Gasifiers	
Number of Gasifiers	2
Gasifier Size, Dimensions	3 meter ID x 21 meter high
Type	Oxygen blown
Coal	
Coal to Gasifiers	Dried, matured brown coal
Analysis	Wt. %
C	38 - 52
H	3.5 - 4.5
O	15 - 18
N	0.5 - 1.5
S	0.8 - 3.0
Ash	20 - 50
Water	4 - 8
LHV	3600-4600 kcal/kg (6480-8280 Btu/lb)
Raw Gas Generation	
Design Capacity	24,000 Nm^3/Hr raw gas
Analysis	Vol %
H_2	39 - 43
CO	31 - 36
CO_2	19 - 21
CH_4	2.4 - 2.9
$N_2 + A_2$	1.5 - 3.0
$H_2S + COS$	0.2 - 0.8
Consumption Figures (Per Nm^3 raw gas generated)	
Coal (34% ash & 4% H_2O), kg	0.9
Oxygen, Nm^3	0.26
Electricity, kw	0.08
Net Heat Recovery (Per Nm^3 raw gas generated)	
Waste Heat Boiler Net Steam Generation, kcal	435
Offsite Steam Generated by Combustible Char, kcal	640
Total Heat Recovered	1075
[LHV of Combustible Char	2400 cal/gm (4320 Btu/lb)]
Total Thermal Efficiency, %	81.0
Gasifier Performance Data	
Total Plant Design Capacity	24,000 Nm^3/Hr raw gas
Maximum Capacity Achieved	36,000 Nm^3/Hr raw gas
Lowest Production Rate	12,000 Nm^3/Hr raw gas

FLUIDIZED BED COMBUSTION
AN OVERVIEW OF DESIGN AND OPERATIONS

Robert M. Patterson

FLUIDIZED BED COMBUSTION

AN OVERVIEW OF DESIGN AND OPERATIONS

Robert M. Patterson

Davy McKee Corporation
Cleveland, Ohio
USA

INTRODUCTION

Fluidized bed combustion (FBC) is an environmentally sound and energy efficient approach to burning a wide variety of fuels. With decreasing and unsure supplies of liquid and gaseous fuels, FBC has lately received attention as a commercial solid fuel combustion technology.

This paper presents an overview of state-of-the-art of fluidized bed combustion technology. The FBC process is described in the first section. Following this, commercial suppliers (or potential suppliers) of FBC systems are listed, and the design parameters of available product lines are briefly reviewed. Data from operating units is then presented; and, finally, the economics of FBC versus competing technologies is reviewed.

THE FBC PROCESS

Fluidized bed processes have been in use in industry for over half a century. Applications for fluid beds include calcining, drying, catalytic cracking, and gasification. Fluidized bed combustion has come into the limelight over the last few years with the intent to use the process advantages of a fluid bed for direct fuel combustion.

A bed of solid particles, supported by a porous or perforated plate, may be transformed into a pseudo-fluid state by a stream of air passing upward through the plate. The air velocity must be sufficiently great to counteract gravitational forces on the particles, but not so great as to transport the entire bed along with the air stream. Under these conditions, the bed of solid particles behaves as a highly turbulent, boiling fluid - hence the term "fluidized-bed."

If a bed of particulate matter is fluidized with air and heated with an auxiliary fuel (oil or gas) to about 1000°F, almost any solid, liquid, or gaseous fuel will ignite. The fuel will be quickly distributed throughout the bed, and controlled combustion at a relatively low temperature may be attained.

The theoretical advantages of fluidized-bed combustion are:

- Utilization of high sulfur fuels without flue gas treatment
- Reduced combustion temperature
- Reduced NO_x emissions
- Reduced excess air requirements
- Reduction of combustor size
- Increased heat transfer to the working fluid
- Fuel versatility
- Compact modular unit, package construction of units
- Easy-to-handle by-product material

Generally, these advantages are a result of (1) intense turbulence within the fluidized bed and (2) long particle residence time in the bed without a correspondingly long linear flow path requirement. A wide range of fuels may be used: gases, fuel oils, coal of all ranks, coal mine wastes, organic wastes, to name several. In theory, the fluidized-bed combustor will operate on any fuel which has a net calorific value greater than its sensible heat at the bed temperature[1].

Perhaps the outstanding advantage of FBC is the ability to burn high sulfur coals in an environmentally acceptable manner without the need for auxiliary SO_2 control equipment. This is accomplished by fluidizing the coal with air and adding a sulfur sorbent, commonly a calcium containing mineral such as limestone or dolomite. The sulfur from the coal is captured by the calcium to form calcium sulfate ($CaSO_4$). This is a solid material which is readily removed with the ash. The sulfur, whether chemically bonded to or loosely associated with the coal, is efficiently removed.

The relatively low combustion temperature of the FBC process, 1400°F to 1700°F, as opposed to open flame temperatures of approximately 3000°F for conventional combustion, results in several other advantages. One is that at these low temperatures and low excess air requirements, NO_x formation is minimal and well within most regulatory emission requirements[2]. Another, is that the FBC temperature is generally lower than the coal ash fusion temperature. This prevents or minimizes slagging and bonded-deposit ash formation in the boiler, which reduces or eliminates the need for soot blowing[3]. The low temperature also results in a clinker-free, granular, smooth flowing ash which is easily handled, as opposed to the wet sludge produced by flue gas scrubbers. The ash may be easily disposed for landfill uses or potentially solid for industrial or agricultural applications.

The fluidized-bed systems described here are all domestic atmospheric pressure systems in contrast to pressurized FBC systems which utilize compressed air and a pressurized combustor to attain greater heat release rates. The advantages of pressurized fluidized-bed combustors are most appropriate for utility applications. The pressurized fluidized-bed combustors, still in the development stage, will be considerably more expensive than atmospheric units. Therefore, they will not be generally useful for industrial users of coal in the foreseeable future[4].

COMMERCIAL FBC SYSTEMS

There are numerous industrial boiler designers and suppliers looking seriously at fluidized bed combustion as a marketable technology. Many of these firms are currently offering FBC systems commercially, and several are working towards commercialization through research and development activities.

This section is intended to present the design and commercial status of each industrial FBC boiler supplier. The information given here should not be considered definitive; the FBC marketplace is rapidly changing as commercialization of the technology proceeds. It should be expected that the number of FBC suppliers and their product lines will change over the next few years.

Further, it is difficult to present all the pertinent technical data on fluidized bed combustion systems in an overview paper such as this. This information merely highlights each supplier's product line or projects to give some indication as to the status of FBC in industry.

Table I lists those boiler suppliers offering or developing FBC product lines. Certain other companies are considering licensing or developing their own FBC systems. Industrial Boiler Co. (Thomasville, GA) and Cleaver Brooks (Milwaukee, WI) are two such companies. This table also presents some key design parameters of FBC demonstration units and commercial product lines.

All FBC boilers require support systems for raw material handling and storage, water treatment, flue gas handling, ash handling and storage, and instrumentation. It is not the intent of this paper to discuss these subsystems, which for the most part are standard designs. The crux of an FBC steam generating plant is the boiler design. This paper stresses boiler design characteristics and operating parameters.

Babcock Contractors/Riley Stoker

Babcock Contractors (BCI) has formed a joint venture agreement with Riley Stoker to manufacture and market a line of industrial size FBC units.

Babcock Contractors Ltd. (U.K.) has four years of operating experience on a 45,000 lb/hr FBC retrofit boiler in Renfrew, Scotland. This unit is a cross-drum forced circulation boiler. The coal feed is sized to minus 1/4 inch, mixed with limestone and fed pneumatically to the boiler.

In addition, BCI has just completed construction on a 60,000 lb/hr FBC retrofit boiler at the Ohio Psychiatric Hospital in Columbus, Ohio. This facility is in the final stages of shakedown, with start-up imminent.

BCI/Riley is currently offering a line industrial-sized FBC boilers in capacities from 50,000 to 500,000 lb/hr, with steam conditions to 1000°F and 1600 psi. These units employ pneumatic underbed feed system which injects a fuel/limestone mix into a turbulent preheated bed. Operating bed temperatures are about 1550°F, which provides for adequate carbon burnup, good sulfur capture, reduced NO_x emissions, and no slagging.

Battelle/Struthers-Wells

In 1973, Battelle Columbus Laboratories began work on developing improved methods of fluidized-bed combustion. Out of these efforts evolved the Multi-Solid Fluidized Bed Combusiton (MSFBC) system.

This process features a tall vertical combustor, coupled with a primary separator and an external boiler. The combustion process and sulfur capture occur in the combustion chamber. Fluidizing air is blown up from under the combustor at about 30 feet per second. The bulk of the solids carried out of the combustor are collected in a cyclone and discharged to an external boiler. Here the hot material transfers heat to a process fluid (water in the case of a steam generator). The external heat exchanger is fluidized at less than 2 feet per second. Bed level in the external boiler is controlled through a variable speed valve in the discharge line to the main combustor.

Struthers-Wells Corporation has licensed the Battelle MSFBC technology to manufacture fluidized bed steam generators for oil field application. Specifically, these units generate high pressure steam for injection into existing oil wells to draw up heavy crudes.

The line of steam generators offered is in the capacity range from 50,000 to 250,000 lb steam/hr. Steam conditions are 1500 psi to 1650 psi, with steam quality typically at 80%. Since a steam generator for oil field flooding is a once-through boiler using high-solids water, some liquid must be present in the steam to ensure that the solids are kept in solution.

Many types of fuels are suitable for combustion in the MSFBC. High grade coal, low grade coal, or petroleum coke can be fired in the same unit. Typically, the fuel is gravity-fed from a volumetric feeding device.

Operating bed temperatures are between 1600 and 1740°F; calcium to sulfur ratios will vary from 3 to 4.5. These values, of course, will depend on the fuel combusted.

Unit turndown is projected to be 3:1. Pilot plant operations have shown the MSFBC design capable of quick response to varying steam demands through controlling recirculation rates.

Struthers-Wells has recently sold one of these units to Continental Oil Co. for a Texas Oil field application. Scheduled for start-up in mid 1981, this unit is sized at 50,000 lb/hr steam.

Babcock & Wilcox

Babcock & Wilcox (B&W) has been engaged in FBC research and development activities for several years. In their Alliance, Ohio Research facility, B&W has (3) FBC test units. The largest of these has a 6 ft x 6 ft bed area, and is currently being operated under a contract with the Electric Power Research Institute (EPRI).

This test unit is rated at 23.8 million Btu/hr input, and can fire coal and other solid fuels. Steam generating tubes are located both in the bed and in the freeboard space. The upper (freeboard) steam section is designed to produce additional 150 psig steam while cooling the combustion gas to 900°F.

Fuel and limestone are mixed in a ratio required for adequate sulfur capture. The mix is then fed pneumatically through (4) underbed feed points.

As stated, this unit is used exclusively for testing. This design configuration is not necessarily representative of a commercial FBC unit which B&W would offer.

Presently, B&W is performing final design on a 20 MW atmospheric FBC unit for TVA. This unit is scheduled for start-up in a couple of years.

Combustion Engineering

Combustion Engineering is presently in the final stages of construction on an FBC demonstration unit, sponsored by the U.S. Department of Energy. This unit is being built at the Great Lakes Naval station in Chicago.

This FBC boiler will generate 50,000 lb/hr of steam at 365 psig and 560°F. The steam will be used for space heating.

The unit employs a waterwall design with the boiler tubes in an "A" configuration - including one steam drum and two lower mud drums. Superheater tubes are located in the bed.

Coal and limestone are mixed in a ratio suitable for good sulfur capture. The mix, sized to minus 1/4 inch, is fed pneumatically underbed. The bed is fluidized by blowing air through six separate air ducts underneath the bed area. The cross-sections of the air ducts correspond to varying portions of the bed area - two ducts at 25% each and four ducts at 12-1/2% each. By shutting of air flow in certain ducts, the appropriate amount of bed area can be slumped and the unit will turndown. Theoretically, this arrangement permits a turndown of 12:1.

As with most FBC design, the CE FBC boiler requires recycle of elutriated fines to achieve adequate carbon burnup. The design value for steady-state recycle is 82% of coal feed. Of course, this value is subject to change based on actual operating experience.

Presently, CE is not offering FBC systems commercially. After successful demonstration of the Great Lakes unit (scheduled start-up in first quarter of 1981) CE plans to actively pursue industrial FBC markets. Their FBC product line will consist of units up to 500,000 lb/hr steam capacity, with steam conditions to 1550 psig and 950°F.

Deltak/Copeland

Deltak Corporation has joined forces with Copeland Associates, Inc. to produce and market the C-D line of fluidized bed boilers. These units are commercially available in a capacity range of 40,000 to 80,000 lb/hr steam per shop-fabricated module, with steam pressures up to 1000 psig.

Solid fuels can be fed to the unit either pneumatically or mechanically, but the system design requires that the fuel be introduced in the bed and overbed. The operating bed height is between four and eight feet, and the fludizing velocity is from four to eight feet per second. With the rather deep bed and average fluidizing velocities, the C-D unit design is offered as achieving good combustion efficiencies without recycle. This is due to increased carbon residence time in the bed, promoting adequate burnup on the first pass.

Dorr-Oliver

Dorr-Oliver presently offers (3) separate lines of FBC units, either directly or through license agreements with E. Keeler Co. and Thermotics, Inc.

For burning sludges, process wastes, or low grade fuels, Dorr-Oliver has a line of fluidized bed boilers in a capacity range of 10,000 to 60,000 lb/hr steam at conditions up to 600 psig and 750°F. These units are suitable for process heating and cogeneration applications.

E. Keeler Co. is offering FBC boilers in sizes ranging from 15,000 to 150,000 lb/hr steam, using Dorr-Oliver technology. These units can fire coal, coal waste, coke, COM, wood, or wood wastes. Generating steam up to 900 psig and 850°F, these units are suitable for process heat or cogeneration application.

Thermotics, Inc. is using Dorr-Oliver technology in offering a line of fluidized bed steam generators for enhanced oil recovery. In sizes from 20,000 to 100,000 lb/hr steam, these units generate high pressure steam (1000 to 2500 psig) to be used in oil recovery operations in existing oil fields.

At the present time, Dorr-Oliver is working in conjunction with Keeler on a DOE-sponsored project to build a 20,000 lb/hr 150 psig FBC boiler firing anthracite culm. Anthracite culm (mining waste) is a low-grade, low-volatile, high ash fuel which is difficult to burn in a conventional combustor. This boiler is currently in shop fabrication.

Energy Resources Co. (ERCO)

ERCO is currently offering FBC units commercially in sizes up to 400 million BTU/hr (up to 125 million BTU/hr per skid-mounted unit). Steam conditions available will be suitable for most industrial applications, including cogeneration.

Recently, ERCO completed construction on a 4 ft. x 9 ft. FBC demonstration unit in their test facility in Cambridge, Massachusetts. The unit is rated at 20,000 lb/hr steam, and can burn a wide variety of coals. This FBC test facility will be used to test various components and operating parameters relevant to commercial FBC systems.

FluiDyne Engineering Co.

FluiDyne has done both materials and parametric testing on FBC units in their lab facilities in Minneapolis, Minnesota. As a result of this background, FluiDyne is offering commercial FBC steam generators and air heaters.

The FluiDyne line of FBC units is available in capacities up to 75 million Btu/hr output in a single unit, with steam conditions up to 650 psig and 750°F. These units can burn a variety of fuels, from high-grade coals to solid waste materials. The FBC air heater uses steel alloy heat transfer tubes located in the fluid bed to heat air to temperatures up to 1000°F. Applications for heated air include paint drying, food and chemical processing and product drying.

Both the FBC air heater and the FBC steam generator can be applied to cogeneration through the use of a gas or a steam turbine.

Foster Wheeler Energy Corporation

Foster Wheeler has sold four FBC steam generators. Two of these units are operational, and two are currently in engineering.

At Rivesville, West Virginia, Foster Wheeler designed, supplied and erected a 300,000 lb/hr FBC (1300 psig/925°F) steam generator for the Monongahela Power Company. This unit was funded by the U.S. Department of Energy. In conjunction with a turbogenerator, the facility produces 30 MW of electrical power. Operations began in 1976.

Foster Wheeler also supplied a 100,000 lb/hr FBC steam generator at Georgetown University in Washington D.C. This produces steam at pressures form 275 psig at 625 psig for heating and cooling loads at the University. Georgetown University and U.S. Department of Energy co-funded this facility, which started operations in July, 1979.

Foster Wheeler is involved in another DOE-sponsored FBC project. This entails engineering and erecting a 100,000 lb/hr FBC steam generator to provide a portion of the heating load for the city of Wilkes-Barre, Pennsylvania. The boiler will burn anthracite culm, and is expected to start-up in 1982.

Currently, Foster Wheeler is designing a coal-fired FBC unit for a European oil company. This facility will generate 110,000 lb/hr of steam at 1300 psig/955F to be used for cogeneration.

Foster Wheeler is offering commercial fluidized bed steam generators in capacities from 50,000 to 600,000 lb/hr of steam, in high pressure and superheated conditions if required.

Johnston Boiler Co.

Johnston Boiler Co. is currently marketing their Fluid-Fire line of FBC packaged boilers on a license from Combustion System Ltd. of UL. This unit incorporates a modified firebox, firetube design with a water-cooled combustion chamber. The fluid-fire boilers are available in capacities from 2500 lb/hr to 50,000 lb/hr of steam, with pressures up to 300 psi.

These units are multi-fuel boilers, capable of firing liquid, gaseous, or solid fuels sized up to 1-1/4 inches. Each boiler consists of three distinct combustion chambers which can be operated independently. Solid fuels are fed through a screw feeder located above each bed; a separate screw feeder in each chamber is provided for metering sorbent.

Each of the fluid-fire units is a package-type skid-mounted boiler.

Johnston has sold approximately twenty of these units, the largest one being a 40,000 lb/hr of steam 200 psig boiler. This boiler has been recently started-up, and although no data is yet available, it appears to be functioning well.

Pyropower

General Atomic Co. has recently come to an agreement with Hans Ahlstrom Co. of Finland to manufacture and market a line of FBC systems in the U.S. These units are to be offered under the name of Pyropower.

Ahstrom has (2) commercial FBC systems operating in Finland. One unit is a multi-fuel 45,000 lb/hr steam generator retrofitted to an existing boiler at a paper mill. The second FBC unit is a 20,000 lb/hr boiler to be used for district heating.

The basic design feature of the Pyropower unit is a fast circulating fluid bed. The system consists of a tall combustion chamber, a cyclone, and a recycle line. The bed is fluidized at a rather high velocity as primary air is introduced through a lower grid. Secondary air is introduced at various levels in the combustion chamber to ensure adequate space velocities to maintain material circulation through the cyclone and back to the combustor.

The Pyropower line of FBC units is to be available in size ranges up to 200,000 lb/hr of steam, with steam conditions suitable for cogeneration or enhanced oil recovery.

Wormser Engineering

Wormser has developed an FBC system that is particularly well suited to retrofit installations. Called the Wormser Grate, this unit employs two-staged combustion and two distinct fluid beds. One of the beds is used for combustion, and the other, located directly above the first, is for desulfurization. The coal is fed pneumatically underbed through an in-line and rotary airlock. The unit may be coupled with an existing boiler by directing hot combustion gases from the fluidized bed through the boiler. This arrangement, utilizing heat transfer surfaces in both the FBC unit and the existing boiler, can result in uprating of steam capacity.

The Wormser Grate is available in capacities from 6,000 lb/hr to 30,000 lb/hr of process steam. Steam conditions and pressure ratings match those of most firetube boilers. All units are shop-assembled package design boilers.

Wormser has an operating unit rated at 3 million Btu/hr input. Connected to an industrial boiler, this unit, as of July 1980, had logged some 1600 hours of operation.

York-Shipley, Inc.

York-Shipley currently manufactures a line of FBC incinerators which fire wood residues, generating hot combustion gases suitable for both steam production and process heating. With 18 operating units, York-Shipley will furnish a wood-burning FBC unit coupled with a firetube boiler capable of generating up to 92,000 lb/hr of steam. For higher pressure steam the York-Shipley unit can be attached to a watertube boiler. Units generating heated air for process applications are available in sizes up to 120 million Btu/hr.

Presently, York-Shipley is pursuing both testing and marketing efforts to utilize other low-grade fuels, such as heat, bio sludge, rice bulls, and pelletized RDF. The company expects to offer a line of integral FBC boilers (as opposed to an FBC incinerator with a waste heat unit) in the future. These units will burn biomass, coal, or oil to generate up to 140,000 lb/hr of steam with conditions up to 350 psi and 650°F.

FBC OPERATIONS

There are several industrial FBC units currently in operation in the United States. However, most operating facilities were installed for private industry, so design and operating data is for the most part proprietary. Generally, operations have been successful.

The largest industrial FBC unit currently operating in the U. S. is that at Georgetown University in Washington, D.C. As stated earlier, this unit was sponsored by DOE; therefore, the operating data is publicly available and will constitute the bulk of the following discussion.

The Georgetown FBC steam generator was started-up in July, 1979. Data on long term operations was taken starting in the first quarter of 1980. The table below compares design data with actual performance for several key values.

	Design	Average 1st Quarter 1980	Average 2nd Quarter 1980
Steam Production (lb/hr)	100,000	33,700	31,500
Calcium/Sulfur Mole Ratio	3:1	4.5:1	6.3:1
SO_2 Emissions (Lb/10^6 Btu)	0.78*	0.5	0.3
Excess Air (%)	20	32.2	35.7
Thermal Efficiency	83.51	66.5	68.6
Hours of Operation	-	797	1,127

*Washington, D.C. Limit

As noted, the values shown in the last two columns represent average data over the first six months of 1980. Data taken on any given day could approach the design values listed. For example, the boiler has generated 100,000 lb/hr of steam, but not for long durations. This is because steam production is dependent upon the University's steam demand, which is expected to rise in the third and fourth quarters of this year.

The calcium-to-sulfur ratios are higher than the design value for a couple of reasons. The freeboard SO_2 analyzer, which is used to trim the limestone feedrate, has not been functioning properly. This means that manual control of limestone flow is necessary, and the precision of automatic control is lost. In addition, the unit has been shut down and started up many times during the six-month period. This entailed charging the bed with fresh limestone (as make-up) more often than necessary for sulfur capture. The result, therefore, is a high overall limestone to coal, or calcium to sulfur ratio.

Thermal efficiency has averaged lower than the expected value for two reasons. First, the fines reinjection system from the mechanical collector has not been functioning as required for good carbon burn-up efficiency. This has resulted in higher than expected unburned combustible losses. Second, the unit has been operated with higher than design excess air. Low load operation has a tendency to increase excess air to maintain adequate fluidization. This is aggravated with two bed operation on light loads.

Even though the Georgetown unit has logged more than 2000 hours of total operation, the longest period of continuous operation to date is about 350 hours. This lack of continuous operating experience means that one must be cautious about drawing conclusions on these results. Operations in the near future should show improvement as steam demand and continuous steam production increases.

FBC MARKET AND ECONOMICS

The future of FBC in the industrial marketplace depends on several technical and economic issues. These issues, which must be demonstrated to the satisfaction of potential users, include

- long term operating reliability
- adequate combustion efficiency
- successful startup and turndown methods
- fuel and limestone feeding
- environmental compliance
- competitive economics

Many of these areas have been addressed in bench scale and pilot plant testing as well as in engineering studies and design work. Most published data has shown FBC systems to be very competitive with conventional heat generating systems, both technically and economically[5,6].

Obviously, such technical issues as reliability, turndown, and, environmental compliance must wait demonstration in actual large-scale plant operations. Most industrial users would require that an FBC unit of at least 20 million Btu/hr input operate continuously in an industrial or commercial setting for about one year. This would confirm the capability of FBC systems to meet the demands of the industrial sector.

Several installations, including those engineered and built by Foster Wheeler, Combustion Engineering, Johnston, and Babcock/Riley, should have a full year of operation during 1981. It is expected that after successful demonstrations next year, the market for FBC should open up.

Capital and operating cost estimates generated to date have concluded that FBC systems are economically competititve with conventional combustion systems in cases where flue gas desulfurization is necessary. For example, a 100,000 lb/hr FBC steam generating plant located in Chicago and burning a high sulfur Illinois coal would require a capital investment of $8,580,000. Annual operating costs would be about $1,954,000. The resultant steam cost is $7.35 per 1000 lbs. of steam. These values are all 1979 dollars.

These costs compare with a capital cost of $8,610,000, an operating cost of $1,940,000, and a steam cost of $7.37 per 1000 lbs. of steam for a spreader stoker boiler plus a flue gas desulfurization system.

Both of these estimates are accurate to ±25%. The costs should not be considered definitive, since design and operating characteristics will be both site and user specific. However, this comparison was done on a consistent basis and, at the very least, reflects the competitive economics of fluidized bed combustion technology.

In conclusion, both suppliers and users have expressed continuing interest in FBC as an alternate technology for heat generation. Within the next couple of years, after technical and economic issues are resolved, fluidized bed combustion should have a distinct role in the industrial marketplace.

REFERENCES

1. Grant, A. J., "A Discussion of Fluidized Bed Combustion and Two Stage Coal Gasification," 1978 Industrial Fuel Conference, Purdue University, October, 1978.

2. Locke, H. B. et al, "Fluidized Combustion in Great Britain - Environmentally Clean Steam and Power Generation from Coal, Heavy Oil and Dirty Fuels," Proceedings of the Fourth International Conference on Fluidized Bed Combustion, Washington, D.C., December 9-11, 1975.

3. McKenzie, E. C., "Burning Coal in Fluidized Beds," Chemical Engineering, August 14, 1978.

4. Reed, R. R., "Fluidized-Bed Combustion for Small Users of Coal." Coal Technology '78 Conference, Houston, Texas, October 17-19, 1978.

5. Davy McKee Corporation, "Cost Comparison Study; Industrial Size Boilers; 10,000 to 400,000 Pound per Hour", October, 1979.

6. Farmer, M. H. et al "Application of Fluidized-Bed Technology to Industrial Boilers, Exxon Research and Engineering Co., January, 1977.

FBC SUPPLIER	STEAM OUTPUT	BOILER EFFICIENCY %	FUELS	TURNDOWN RATIO	FEED SYSTEM	BED TEMPERATURE °F	SUPERFICIAL VELOCITY FT/SEC	BED DEPTH FT.
Babcock Contractors/ Riley	50,000 to 500,000 lb/hr 1000°F 1600 psig	80-85	Fuel Oil, Gas, Coal, Wood, or Refuse	Variable, as required	Over-bed for units less than 100,000 pph under-bed for larger units	1500-1600 Nominal	8 Nominal	3 Typical
Battelle/ Struthers Wells Corp.	50,000 to 250,000 lb/hr 80% quality 2650 psig	84 Typical	Coal and Petroleum Coke	3:1	Gravity Feed with Pneumatic Assist	1600-1740	20-30 in combustor; 1.5-2 in external ht. exch.	---
Combustion Engineering	Up to 500,000 lb/hr 950°F 1550 psig	80 Typical	Coal	8:1 Nominal 12:1 Max	Pneumatic Under-bed	1550 Nominal	7 Nominal	3 Typical
Deltak/ Copeland Associates	40,000 to 80,000 lb/hr per module. Up to 1000 psig	80 Typical	Oil, Gas, Coal, Solid Refuse	Variable, as required	In-bed Pneumatic or Mechanical	1500-1600	4-8	4-8
Dorr-Oliver	10,000 to 60,000 lb/hr 750°F 600 psig	70	Sludges, Process Wastes, High Water Fuels	1.4:1	Mechanical Over-bed	As Required for Fuel	As Required for Fuel	As Required for Fuel
E. Keeler (with Dorr-Oliver)	15,000 to 150,000 lb/hr 850°F 900 psig	82	Coal, Coal Waste, Coke, Com. Wood, Wood Wastes	5:1	Mechanical	As Required for Fuel	As Required for Fuel	As Required for Fuel
Thermotics Inc. (with Dorr-Oliver)	20,000 to 150,000 lb/hr 80% quality 1000-2500 psig	82	Coal, Coke, Com.	4:1	Mechanical	As Required for Fuel	As Required for Fuel	As Required for Fuel
Energy Resources Co. (ERCO)	Up to 125,000 lb/hr per module	---	Coal, Waste Oil, Waste Solids	---	(1) Pneumatic In-bed	(1) 1500-1600 Typical	(1) 3-9	(1) 3-6

(1) Based on design values for ERCO's 4 ft. x 9 ft. demonstration unit.

TABLE I

FBC SUPPLIER	STEAM OUTPUT	BOILER EFFICIENCY %	FUELS	TURNDOWN RATIO	FEED SYSTEM	BED TEMPERATURE °F	SUPERFICIAL VELOCITY FT/SEC	BED DEPTH FT.
FluiDyne Engineering Co.	Up to 75 x 10^6 BTU/hr 750°F 650 psig	84	Coal, Wood Solid Waste	2 1/2:1	Pneumatic	1600 Nominal	6	Variable with Fuel and System Characteristics
Foster Wheeler	50,000 to 600,000 lb/hr High Pressure & Superheat	To 85	Coal and Petroleum Coke	4:1 Normal	Mechanical Spreader (Typical)	1550-1600	8	3 Typical
Johnston Boiler Co.	2,500 to 50,000 lb/hr Saturated 300 psi	80	Oil, Gas, Coal, Wood, Solid Refuse	3:1 Standard 6:1 Optional	Screw Feeders	1500-1600 Nominal	9 Ave. 6 At Top of Bed 12 At Bottom	2 1/2 - 3
Pyropower (General Atomic Co. & Ahlstrom)	20,000 to 200,000 lb/hr 970°F 2600 psia	---	Coal, Peat, Wood, Petroleum Coke, Solid Wastes	3:1 Typical	Screw Feeders	1550	---	---
Wormser Engineering	6,000 to 30,000 lb/hr Temp. & Press. Match Firetube Boiler Design	(2) 84	Oil, Gas, Coal	(2) 3:1 Modulation 10:1 On-Off 30:1 Overall	(2) Pneumatic Under-bed	(2) 1550	(2) 7.2	(2) 1
York Shipley	Up to 140,000 BTU/hr 650°F 350 psi	75-90	Biomass, Oil, Coal	5:1	Pneumatic	1400-1600	11-12	2

(2) Based on 3 MM BTU/hr boiler Wormser is currently operating.

TABLE I (CONTINUED)

PRELIMINARY RESULTS FROM THE DOE COAL-OIL MIXTURE DEMONSTRATION AT SALEM HARBOR

Richard M. Dunn

PRELIMINARY RESULTS FROM THE DOE COAL-OIL MIXTURE DEMONSTRATION AT SALEM HARBOR

Richard M. Dunn
New England Power Service Company
Westborough, Massachusetts 01581

1.0 Abstract

An 80 MW coal fired boiler which had been converted to burn #6 oil in 1969 was equipped with new burners and accessories to burn coal-oil mixture. A description of the 30% coal and 70% oil mixture preparation facility is given. Results of initial coal-oil mixture (COM) blending, combustion and stack emission tests are presented along with a discussion of initial startup problems encountered. EPRI will support an expanded boiler performance test program designed to yield information which may be utilized in the conversion of oil fired designed to COM. Recommendations are made for additional COM research and product development to improve reliability.

2.0 Introduction

The use of COM as a substitute boiler fuel for oil has been studied since the late 1800's. The recent history of COM progress in the United States, Canada and Japan has been discussed adequately elsewhere.[1,2,3] The bibliography in the Proceedings of the First International Symposium on COM held in May 1978 lists many significant publications involving basic research and industrial experience.

The COM program in the United States received additional momentum as a result of the 1973 OPEC embargo. In February 1976, the U.S. Energy Research and Development Administration (ERDA) issued a program opportunity notice (PON FE-3) which identified seven distinct categories for industrial COM combustion demonstrations. New England Power Service Company (NEPSCo), the engineering and construction subsidiary of the New England Electric System, was selected by ERDA to construct a COM demonstration facility on a utility boiler which was originally coal fired but later converted to residual oil.

Salem Harbor Unit #1 is rated at 80 MW and was installed in 1951. The Babcock & Wilcox boiler is front wall fired with a rated steam flow of 625,000 pounds per hour, a design pressure of 1675 psi and superheat and reheat temperatures of 1000 F. Although the unit was originally designed to fire both coal and residual oil, coal remained the primary fuel until 1968. The primary superheater tube side spacing is 3½ inches, the secondary superheater spacing is 12 inches and the reheater spacing is 6 inches. The 2½-inch economizer tubes are on 4-inch centers without fins. The unit also has two banks of tubular air preheaters arranged in series with vertical arrays of 2-inch tubes through which the flue gases pass.

The boiler is equipped with retractable lance type soot blowers in the superheater, reheater and convection pass and intermittently revolving air puff wall blowers in the furnace. In addition, the ash removal system was designed for coal firing with a sloping furnace bottom, a bottom hopper ash conveying system and air preheater and economizer ash hoppers.

The flue gas passes through an electrostatic precipitator which was installed in 1951 prior to the introduction of modern solid state voltage and rapping controls and is equipped with expanded metal collecting plates. The original collection efficiency was guaranteed at 97% in accordance with the ASME code in common use at the time.

The twelve original B&W coal burners which had been equipped with high pressure (1000 psi) mechanical atomizers for residual oil firing were replaced with low pressure (130 psi) Forney Verloop burners. The two-stage air atomized Forney burner is rated at 80 million Btu's per hour. Low pressure air (38 inches H_2O) is utilized in two stages to atomize COM. Approximately 5% of the total combustion air is contributed as primary air for fuel atomization.

The Forney burners was selected primarily because of the relatively large size of the atomizer orifice and swirl block passages which was expected to resist plugging with COM. It was also expected that the low fuel pressure would tend to minimize erosion of the stainless steel atomizer. NEPSCo had also successfully used Forney Verloop burners on two 450 MW units firing #6 oil with magnesium oxide fuel additives and atomizer erosion had not been severe. Further assurance that the Forney burner would be suitable for long term COM firing came from results of the General Motors pilot program[4] along with reports of high pressure mechanical atomizer erosion in Canadian tests.

3.0 Objectives

In responding to the original PON in 1976, it was noted that various investigators observed a rapid increase in COM viscosity above 30% coal concentration.[5,6,7] In addition, although the non-Newtonian nature of high concentration COM had been noted, insufficient data on the fluid properties had been published to permit reliable predictions of pressure drop and other parameters for design purposes. In the case of the demonstration planned on the base loaded Salem Harbor Unit #1, dual fuel feed capabilities were to be included so that if difficulties with COM were experienced, a rapid switch to #6 oil would be possible. NEPSCo subsequently submitted a proposal to ERDA based on a maximum coal concentration of 30%.

When the project was initiated, the Massachusetts air emission regulations required the use of oil with a sulfur content less than 1%.

DOE advised that COM commercialization, at least for the Northeast, should encourage the use of medium or high sulfur Eastern coal. A 2.6% sulfur Pittsburgh Seam #8 coal was specified with an assumed range of grindability of 55 to 70. After discussions with State and Federal environmental agencies, it was decided to request a one year variance to permit the burning of 2.2% sulfur COM blended from 1.0% sulfur oil and the 2.6% sulfur coal. The variance was approved in 1978 with additional provisions that permitted opacity excursions for short periods during COM firing tests.

Early in 1979, however, the use of 2.2% sulfur fuel was approved for the Metropolitan Boston Air Pollution Central District and annual renewal of the sulfur variance for the COM demonstration was no longer required.

In addition to the sharp increase in overall residual oil prices since the COM project was authorized, there has also been a widening price gap between 1.0% sulfur and 2.2% sulfur oil. Although the gap has narrowed somewhat in recent months, it has averaged as much as $6 per barrel during the last year. In view of these changing economic factors, production of COM from low sulfur oil will be discontinued at Salem. The 2.2% sulfur oil currently being burned on other units will be blended with an Eastern coal with a sulfur content less than 1.5%.

Additional objectives were to continuously blend a stable COM without stabilizing additives, if possible, by utilizing a medium shear bladed agitator mounted on the top of a cylindrical blending tank. COM would be stored in an existing oil storage tank equipped with a roof mounted agitator, thermal insulation and a heating coil for periods of up to one week. The unit would be normally base loaded on COM but during startup would be fired with #6 oil. (See Figure 1, simplified COM schematic)

4.0 Test Facility

Since on-site COM production was part of the original DOE objective, it was decided to locate the COM preparation facility in a separate metal building constructed as a temporary addition to the existing boiler enclosure. One of the first decisions to be made was related to coal fineness and grinding equipment. The fine grinding technique utilized by Florida Power Corp.[8] was reviewed and rejected on the basis of high grinding energy and reported difficulties with grinding equipment. A program had been initiated by NEPSCo at the University of Massachusetts in 1976 by Prof. R.L. Rowell[9] to investigate the stability of 200 mesh coal in oil. Many of the initial observations during this study indicated that COM stability could be obtained with 200 mesh coal when utilizing a variety of surfactants. With indications that COM stability could be demonstrated with typical utility grinds, one of the four existing Babcock & Wilcox Type E pulverizers installed when the unit was originally constructed was selected for integration into the COM production facility.

The three options available for coal drying were a separate boiler to produce hot air or flue gas, primary air from an existing boiler or flue gas from an existing boiler. Since the original pulverizer primary air ducts and dampers were still intact, primary air from the existing boiler was selected.

A 100 ton pulverized coal storage bin was constructed above the new COM blending facility. The bottom of the bin as originally installed was an inverted pyramid with a slope of 60^{o}. The bin was thermally insulated and heated with electric heat tracing cable to insure that coal remained dry. Since 13 tons of coal per hour would be transported from the pulverizer to the top of the storage bin after mixing with primary air, a booster fan developing an additional 8 inches H_2O was required.

The installation of stored pulverized coal systems had essentially ceased in the United States by the late 1950's in favor of direct fired systems. Early experience with storage systems and with pulverized coal in general resulted in the National Fire Protection Association (NFPA) issuing guidelines to minimize the hazards related to coal dust accumulation. It was concluded that the COM preparation facility must comply with the classification for dust hazard areas as defined in Volume 6 of the NFPA code. Section 85F specifies that stored pulverized coal should be kept under an inert atmosphere. Carbon dioxide was selected based on its availability, relatively low cost and a density exceeding that of air which promotes effective blanketing. Laboratory experiments later established that carbon dioxide reduces stability and that the COM exhibited properties resembling a mixture without surfactant. Further tests indicated that nitrogen does not have this effect and nitrogen was substituted for the inerting gas in the pulverized coal storage bin and in the COM blending tank.

The NFPA standard recommended other measures which complicated the design of the COM preparation facility. Totally enclosed explosion proof motors and other electrical accessories were required within the building.

A means of separating the pulverized coal from the hot transport air was required prior to discharging the coal into the storage bin. A cyclone separator was selected but disposal of the transport air which would be contaminated with coal fines was still required. The options here appeared to be a bag house separator or to inject the air and coal fines into the boiler. The latter option was selected and two Inconel 601 injection nozzles were installed in the front waterwall above the top burner elevation. This arrangement must be integrated into the boiler fuel safety interlock system, however, since it becomes a source of combustibles entering the furnace. The cyclone separator is isolated from the pulverized coal storage bin with a rotary airlock so the bin can be maintained at a positive nitrogen pressure of 4 inches H_2O.

The 100 ton pulverized coal storage bin was designed to function as a day tank. Coal grinding is permitted only when boiler output exceeds a 70% rating to insure that the coal fines entering the boiler from the cyclone separator are consumed. COM blending can be scheduled independently of boiler or pulverizer operation with this arrangement.

Continuous COM blending was preferred over batch operation. The evaluation of COM mixing equipment conducted by Adelphi University for NEPSCo has been reported previously.[10] The COM stability was the primary standard for mixing equipment acceptance. Mixtures prepared with various equipment were examined in a sedimentation column described previously[9] and exposed to high shear rates in a viscometer. A conventional bladed turbine agitator was ultimately selected over various peripheral and injection type mixers. As a further precaution, laboratory tests were conducted by Chemineer, Inc., the manufacturer of the turbine agitator. The type HTD agitator, a medium shear devide with two sets of pitched blade turbine impellers was installed on top of the 8000 gallon COM blending tank. Prior to initial startup, a rotary airlock was added between the COM blending tank and the gravimetric coal feeder to isolate the feeder from pressure fluctuations in the tank (See Figure 2).

An Acrison gravimetric feeder Model 203B was provided by Combustion Engineering, supplier of the storage bin, cyclone separator, rotary air locks and related coal piping.

The continuous COM blending controls are very straightforward. The #6 oil flow to the COM blend tank is controlled by the blend tank level controller. Since, at the present time, equipment for measuring or controlling COM concentration has not been found for continuous in-line blending, COM concentration is controlled by a pulverized coal flow ratio controller. Coal flow and oil flow ratio is preset to produce the desired COM concentration. COM concentration is verified in the laboratory using a simple solvent extraction technique.

The COM stability studies conducted at the University of Massachusetts[9] indicated that a number of commercially available surfactants were effective stabilizers. Ultimately a BASF Wyandotte Corp. product, ES-7071, was selected for use during the first six months of the one year demonstration based on laboratory evaluations with a sedimentation column and on the approximately 17¢ per million Btu cost for the .25% surfactant continuous feed. Dynamic as well as static stability is being observed during the demonstration. COM samples can be withdrawn from the system at a number of locations in order to detect coal particle setting, attrition or agglomeration.

In order to provide for periodic COM blending system forced outages or inspections, an existing 700,000 gallon oil tank was modified to store COM. A 75 HP Chemineer bladed turbine agitator with 120 inch blades was supported above the tank roof. Four vertical mixing baffles were installed on the tank walls. Thermal insulation was applied to the tank and an internal steam heating header was installed. The COM storage tank is located several hundred feet from the boiler and the COM preparation facility requiring several oil and COM pipelines of up to 500 feet. Three in-line Kenics Turbumixers which are vaned static devices were installed in an attempt to resuspend coal particles that may tend to settle out in the piping. All fuel piping was insulated and heat traced with electric heat tracing. Electric tracing was recommended over steam tracing because it tends to permit more accurate temperature control and condensate return lines are avoided. (See Figures 3 and 4)

The COM is transferred from the blending tank to the storage tank from the storage tank to the burner supply pumps utilizing two Tuthill Model 600 lobe type pumps. Each pump was equipped with #6 oil continuous flushing connections on the shaft seals to minimize damage from coal particles.

The four 25 GPM burner supply pumps are Viking abrasive liquid rated which were supplied by Forney Engineering with the burners. Pumps equipped for abrasive service are equipped with ceramic shaft seals, a carbide idler pin and case hardened gears. Each pump is driven by a variable speed DC motor SCR drive. The variable speed pump operates as a fuel orifice delivering COM to the burners at approximately 50 psi. An advantage of this arrangement is that fuel flow control valves can be avoided eliminating valve wear with modulating COM flows.

One of the two major obstacles was the selection of instrumentation for COM service. A preliminary DOE survey of instrumentation for COM viscosity and coal concentration measurements[11] recommended further field testing. Two MAPCO sonic flow meters were selected for COM fuel supply and blending plant output flow measurement.

The other obstacle encountered in component selection was related to the potential for erosion or wear of parts exposed to COM. Serious deterioration attributed to erosion had been observed by General Motors[4] and in a pilot program in New Brunswick, Canada.[12] When NEPSCo initiated engineering late in 1977, few manufacturers providing equipment for handling liquid fuels expressed much interest in the COM market. This was especially true in the positive displacement pump area where the market potential for COM was apparently considered small and produce development was not being emphasized. Manufacturers were in most cases unwilling to provide equipment with improved erosion resistant alloys. EPRI had funded two studies on pump alloys for the coal liquefaction program[13,14] which did identify specific alloys that tended to resist erosion from 200 mesh coal in an oil which was similar to COM. Some manufacturers, however, did propose equipment rated for erosion service but few were able to cite experience with slurries similar to COM. Spare parts have been purchased for most process equipment exposed to COM and the ability of certain specialty contractors to apply hardened surface coatings has been investigated. In order to docoment the erosion encountered, a number of component parts have been carefully weighed and their dimensions recorded. In addition, 13 pipe fittings, valves and other process elements have been identified as erosion specimens for later removal and examination

5.0 Demonstration Program

The dmeonstration is divided into a feasibility phase and the long term demonstration phase. During the feasibility phase which commenced on August 1979 and is currently being completed, the COM preparation and storage facility has been placed in operation and stack emission compliance tests have been conducted at 10, 20 and 30% coal concentration. The results of the emission tests are shown in Figure 5.

5.1 Supplementary Test Program

The Alliance Contract Research Division of the Babcock & Wilcox Company will conduct a series of boiler performance tests with additional support from EPRI. The objective of this expanded test program is to provide information which can be utilized in the conversion of oil fired designed boilers to COM. A fuel, ash and slag characterization program is included. Heat absorption in the furnace, superheater, reheater and economizer will be measured when burning COM and #6 oil. Flame photographs will be made and high temperature furnace probes will be inserted to evaluate slag properties. Although a boiler derating when burning COM in the coal designed boiler at Salem is not expected, some derating of oil fired boilers may be required unless the ash fraction can be removed prior to COM blending.

5.2 Test Plan

The test program which has been described in a recent paper[15] consists of a series of four structured tests on COM and one on #6 oil. After each test, boiler efficiency by the heat loss method will be computed. Additional measurements will be taken at the COM preparation facility. Raw coal, oil and COM samples will be taken at frequent intervals throughout the demonstration. Flue gas analysis will be supplemented by measurements of fly ash resistivity and particle size distribution.

In addition to the erosion specimens discussed previously, 20 boiler waterwall, superheater and reheater tube samples were installed prior to burning COM. These samples will be removed at the end of the demonstration to observe fireside deposits or erosion. Significant erosion is not expected, however, since it was not a problem when the unit burned 100% coal.

6.0 Results

When the 12 80 million Btu Forney Verloop burners were installed in July 1978, the burner throat diameter was decreased to 26" with the addition of refractory. The original 31" throat for coal firing had been reduced to 29" when the unit was converted to residual oil in 1979. Initial operation of the Forney burners on oil was unsatisfactory. The fires were long and smoky with many sparklers or fireflies observed. Fuel coking was experienced on burner tips and diffusers which required cleaning several times during the first six months of operation. The burner throat diameter was reduced to 22" in an effort to increase the pressure drop from the windbox to the furnace from less than 2 inches H_2O to 4 to 6 inches. This change was only partially successful although the fires appeared brighter and shorter. Forney then concluded that secondary air was being distributed unequally to the burners and that the burner secondary air registers were open too wide to permit optimum mixing or acceptable burner throat pressure drop. As a result of this analysis, the burner throats were increased to a 24" diameter with the installation of 309 stainless steel tapered throat inserts. The secondary air registers were partially closed resulting in no more than a 3 inch H_2O windbox to furnace pressure drop. In addition, a number of turning vanes were installed in the upper and lower windboxes to improve the distribution of secondary air to each burner. As a result of these changes, the flames appeared somewhat shorter and brighter and sparklers were reduced.

Forney advised during initial engineering that they had developed a bladed or vaned swirler for COM combustion which was proposed for installation on one burner on an experimental basis. After two modifications and a number of field observations of the burner equipped with the swirler, Forney recommended the installation of the 11 remaining swirlers in early 1980. After installation and further field observations, it was noted that coking of burner parts had been eliminated, flames became shorter and brighter and sparklers were further reduced.

The unit has been switched from residual oil to COM one three burner level at a time without difficulty. Burner adjustments are not required when fuel is switched. The flames are just about indistinguishable from oil fired at 10 and 20% COM. When firing 30% COM, there is clearly a darkening at the burner throat opening but a somewhat brighter radiant flame has been observed.

Stack opacity as measured by the Environmental Data Corporation (EDC) flue gas monitor located at the base of the stack was maintained below the 20% regulatory limit when firing COM with a 20% coal concentration by reducing unit output from 84 to 78 MW. Particulate emission levels were measured by KVB, Inc. at the stack midheight using EPA Method 5 procedures with a twelve point traverse. The results of particulate emission tests on oil and COM summarized in Figure 5 indicate that compliance with the .12 pounds per million Btu's regulatory limit was achieved with 10 and 20% COM. When 30% COM was initially fired in Feburary 1980, the 20% stack opacity was exceeded at full load. Load was reduced to 68 MW to reduce opacity to the 20% limit but a particulate test was not performed due to severe winter weather.

During the remainder of February, boiler output particulate measurements were taken when firing residual oil only in an attempt to determine why emissions varied widely with various oil cargoes. Data published by Exxon[16] and discussions with Florida Power & Light Company indicated that particulates may increase with increased fuel asphaltene content. Two fuel cargoes with 7.75% and 10.9% asphaltene content showed sharp increases in stack opacity when burned. Future fuel cargoes will be analyzed and efforts will be made to blend COM using oil with asphaltene content less than 4%.

An additional effort was made to resolve the problem of high stack opacity experienced on residual oil and COM. Professor Janos Beer of MIT was given a Forney burner assembly to conduct atomization tests in the laboratory. Initial observations of flame in the Salem boiler followed by laboratory tests with water suggested that fuel atomization may contribute to the opacity problem and also explain high levels of unburned carbon in the ash. Forney has suggested that both problems may be the result of high asphaltene oil. MIT advised that they have observed fuel droplets in the 1000 micron range and 100 to 200 microns should be the maximum size for proper atomizer performance. The Forney atomizer tip opening was reduced from 5/16" to 1/4" diameter in an attempt to increase pressure and improve atomization. Fuel pressure increased approximately 4 psi but a significant reduction in opacity was not observed.

The burner atomizer swirl block or tip is fabricated from 304 stainless steel and there was initial concern that erosion would be a serious problem. To date significant material deterioration has not been observed. Total COM firing, however, has been limited to no more than 15 days because of the problems related to stack opacity and particulate emissions.

Another serious problem encountered involves the inability of the pulverized coal storage bin to feed coal through the gravimetric feeder to the COM blending tank at the design rate of 30,000 pounds per hour. The maximum initial coal feed rate was about 10,000 pounds per hour and bin outlet flow fluctuated widely as the bin outlet alternately flooded and plugged. Combustion Engineering recommended the installation of a bin flow modifier similar to the "easy flow" device introduced by Bituminous Coal Research in the 1950's. It consisted of two slender pyramids with a common based oriented vertically above the bin outlet. The bin insert did not improve flow significantly and Combustion Engineering engaged Jenike & Johanson, consultants in the storage and flow of solids, to review the coal flow problem. As a result of field observations, it was recommended that the bottom slopes of the bin be lined with stainless steel and the bin bottom be aerated with nitrogen. The initial test after these additional modifications was only moderately successful. The nitrogen aeration nozzle flows were varied and gravimetric feeder flow indication and feeder speed recorded. Since the bin outlet flooded less often and the outlet flow was steader, it was hoped that further nitrogen feed and coal feeder adjustments would resolve the flow problem. These efforts were not successful and Combustion Engineering asked Jenike and Johanson to redesign the bin outlet to achieve the design flow rate. As a result, COM blending was suspended in late April 1980 and the bin outlet was removed. A slotted outlet opening 2' wide and 12' long was added with a 2' diameter variable pitch auger located in a housing below the slot. A variable speed auger drive was integrated into the existing gravimetric coal feeder control circuit. This system was placed in service in early June and the cost flowed smoothly from the bin although limited to approximately 22,000 pounds per hour. Combustion Engineering now expects to replace the auger drive motor and sprokets to achieve the desired 30,000 pounds per hour.

When the COM storage tank equipped with the agitator was initially filled with oil, pulsations in the tank walls were noted. Deflections of up to 1/16" were recorded and the agitator drive gear was changed temporarily to reduce the speed from 25 to 20 RPM. Tank wall stresses were ultimately determined to be well below acceptable limits. When the storage tank was filled with COM, the wall pulsations diminished due probably to the higher viscosity of COM and possibly higher average tank level.

The MAPCO sonic flow meter in the 4-inch COM line from the blend tank to the storage tank has not operated successfully at COM concentrations of up to 40%. A similar MAPCO flow meter in the 4-inch line from the storage tank to the burner fuel supply pump suction has been operating with reasonable consistency at up to 20% coal concentration. Tests were scheduled with MAPCO in late July 1980 to field test redesigned equipment with 30% COM.

During initial COM blending, a rubber expansion joint isolating the gravimetric coal feeder from the coal feed piping parted. The slide gate above the feeder was not closed for about five minutes and during this period possibly a ton or more of pulverized coal filled the COM preparation building with coal and dust. COM blending was interrupted and cleanup of the facility required a two-day effort with a vacuum track and water hoses. Since the pulverized coal dust hazard is real, every effort should be made to maintain a tight system and to prevent spills, especially if COM blending is conducted indoors.

Since the results of the limited COM firing tests to date indicate that compliance with Massachusetts opacity and particulate regulations is not possible when firing 30% COM at full load, efforts were initiated in May 1980 to improve the performance of the existing precipitator. The precipitator was installed in 1951 with a specific collecting area (SCA) of 142 with the design flue gas flow of 260,000 ACFM. Recent internal inspections indicate that clearances between discharge wires and collecting plates have been reduced due to plate warping and distortion. Recent calculations by the manufacturer, Research Cottrell, indicate that an SCA of 238 would be required to meet the .12 pounds per million Btu's particulate limit. Studies are still under way but it appears that the expanded metal plates in the existing precipitator should be replaced along with the installation of a new inlet collecting field 12' long.

While the matter of precipitator improvements is still being investigated, it is planned to reduce COM concentration from 30 to 20% coal in August 1980. A continuous COM burn will commence along with improvements to the precipitator plate rapping system to minimize particulate reentrainment. KVB will conduct tests to determine the maximum coal concentration which will permit compliance with particulate regulations. The demonstration will continue at this concentration until precipitator improvements can be made.

It is significant to point out that no evidence of coal particle settling has been detected. A 30% COM was stored in a 16,000 barrel tank for six months with continuous operation of the 25 RPM agitator. Samples were withdrawn from the tank at various elevations with no measurable variation in coal concentration. There has also been no evidence of coal settling in process piping, heat exchangers and pump casings.

7.0 Conclusions

Initial startup and feasibility testing of the coal blending and combustion demonstration facilities have been very encouraging. A maximum COM concentration of 42% coal has been blended and pumped to the storage tank with no evidence of excessive coal particle settling or plugging. Although only some of the process equipment has been dismantled for internal inspection, available pressure and flow readings do not indicate excessive wear. There is no evidence of a shift in furnace heat absorption patterns from the standpoint of changes in superheat or reheat steam temperatures. Since the unit was originally designed to burn coal, it is equipped with soot blowers on both the furnace walls and the convection pass. There has been no unusual slag buildup on furnace walls and tube surfaces although this was expected to be the case for a coal designed unit.

There is increased confidence that a stable COM at the 30% coal concentration will not be a problem. This is at least partially attributable to the extensive COM research sponsored by DOE, EPRI and private industry over the last three years.

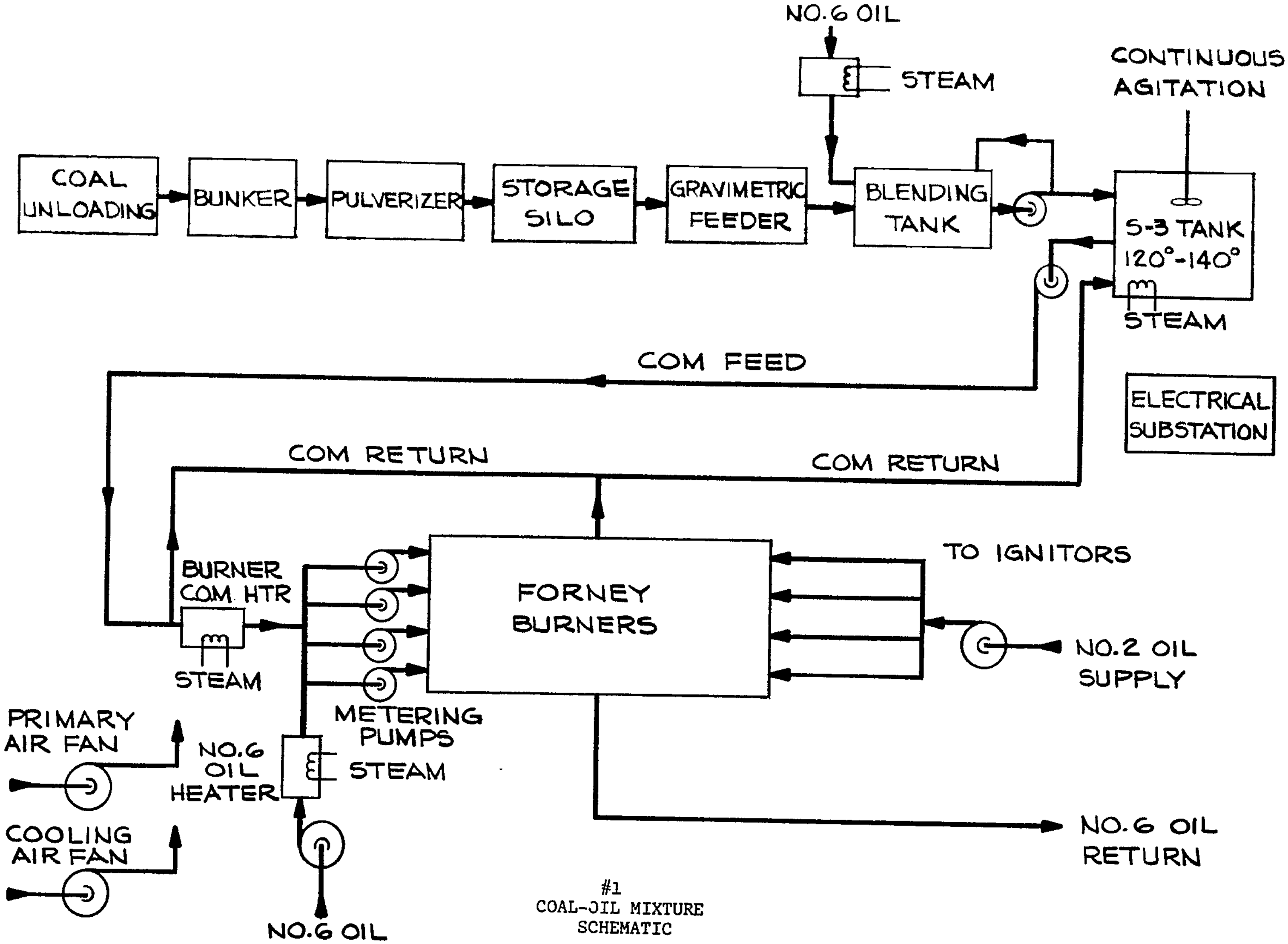
8.0 Acknowledgements

The work described in this paper is sponsored by the U.S. Department of Energy under Contract No. DE-AC22-76E10380. The additional boiler performance testing will be sponsored by the Electric Power Research Institute of Palo Alto, California under Contract Nos. RP-1455-1 and RP-1455-4. The assistance and cooperation of Mr.Andrew M. Wood of New England Power Service Company is also acknowledged.

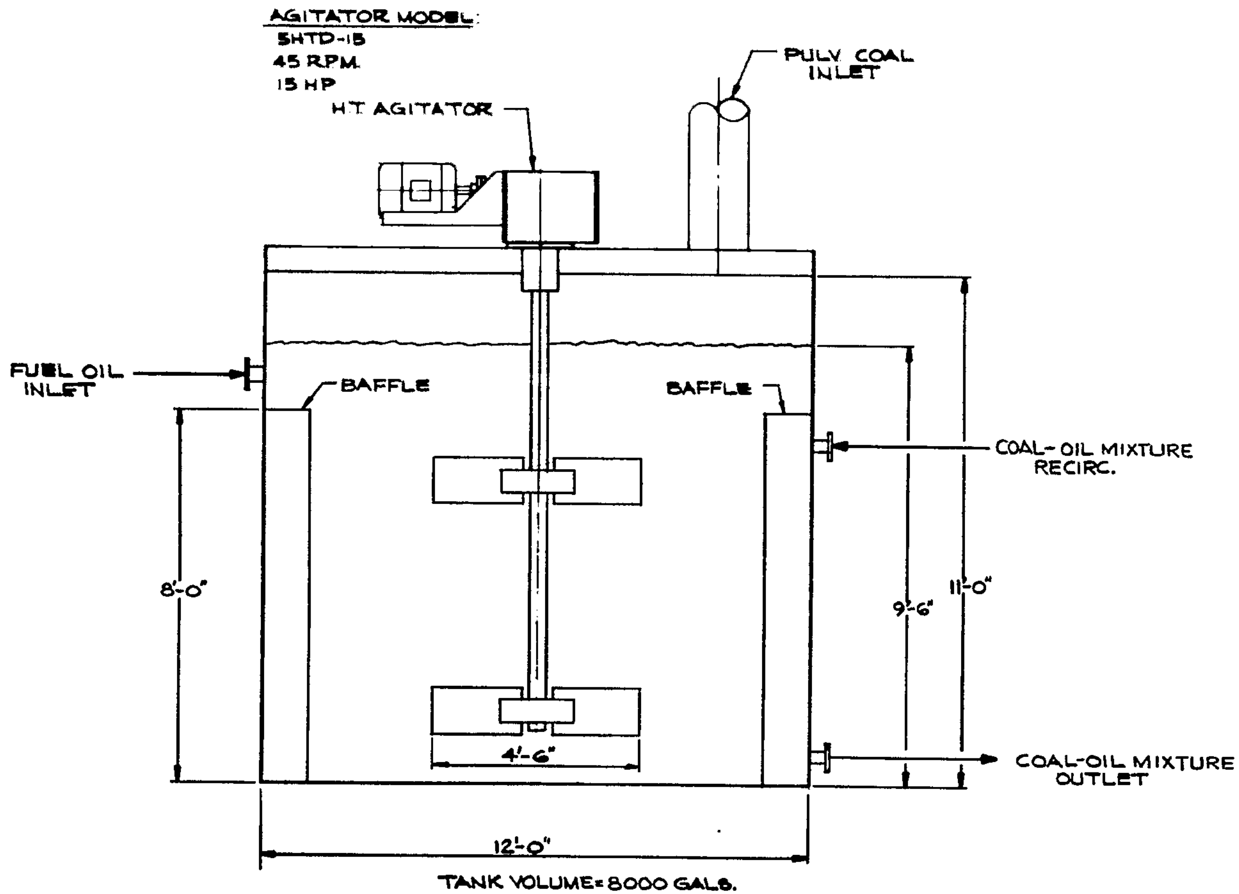
9.0 Bibliographical References

1. C.B. Foster, November 1977. ERDA's Coal-Oil Mixture Combustion Program, 77-WA/FU-4, 1977 ASME Winter Annual Meeting.
2. Y. Nakabayaski, May 1978. Outline of COM R&D in Japan. First International Coal-Oil Mixture Combustion Proceedings, Page 21-27, The MITRE Corp., McLean, VA.
3. H. Whaley, May 1978. Coal-Oil Mixture Projects in Canada. First International Coal-Oil Mixture Combustion Proceedings, Page 28031, The MITRE Corp., McLean, VA.
4. A. Brown, August 1977. Final Report of the General Motors Corporation Powdered Coal-Oil Mixtures (COM) Program. Warren, Michigan, General Motors Corp.
5. J.F. Brakley, A.B. Hersberger and L.R. Burdick, October 1943. Laboratory and Field Tests on Coal-in-Oil Fuels. Trans. of the ASME, April 1944, Page 185-198.
6. C. Moreland, Viscosity of Suspension of Coal in Mineral Oil. Canadian Journal of Chem. Engin., February 1963, Page 24-28.
7. E.M. Rudzki, B.K. Pease, T.H. Weidner. Use of Coal-in-Oil Mixtures to Improve Open-hearth Furnace Performance. Journal of the Institute of Fuel, April 1965, Page 154-165.
8. L. Rodriguez, F. Sell, May 1978. Florida Power Corporation/Drave Corporation Coal/Oil Composite Fuel Program. First International Coal-Oil Mixture Combustion Proceedings, Page 51-65, The MITRE Corp., McLean, VA.
9. R.L. Rowell, S.R. Vasconcellos, J.R. Ford, E.E. Lindsey, C.B. Glennon, S.Y. Tsai, S.K. Batra, May 1978. Investigation and Measurement of Stability in Coal-Oil Mixtures. First International Coal-Oil Mixture Combustion Proceedings, Page 288-299, The MITRE Corp., McLean, VA.
10. J. Dooher, S. Jakatt, S. Batra, R. Dunn, A. Wood, T. Tsai, May 1978. Evaluation of Mixing Equipment and Flow Properties for the NEPSCo Coal-Oil Mixture Demonstration Plant, First International Coal-Oil Mixture Combustion Proceedings, Page 207-221, The MITRE Corp., McLean, VA.
11. A.J. Sabadell, October 1978, Preliminary Evaluation of Commercial Continuous Instrumentation for the Measurement of Viscosity and Coal Concentration for Coal Oil Mixtures. Washington, D.C., The MITRE Corp.

12. P.J. Whalen, F.W.Davies, June 1978. Coal-Oil Slurry Firing of a Utility Boiler. Fredericton. New Brunswick, Canada, New Brunswick Electric Boiler Commission.
13. J.J. Mueller, I.G. Wright, January 1978. Evaluation of Materials for Use in Letdown Valves and Coal Feed Pumps for Coal Liquefaction Service, Columbus, Ohio, Battelle Columbus Laboratories, AF-579.
14. G.S. Wong, September 1978. Coal Slurry Feed Pump for Coal Liquefaction, Canoga Park, Claifornia, Rockwell International, AF-853.
15. O. Foo, November 1979. A Test Plan for the NEPSCo Coal-Oil Mixture Combustion Demonstration Project.
16. K.C. Bachman and C.W. Siegmund, September 1979. The Effect of Fuel Oil Compositon on Particulate Emissions, Exxon Research and Engineering, Linden, New Jersey.



#1
COAL-OIL MIXTURE
SCHEMATIC



#2
COM BLENDING TANK

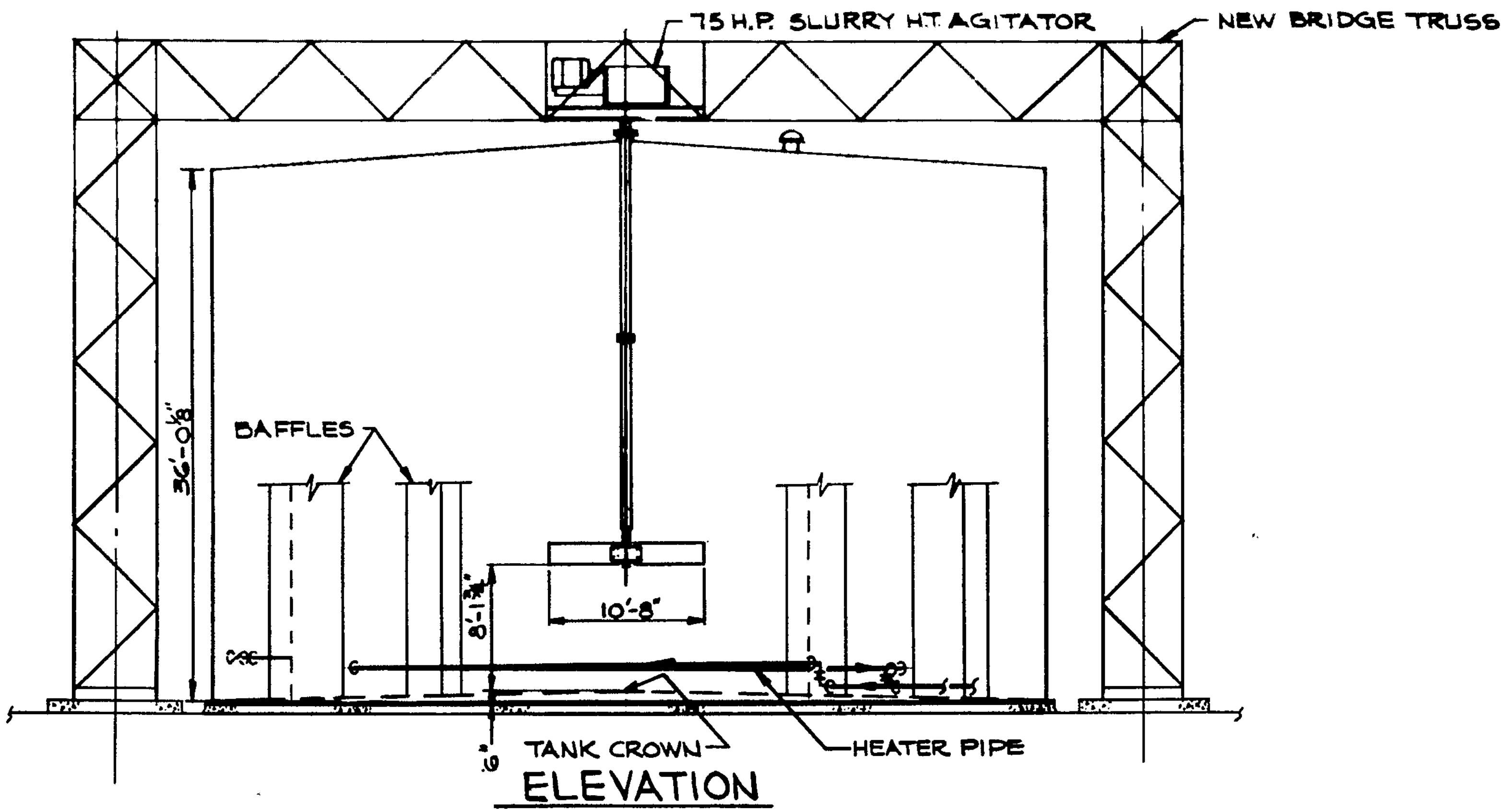

75 H.P. SLURRY H.T. AGITATOR
NEW BRIDGE TRUSS
BAFFLES
36'-0⅛"
10'-8"
TANK CROWN
HEATER PIPE
ELEVATION

Figure 3

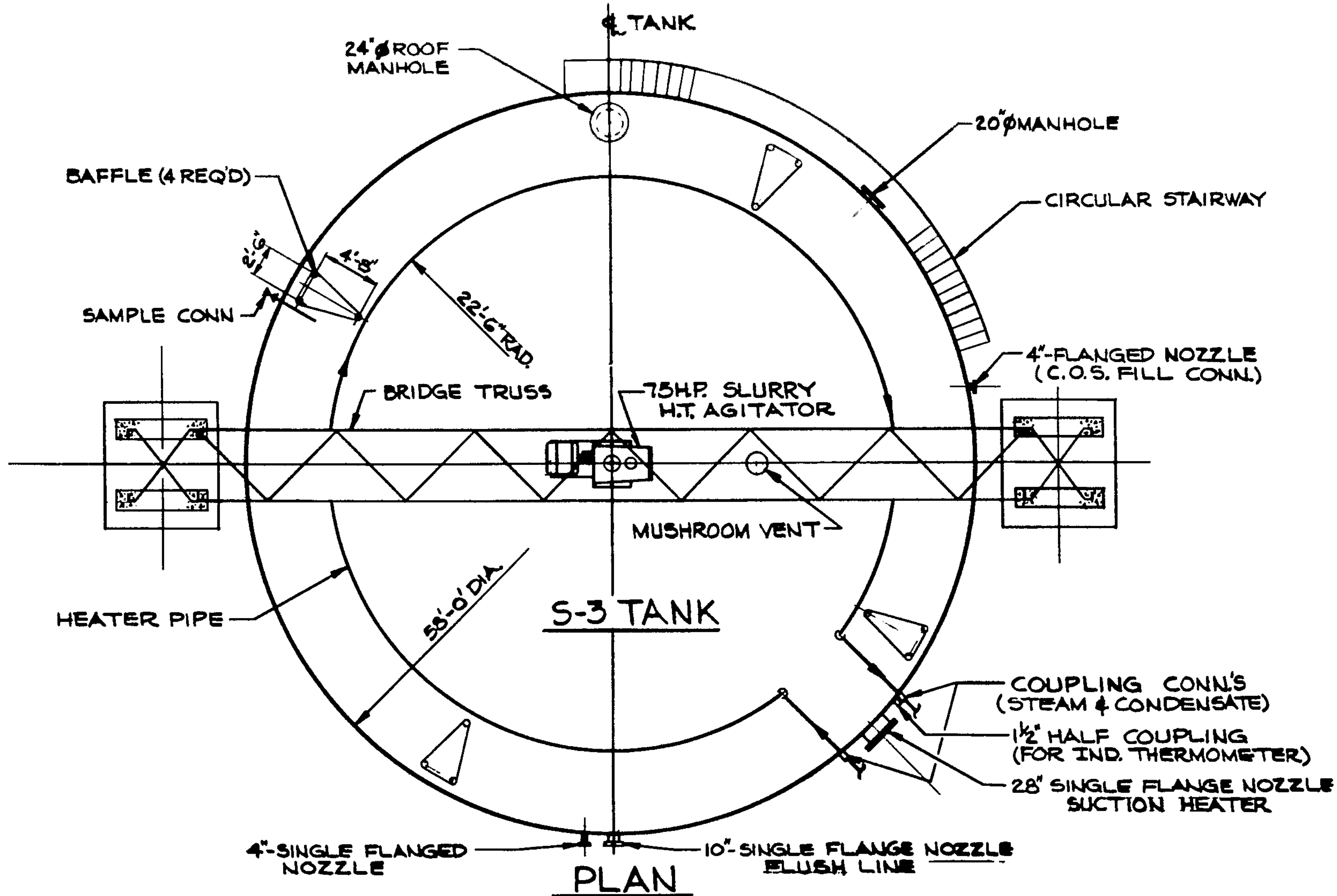

℄ TANK
24"ø ROOF MANHOLE
20"ø MANHOLE
BAFFLE (4 REQ'D)
CIRCULAR STAIRWAY
2'-6"
4'-8"
22'-6" RAD.
SAMPLE CONN
4"-FLANGED NOZZLE (C.O.S. FILL CONN.)
BRIDGE TRUSS
75H.P. SLURRY H.T. AGITATOR
MUSHROOM VENT
HEATER PIPE
58'-0" DIA.
S-3 TANK
COUPLING CONN'S (STEAM & CONDENSATE)
1½" HALF COUPLING (FOR IND. THERMOMETER)
28" SINGLE FLANGE NOZZLE SUCTION HEATER
4"-SINGLE FLANGED NOZZLE
10"-SINGLE FLANGE NOZZLE FLUSH LINE
PLAN

Figure 4

Particulate Emission Tests
Salem Harbor Unit #1

Date	Fuel	Load	Precipitator Inlet lb./million Btu	Precipitator Outlet lb./million Btu	Unburned Carbon	Opacity
9/20/79	10% COM	78 MW		.038		4-7%
10/26/79	20% COM	78 MW		.094		10-20%
2/6/80 (5/16" burner tips)	2.2%S oil	79 MW	.300		60%	10-20%
2/26/80 (¼" burner tip)	2.2%S oil	70 MW	.287		25%	
4/16/80	30% COM	58 MW	1.461	.148		20%
4/17/80	2.2%S oil	75 MW	.353	.107		10-15%
7/26/80 (new precip. rappers)	30% COM	77 MW	1.005	.383		20-50%

Figure 5

OPERATING FLORIDA POWER & LIGHT COMPANY'S SANFORD PLANT ON COAL-OIL MIXTURE

Michael C. Cook

OPERATING FLORIDA POWER & LIGHT COMPANY'S SANFORD PLANT ON COAL-OIL MIXTURE

Michael C. Cook
Florida Power & Light Company
Miami, Florida 33152
United States of America

On April 20, 1980, Florida Power & Light began burning a coal-oil mixture (COM) at its Sanford Power Plant. We believe that mix, which contained 10% by weight coal, was the first COM ever burned in a utility power plant which had originally been designed to be fired with oil. Since April, FPL has continued its Sanford COM experiment, gradually increasing the proportion of coal in the mix. As of August 1, 1980, FPL had successfully burned a COM fuel containing as much as 40% coal. In the weeks ahead, the concentration will gradually be increased to 50%, and we hope to conduct sustained combustion tests at that level.

The purpose of this paper is to describe to you how and why FPL began burning COM at its Sanford plant, and to give you a status report on the project. I will not dwell on the technical details. The Exhibits provide some relevant technical information on the power plant, the fuel, and the COM preparation facility. I will concentrate on the management processes which FPL pursued in successfully undertaking this project so that you may judge whether such an approach might be helpful to others in pursuing significant new energy technologies.

What are some of the unique features of FPL Sanford project?

- The project is being totally funded by Florida Power & Light Company at a cost of over $10 million; no governmental funds are involved.

- Sanford Unit No. 4 is a 400 MW oil-designed plant, and thus provides a large scale commercial test of the technical and economic viability of using COM in an oil utility boiler.

- FPL Management approved the Sanford COM Project on October 12, 1979. It then took us only 6 months to engineer, build and start up a complete COM preparation facility, and to modify the Sanford Plant to handle and burn COM.

·Federal, state and local regulatory agencies gave the Sanford Project their whole-hearted support, issuing approvals and permits as necessary to permit the project to move forward without any regulatory delay.

Some folks, to whom we told this story, seem to think this last point alone constituted a miraculous accomplishment, even if we never burned the first barrel of COM. But, in fact, the test seems to be quite successful. While we have not yet burned enough COM to determine the long term effects, initial results are quite encouraging. It appears at this point that we have a high probability of achieving full power operation of the 400 MW Sanford Unit with a COM mixture containing as much as 50% by weight of coal. Subsequent testing will be needed to determine whether any long-term degradation of plant components or performance is likely to occur.

To help put the Sanford COM Project in perspective, I will first review the history which led up to this experiment. FPL is one of the largest oil-burning utilities in the United States. During 1980 we will use approximately 40 million barrels of oil to generate about 55% of the company's electric sales. This large oil requirement reflects the historical economics of the company's proximity to the oil rich U.S. Gulf Coast and Caribbean oil refineries, and FPL's remoteness from U.S. coal fields. The company's other energy sources are natural gas and nuclear power.

With the rapid increase in oil prices in recent years, the company naturally supported the many national efforts to develop alternative fuels which we might be able to use in our oil-designed power plants. However, as work on these alternatives continued, it became increasingly clear that synthetic fuels for our oil-designed power plants would not be available in significant quantities until the 1990's. We concluded that the most likely substitute fuel which we could burn near-term in our power plants would be a mixture of coal and oil.

COM, however, presented significant uncertainties. Nobody had ever made COM in commercial quantities, and no experiments had been done to determine the effects of COM in oil-designed power plant units. Towards the end of 1979, small amounts of COM had been successfully burned in an oil-designed industrial boiler at the DOE's Pittsburgh Laboratory, and New England Electric System was about to burn some COM in its 80 MW coal-designed plant at Salem Harbor. We concluded that the only way to find out whether COM would successfully burn in one of our oil-designed plants was to try it. But this seemed to be economically unfeasible and presented too high a risk at that time. The problem was stack emissions. Since our plants are not equipped with precipitators (having been designed to burn oil with an ash content of approximately 0.1%), there was no way we could meet emission limits while burning COM. And the outlook for a successful burn of COM was just too uncertain to warrant a $25 million investment for a precipitator on one of our 400 MW units. (Since FPL has nine 400 MW units on its system of similar design, it was concluded that the most meaningful test of COM was to try it in one of these units.)

In September of 1979, discussions with senior personnel at the Department of Energy led them to suggest they might be able to help us avoid the huge cost and multi-year delay associated with adding a precipitator by supporting the granting of an emission variance to permit us to conduct a COM test on one of our 400 MW units. A series of discussions were held with the Florida Public Service Commission, the Florida Department of Environmental Regulation, the EPA, and the DOE which indicated that indeed the granting of such a variance was possible, and that the project would be enthusiastically supported by the concerned governmental agencies.

What was the reason for this active support? The potential benefits were clear. For FPL alone, with nine 400 MW units and four 800 MW units on our system, a conversion to 50% by weight coal COM in those units would permit the company to displace up to 16 million barrels a year of residual fuel oil. Not only would this significantly lessen our company's dependence on foreign oil, but the preliminary economics indicated it could do so at a sufficient fuel savings to more than pay for the extra fuel handling equipment and environmental control equipment which would be needed for permanent conversion to COM. Based on these preliminary indications, FPL management approved a demonstration project for Sanford on October 12, 1979. Three days later, we selected Bechtel Corporation as the engineer/constructor for the COM Project and operator of a COM preparation facility to be located at the Sanford Plant site. It was necessary to build a COM facility from scratch because no one could produce the 10,000 barrels a day of COM product we needed to sustain full power operation of the Sanford Plant. We wanted to move as quickly as possible for a number of reasons, including the need to develop a long term fuel strategy at the earliest possible date, the momentum of the regulatory support which we were getting, and the hoped-for economic benefits of an oil displacement program. Licensing activities, engineering, procurement and site planning were all started immediately. Site work at the Sanford Plant was initiated in mid-November. On November 14, 1979 the Florida Public Service Commission approved a rate-making treatment of the Sanford Project cost which would permit FPL to recover the investment over a proposed one year test program. On January 2, 1980 the Florida Department of Environmental Regulation granted an air emission variance which permitted the use of COM fuel at Sanford without precipitators for 120 full power burn days during a period of up to one year.

The first coal arrived on a newly-built railroad siding at FPL's Sanford site on February 22, 1980. On April 13, 1980 the COM preparation plant was started up and successfully began producing a mixture containing 10% by weight of coal in residual fuel oil. The first burn of this mixture took place in the newly modified Sanford Unit No. 4 power plant on April 20, 1980. The burn was successful, and after testing at that level a program of incremental increases in the coal concentration was begun. By July 18, 1980 the company had achieved a coal concentration of 40% by weight, having overcome numerous operational problems during the progress of the experiment.

Such operational problems were not unexpected. Indeed, we were blessed with a fine project team and plant operating staff whose imagination and initiative were an absolute requisite to the successful performance of the Sanford experiment. A high level of support, enthusiasm and cooperation was necessary from all levels of the Bechtel organization, the FPL organization, and the numerous consultants and suppliers who supported this work. The accelerated schedule on which we were operating lent an air of challenge and enthusiasm to a number of staid industrial organizations around our nation. Items which typically have a one year lead time were often delivered in a matter of days. Did we make mistakes and do rework because we were moving so quickly? No more so than on a "normal" project. On balance, there was probably a net reduction in cost because the rapid progress of the work didn't leave time for overheads to build up, for inflation to take its toll, or for all sorts of "extras" creep into the project.

What have been the tangible results of the experiment to date? We have demonstrated that COM can be made in a continuous process facility using relatively conventional industrial equipment under production conditions. We have shown that COM can be stored and handled at a regular power plant site by utility personnel. We have shown that, with appropriate plant modifications, COM containing 40% by weight of coal, and probably up to 50%, can be burned in a 400 MW oil-designed power plant with no de-rating. We have not yet determined the extent to which corrosion, erosion or other operational problems may be associated with such fuel.

As indicated earlier, the facilities which were used to make and burn COM are described more fully in the attachments to this paper, which also includes specifications for the coal and oil used in the Sanford experiment. Anyone desiring more detailed information than this is requested to wait until our experiment is concluded. We have been moving along at a very rapid pace and have elected not to issue public reports during the progress of the experiment. We do intend at the conclusion of the Sanford project to issue a full report. While certain of the process technology which has been developed during the conduct of this experiment is proprietary, we do expect to be able to publish substantial operating data which will be useful to other utilities considering the burning of COM.

Let me pause and ask the bottom line question. "Will FPL switch to COM?" The answer, at this point, is "I don't know". Even if our continued testing demonstrates that COM does not present any long-term technical problems, there are certain economic penalties associated with burning COM. It requires modifications to the fuel delivery system at each plant, changing of burners, and installation of precipitators and ash handling equipment. In addition, one must either build a large COM preparation facility or contract for such services from a third party. These extra capital and operating costs offset to a great extent the current price differential between coal and oil.

Furthermore, we must remember that coal has less Btu value than oil, so a barrel of COM which is 50% by weight of coal has only up to 40% of the energy provided by the coal, with the remaining 60% coming from the oil. Conversion of a plant to COM only displaces about 40% of the oil used in that plant. Therefore, at today's prices, COM economics are perhaps only slightly better than breakeven and, for some plants, COM may be more costly than oil. But oil prices are likely to continue to rise faster than coal prices, and COM economics will improve in the years ahead. The odds are good that if one were to make a current commitment to COM one would almost certainly come out ahead over time.

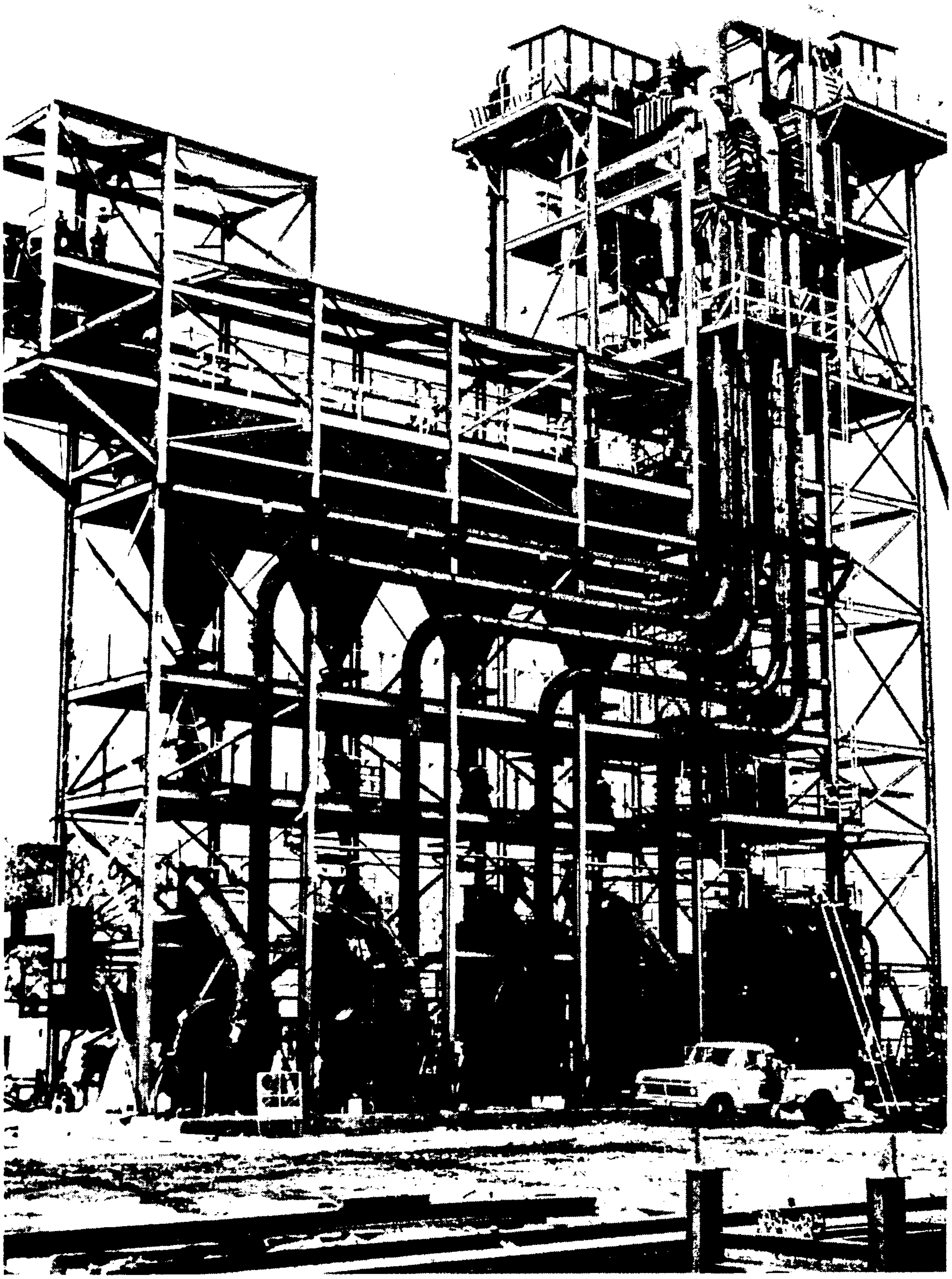
At FPL we are now asking ourselves if, with our successful experience in burning COM, we can eliminate the oil altogether.

The engineers among us know there are several "minor" obstacles to burning coal in an oil-designed plant. One of these is the necessity for a bulk fuel handling facility and all the hardware needed to get the coal to the burner. One of the merits of COM is that one can adapt an existing fuel oil delivery system. This is not possible with bulk coal. A second problem in coal burning is the slagging associated with the coal ash. This problem also exists with COM, but our Sanford experiment seems to indicate that the problem can be overcome at the ash levels experienced in the COM mixture. The third problem is the heat transfer characteristics of coal versus oil, involving radiant versus convective heat and associated boiler design considerations. COM seems to behave enough like oil so we don't have to change our basic boiler configuration.

We are examining each of these problem areas in depth to see if it is at all feasible to engineer modifications tp our fuel as well as our power plant to overcome these obstacles without mixing oil with the coal. Perhaps we can slurry the coal with water instead of oil so as to continue to be able to use our present fuel handling system. Perhaps we can remove enough of the ash from coal so that ash loading is reduced to the levels we are successfully handling in our COM mixture. And perhaps we can modify the characteristics of the coal particles and of the furnace so as to overcome the convective versus heat radiant heat transfer problem. During the months ahead, FPL will be exploring each of these problem areas with engineers, consultants, suppliers and research people. Our objective will be to see if we can come up with some engineered fuel mixes that we might try out as part of our Sanford project. At the same time we will be looking at the longer term engineering and economics of systems which could be used to process and clean fuel and deliver it to our burners in a form compatible with the existing plants so as to minimize required plant modifications.

Where are we headed? I'm not quite sure. We do though have a corporate determination to reduce our use of fuel oil without throwing away our present power plants or spending billions of dollars on their modification. The best hope seems to be in a combination of fuel engineering and plant modification, and this is the approach we're pursuing. We believe that FPL and other private industry participants can get the job done with our own resources. We're moving as quickly as prudent, and I expect that the trade media will be reporting on the results of our work.

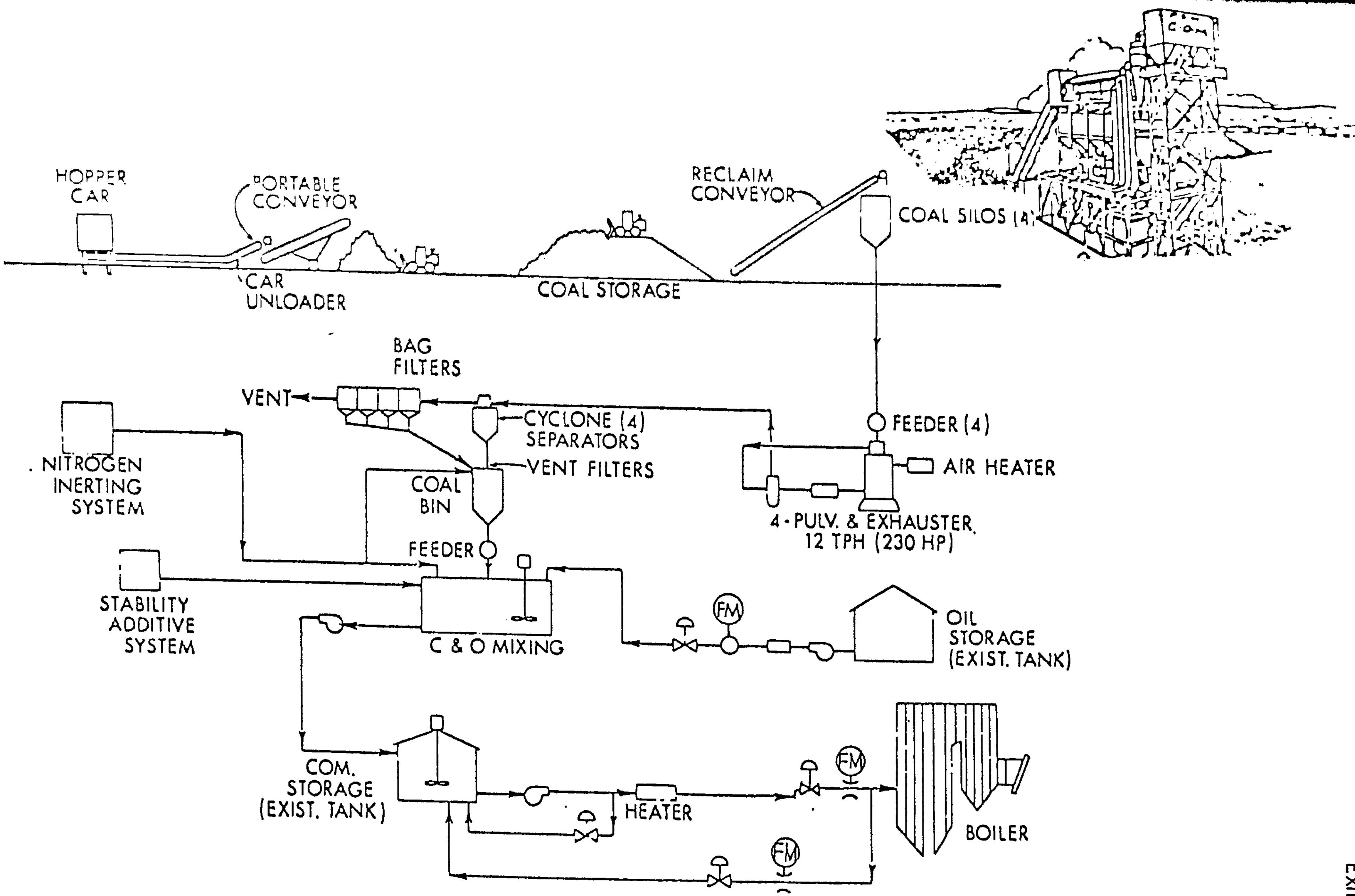
FIGURE 1



FLORIDA POWER & LIGHT COMPANY

COAL-OIL MIX PREPARATION FACILITY

SANFORD PLANT



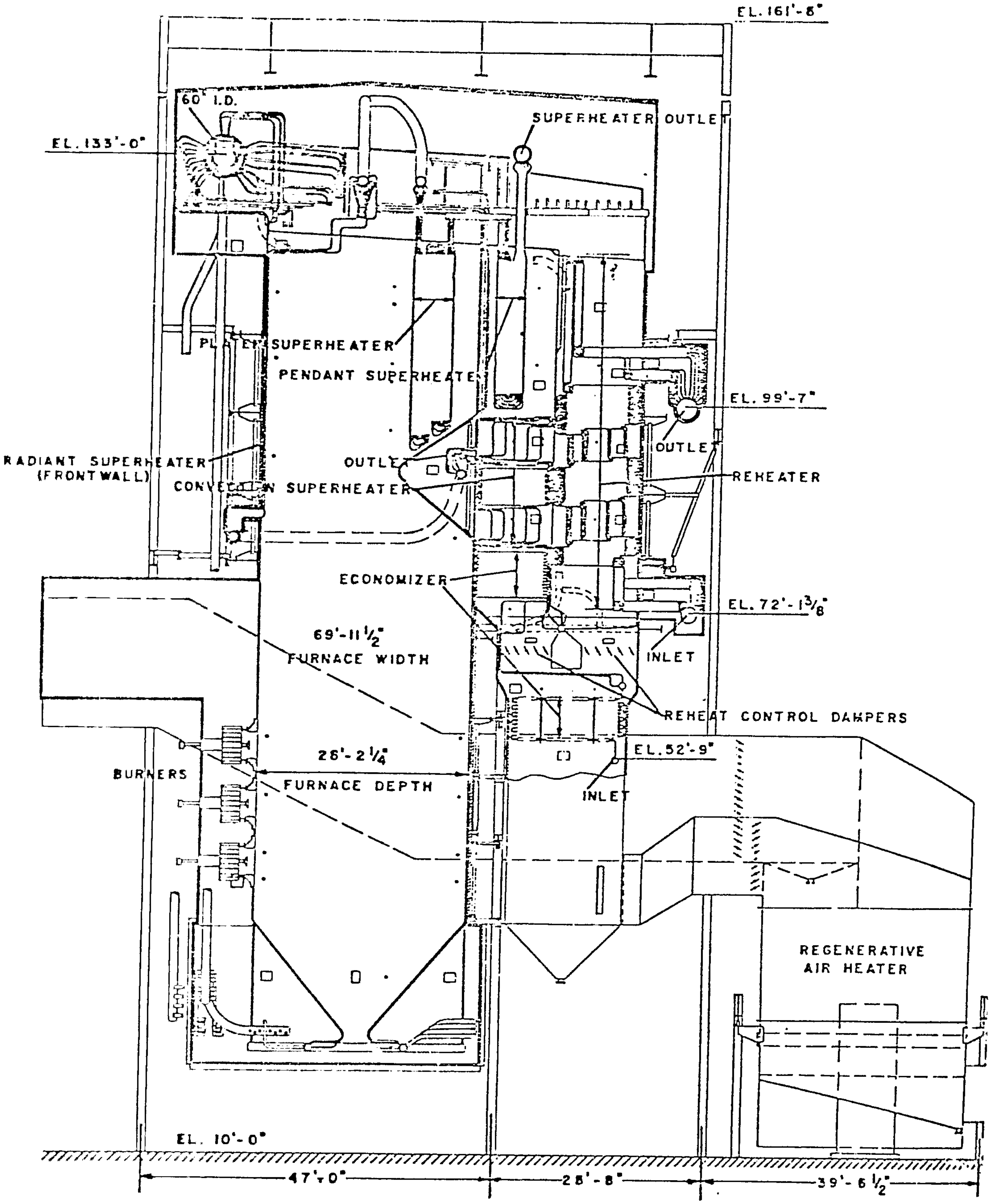
COM = COAL/OIL MIX

FLORIDA POWER & LIGHT CO.
SANFORD STATION
COAL/OIL MIX FACILITY

Exhibit 1

Exhibit 2

FLORIDA POWER & LIGHT COMPANY
SANFORD UNIT NO. 4
STEAM GENERATOR CROSS SECTION



SUMMARY PERFORMANCE SHEET

Purchaser . . . . Florida Power & Light Co. . . . .
Location . . . . Sanford Steam Plant Units No. 4 & 5 . . . .
Design pressure . . . 2850 psig . . . .

Fuel		Oil	Oil	Oil
Steam	M lb hr	1721	2398	2651
Pressure superheater outlet	psi	2460	2460	2590
Temperature steam superheater outlet	F	1005	1005	1005
Pressure boiler drum	psi	2530	2580	2720
Reheat steam	M lb/hr	1500.5	2074	2560 *
Temperature steam entering reheater	F	589	640	655 *
Temperature steam leaving reheater	F	1005	1005	1005
Pressure steam entering reheater	psi	400	558	689
Pressure steam leaving reheater	psi	381	532	657 *
*Includes 1% spray & top heater		out of service		
Temp feed entering unit	F	452	485	425
Temp feed leaving econ.	F	536	578	540
Temp air entering unit	F	170	155	150
Temp air leaving air heater	F	470	503	508
Temp gas leaving furnace	F	1800	1980	2040
Temp gas leaving boiler	F			
Temp gas leaving economizer	F	550	621	631
Temp gas leaving air heater	F	300	315	320
Ditto corrected for leakage	F	293	308	313
Excess air leaving economizer	%	10	10	10
Wet gas entering air heater	M lb/hr	2025	2770	3310
Wet gas leaving air heater	M lb/hr	2163	2930	3483
Air entering air heater	M lb/hr	2035	2756	3276
Air leaving air heater	M lb/hr	1897	2596	3103
Draft in furnace	in. H_2O	.10	.10	.10
Gas side loss thru boiler	in. "	1.05	1.80	2.55
Gas side loss thru suphtr. & rehtr.	in. ")			
Gas loss thru economizer	in. ")	3.36	4.75	6.35
Gas side loss thru airheater	in. "	2.81	4.55	6.10
Gas side loss thru flues (purch.)	in. "	.23	.36	.50
Gas side loss thru dust collector	in. "	1.20	1.85	2.55
Gas side loss total	in. "	8.75	13.41	18.15
Air side loss thru air heater	in. H_2O	2.24	3.45	4.80
Air side loss thru ducts	in. "	1.12	1.85	2.55
Air side loss thru burners	in. "	2.33	3.50	4.85
Air side loss, meas. device	in. "	.56	.95	1.30
Air side loss, stm. coil	in. "	.23	.27	.52
Air side loss total	in. "	6.49	10.12	14.02
Air & gas loss total	in. H_2O			
Pressure loss drum to S.H. out	psi	70	120	130
Fuel burned	M lb/hr	128	174	207
Liberation, total vol.	Btu/hr x cu ft.	12,600	17,200	20,500
Furn. cooling Factor net,	Btu/hr x sq. ft.	55,000	79,600	96,600
Heat Losses				
Dry gas	%	2.50	3.14	3.34
Hydrogen and moisture in fuel	%	5.30	5.44	5.46
Moisture in air	%	.06	.08	.08
Unburned combustible	%			
Radiation	%	.28	.20	.18
Unaccounted for and mfrs. margin	%	1.50	1.50	1.50
Total losses	%	9.64	10.36	10.56
Efficiency	%	90.36	89.64	89.44

Each unit will include the following
Convection surface . . . . 8,080 . . . . sq ft
Walls in furnace . . . . 21,675 . . . . sq ft
Radiant superheater . . . . . 21,800 . . . . sq ft
Convection superheater . . . 59,750 . . . . sq ft
Reheater . . . . 84,900 . . . . sq ft
Economizer (bare tube) . 17,670 . . . . sq ft
Economizer (ext. surf.) . . 82,500 . . . . sq ft
Air heater . . . . 302,000 . . . . sq ft

Total furnace volume . . . . 184,000 . . . . cu ft
Total furnace surface . . . . 41,000 . . . . sq ft
Firing equipment . . . .

Superheat control by . . . .
Performance based on fuel specified below:

SOLID

Kind . . . .
Grindability . . . .(ASTM; D-409-37-T)
Size . . . . Max moisture . . . .

Prox. Analysis Per Cent:
- Moisture . . . .
- Volatile matter . . . .
- Fixed carbon . . . .
- Ash . . . .

Fusion temp. of ash . . . .

LIQUID

Kind . . . .
Gravity API . . . .
Req'd viscosity at burner . . . .
Oil press. at burner . . . .
Oil temp. at burner . . . .
Steam press. at burner . . . .

GASEOUS

Kind . . . .
Sp gr relative to air . . . .
Gas pr at burner in. Hg . . . .
Gas temp at burner F. . . . .

ULTIMATE ANALYSIS

Fuel . . . . Oil . . . .
Per cent by . . . . Weight . . . .
Ash . Vanadium (bal) . . . .01 . . . .
S . . . . 2.75 . . . .
H_2 . . . . 10.30 . . . . . .
C . . . . 85.79 . . . . . .
CH_4 . . . .
C_2H_4 . . . .
C_2H_6 . . . .
CO . . . .
CO_2 . . . .
SO_2 . . . .
H_2O . . . . .20 . . . . . .
N_2 . . . . .50 . . . . . .
O_2 . . . . .45 . . . . . .
Btu/lb dry . . . . as fired . 18,250 . . .
Btu/cu ft at 60 F–30 in. Hg. . . . .

Steam Temp. Control Range From . . . . To . . . . Guarantee Point . . . .

TUBE DATA

WATERWALLS

Front WW Tubes - 3"O.D. x .286"MW SA210A1
Side WW Tubes - 3"O.D. x .286"MW SA210A1
Rear WW Tubes - 3½"O.D. x .400"MW SA210A1
Rear WW Support Tubes - 3½"O.D. x .440"MW SA210A1
HRA Side WW Tubes - 3"O.D. x .286"MW SA210A1
Front WW Risers - 3"O.D. x .286MW SA210A1
Side WW Risers - 3"O.D. x .286MW SA210A1
Rear WW Risers - 3½"O.D. x .340MW SA210A1
HRA Side WW Risers - 3½"O.D. x .340"MW SA210A1
Front WW Feeders - 4"O.D. x .380"MW SA210A1
Front WW Feeders - 5"O.D. x .480"MW SA210A1
Rear WW Feeders - 6 5/8"O.D. x SCH.160 SA106B
Side WW Feeders - 6 5/8"O.D. x SCH.160 SA106B

(165) ECONOMIZER ELEMENTS

C 2¼"O.D. x .260"MW SA210A1
C1 2¼"O.D. x .260"MW SA210A1
C2 2"O.D. x .212"MW SA210A1

STEAM TUBES

(166) CONVECTION S.H. ELEMENTS

A 2½"O.D. x .300"MW SA210A1
B 2½"O.D. x .320"MW Sa210A1
B1 2½"O.D. x .260"MW SA213T2

(221) REHEATER ELEMENTS

D 1 3/4"O.D. x .148"MW SA210A1
E 1 3/4"O.D. x .148"MW SA213T12
F 1 3/4"O.D. x .148"MW SA213T11
G 1 1/2"O.D. x .156"MW SA213T22
H 1 1/2"O.D. x .156"MW SA213T11
G1 1 1/2"O.D. x .165"MW SA213T22

(54) ROWS PLATEN S.H. TUBES

J 2 1/2"O.D. x .340"MW SA213T22
K 2 1/2"O.D. x .240"MW SA213T11
L 2 5/8"O.D. x .480"MW SA213T22

(54) ROWS PENDANT S.H. TUBES

M 2 1/8"O.D. x .270"MW SA213TP304H
N 2 1/8"O.D. x .320"MW SA213TP304H
P 2 5/8"O.D. x .480"MW SA213T22

TUBE DATA (Continued)

(222) PARTITION WALL TUBES

R 1 1/2"O.D. x .188"MW SA213T11
S 2 1/4"O.D. x .280"MW SA210A1

(167) CONVECTION S.H. R.W. TUBES

T 1 1/2"O.D. x .188"MW SA213T11
U 2 1/4"O.D. x .280"MW SA210A1

(444) RADIANT FW TUBES

V 1 3/4"O.D. x .280"MW SA213T22

HEADER DATA

WATERWALLS

Lower Furn. Front Wall Header is 12 3/4"O.D. x 1 5/8"MW Pipe SA106C

Lower & Upper HRA Side WW Headers are 10 3/4"O.D. x 1 1/2"MW SA106C

Economizer Inlet & Intermediate Hdrs. Are 10 3/4"O.D. x 1.125"MW Sa106C

Economizer Outlet Header is 10 3/4"O.D. x 1.312"MW SA106C

Lower Furn. Rear Wall Header is 12 3/4"O.D. x 1 3/4"MW Pipe SA106C

Lower Furn. Side Wall Header is 12 3/4"O.D. x 1 9/16"MW Pipe SA106C

STEAM

Rehtr. Inlet Hdr. 25"I.D. x .850"M.W. SA-515 GR-70

Rehtr. Outlet Hdr. 30"I.D. x 2.169"M.W. SA-378 GR-D

Pendant Outlet Hdr. (2-79-769) 25"I.D. x 5.625"M.W. SA-335 P-22- (2-79-770) 25" I.D. x 5.125"M.W. SA-335 P-22.

Radiant FW Inlet Hdr. 16"O.D. x 1.531M.W. SA-106-C

Radiant FW Outlet Hdr.- 17"O.D. x 1 3/4"M.W. SA335P11

Conv. Roof Inlet Hdr.- 10 3/4"O.D. x 1 1/8" M.W. SA106C

Conv. Intermediate Hdr.- 12 3/4"O.D. x 1.969"MW SA106C

Conv. Outlet Hdr.- 16"O.D. x 2.290"M.W. SA106C

Platen Inlet Hdr. - 17"O.D. x 1 3/4" M.W. SA335-P11

Transition Hdr. - 9"O.D. x 1 1/4"M.W. SA335P11

FLORIDA POWER & LIGHT COMPANY
FUEL SPECIFICATIONS

No. 6 Fuel Oil

Property	ASTM Test Method	LOW ASPHALTENE LOW VANADIUM 2.5% Max Sulfur Min.	Typ.	Max.
Sulfur % by WT.	D-129	1.0	-	2.50
Flash Point Pensky-Martin °F	D-93	150	160-200	-
Pour Point	D-97	-	30-50	.60
Water & Sediment % (See Note 5)	D-95 D-473	-	0.30	2.0
Ash %	D-482	-	0.03	0.10
Viscosity				
Foreign SSU @ 100°F		-	-	-
Domestic SSU @ 100°F	D-88	-	-	-
SSF @ 122°F		-	150	225
Gravity °API @ 60°F	D-287	-	12-13	17
Heat of Combustion MBTU/BBL	D-240	6200	6300	-
Total Metals PPM	-	-	150	-
Vanadium PPM	D-1548	-	50	100
Asphaltenes %	*	-	3.0	5.0

*BRITISH STANDARD BS-4676:1971, MP-143/57

Sanford #4
COM Coal Specification

I. Proximate Analysis

% Fixed Carbon	52.7
% Volatile Matter	32.8
% Ash	8.0
% Moisture	6.5

II. BTU/lb 13200

III. Hardness (HGI) 60

IV. Ash Fusion Temperature (Reducing)

Initial Deformation	2635°F
Softening H=W	2735°F
Hemispherical H=½W	2800°F
Fluid	2800°F

CURRENT STATUS OF DRY FLUE GAS DESULFURIZATION SYSTEMS

M.E. Kelly
J.C. Dickerman

CURRENT STATUS OF DRY FLUE GAS DESULFURIZATION SYSTEMS

M. E. Kelly
J. C. Dickerman
Radian Corporation
Durham, North Carolina
United States of America

Introduction

Radian Corporation is currently conducting a survey of the commercial and developmental status of dry flue gas desulfurzation (FGD) systems in the United States. This project is being funded through the Environmental Protection Agency's (EPA) Industrial Environmental Research Laboratories. The paper presented today will discuss the current commercial status of dry SO_2 control systems and the focus of current research and development (R&D) activities. Also discussed are the possible advantages of dry systems vs. conventional wet lime/limestone systesm. Finally, the possible technical and economic limitations of dry systems are briefly addressed.

Definition of "Dry FGD Systems" Considered

For the purpose of this study, dry FGD is defined as any pollution control system where an alkaline material is contacted with SO_2-laden flue gas and a dry waste product results. This definition excludes several dry adsorption or "acceptance" processes, such as the Shell/UOP copper oxide process or the Bergbau-Forshung adsorptive char process.*

Dry FGD systems can be grouped according to system type: (1) spray dryer based systems with ESP or fabric filter collectors, (2) dry injection systems, primarily with baghouse collectors, and (3) other systems, primarily those where alkaline material is added directly to the fuel prior to combustion such as a coal/limestone combustion system.

*Rockwell's regenerable Aqueous Carbonate Process (ACP) was excluded as no solid waste product results. Rockwell has, however, adapted the open-loop portion of this process for a "throwaway" system.

In the spray drying process, flue gas is contacted with a slurry or solution such that the flue gas is adiabatically humidified and the slurry or solution is evaporated to apparent dryness. For FGD applications the sorbent is often a calcium-based slurry or a sodium solution which reacts with flue gas SO_2 during and following the drying process. The spray dryer can use rotary, two-fluid or nozzle atomization, and the vessel can be anything from the back-mix reactor typically used in conventional spray dryer applications to a large horizontal duct. The dried product salts and fly ash are collected in a downstream fabric filter or ESP.

Dry injection is defined as the process of introducing a dry sorbent into a flue gas stream. This can take the form of pneumatically injecting sorbent into a flue gas duct upstream of the particulate collection device, precoating or continuously feeding sorbent onto a fabric filter surface, or any similar form of mechanically introducing a dry alkaline sorbent to a flue gas stream.

Coal/limestone combustion is defined as the process of burning a mixture of coal and limestone whereby the SO_2 released from the coal reacts with the limestone to form solid calcium salts that are collected with the ash. Two specific combustion processes are currently being developed: one involves burning coal/limestone pellets in a stoker fired boiler, and the other involves burning a pulverized coal/limestone mixture in a low NO_x burner.

Current Status of Dry FGD Technology

Of the three system types, only spray dryer systems have been commercially applied. Ten utility systems (primarily low sulfur coal applications) and four industrial systems have been sold as of May 1980. (Table 1). The spray dryer/baghouse system at Celanese Fibers Company's Amcelle plant in Cumberland, Maryland has passed the State compliance tests for particulate and SO_2. Designed for 1.5 percent sulfur coal, the system has been achieving 85 percent SO_2 removal with a 3 percent S coal. Another industrial system, at Swarthmore Paper, is in the start-up phase. The earliest scheduled utility start-up is the Joy/Niro system at Northern States Power Company's Riverside Station. Start-up is scheduled for the late summer of 1980.

There are currently 13 firms offering a commercial spray drying system. Most of these systems use a lime sorbent: lime is less expensive than sodium alkalis and calcium based product salts present less of a waste disposal problem than sodium-based salts. Most of the systems offer a fabric filter for particulate collection, although an ESP is used in at least one of the utility systems sold (B&W at Laramie River #3). Some vendors claim an additional 10 to 20 percent SO_2 removal across the fabric filter. However, vendors that offer ESPs claim that the gas can be cooled closer to saturation if an ESP is used instead of a fabric filter, resulting in additional SO_2 removal across the spray dryer.

TABLE 1

COMMERCIAL DRY FGD SYSTEMS SOLD TO DATE

Utility

Purchaser	Plant	Size	System Supplier
Otter Tail Power Co.	Coyote, Unit 1	410 MW	Rockwell/Wheelabrator-Frye
Basin Electric Power Coop	Antelope Valley, Unit 1	430 MW	Joy/Niro
Basin Electric Power Coop	Laramie River, Unit 2	500 MW	Babcock & Wilcox
Nothern States Power	Riverside, Units 6 & 7	130 MW	Joy/Niro
Tucson Electric Co.	Springerville, Unit 1	350 MW	Joy/Niro
Tucson Electric Co.	Springerville, Unit 2	350 MW	Joy/Niro
United Power Association	Stanton Station	60 MW	Research-Cottrell
Colorado-Ute Association	Craig, Unit 3	450 MW	Babcock & Wilcox
Platte River Power Authority	Rawhide, Unit 1	250 MW	Joy/Niro
Sunflower Electric Coop	Holcombe, Unit 1	310 MW	Joy/Niro

Industrial

Purchaser	Plant	Size	System Supplier
Celanese Fibers Co.	Cumberland, Maryland	65000 acfm	Rockwell/Wheelabrator-Frye
Swarthmore Paper	Woronco, Massachusetts	40000 acfm	Mikropul
Calgon			Joy/Niro
University of Minnesota		2 units at 120,000 acfm each	Kennecott, Environmental Products Division

In addition to commercial systems in operation or under construction, there are several large-scale demonstration units operating or being constructed. (Table 2). Most of the demonstration test work involves investigation of various parameters such as sorbent type, solids recylce, inlet SO_2, sorbent stoichiometry, spray down temperature and waste solids properties. Some vendors are also privately conducting small-scale test work (1000 to 5000 acfm) aimed at investigating the SO_2/sorbent reaction mechanism, effect of flue gas distribution in spray dryer, and the effect of fly ash alkalinity on SO_2 removal, to help them better respond to bid requests.

The dry injection/baghouse systems have been the subject of numerous past and on-going bench and pilot scale studies. Table 3 lists the current dry injection programs. No commercial systems have been sold to date and few vendors even offer a commercial dry injection system. Fairly high temperatures (600°F+) are required to achieve significant SO_2 removal using lime or limestone. The most reactive sorbents are sodium based (nacholite or trona), resulting in waste disposal problems.

Technologies involving combustion of a coal/limestone fuel mixture are still in the early stages of development. Two of the most promising systems are combustion of coal/limestone pellets and firing of a pulverized coal/limestone fuel mixture in a low-NO_x burner. EPA is sponsoring development of both these technologies.

A 30-day test program with the coal/limestone pellets will start in July 1980. The tests will be conducted at General Motor's Indianapolis plant (60,000 lb steam/hr stoker boiler). This technology is still in the developmental stage and has only been tested on small boilers to date.

The SO_2 removal effectiveness of the coal/limestone mixture fired in a low-NO_x burner is apparently related to the lower flame temperatures present in the low-NO_x burner. The relatively high flame temperatures in conventional burners may result in "glazing" of the reagent particles, leading to significantly lower reactivity.

Dry FGD vs. Conventional Wet (Lime/Limestone) Scrubbing

Technology comparisons between dry and wet scrubbing systems can be drawn in several major areas: waste disposal, reagent requirements, operation and maintenance, energy requirements, and economics. This comparison will focus on general aspects of dry FGD systems as compared to conventional lime/limestone wet scrubbing systems.

TABLE 2

MAJOR SPRAY DRYING DEMONSTRATION ACTIVITIES

Vendor	Location	Size	Comments
Babcock & Wilcox	Pacific Power & Light's Jim Bridger Station	120,000 acfm	Testing in progress
Buell Envirotech/ Anhydro	Colorado Springs-Martin Drake Station	20,000 acfm	Also EPA-funded dry injection program at same location
Combustion Engineering	Northern State Power Sherburne County Unit #1	20,000 acfm	Testing complete
Combustion Engineering	Alabama Power-Gadsden Station (under contruction)	100,000 acfm	Testing to start in May 1980
Ecolaire Systems, Inc.	Nebraska Power- Gerrel Gentlemen Station	10,000 cfm mobile pilot plant	
Research-Cottrell	Colorado Ute Power Comanche Station	10,000 acfm	EPA-funded, tests in progress

With regard to waste disposal, dry FGD systems have an inherent advantage over wet lime/limestone systems in that they produce a dry, solid waste product that can be handled by conventional fly ash handling systems, eliminating requirements for a sludge handling system. However, the waste solids from sodium-based dry FGD systems are quite water soluble and can lead to leachability and waste stability problems. Waste solids from lime spray drying systems and coal/limestone fuel systems should have similar environmental impacts as waste from lime/limestone wet systems, for which waste disposal technology is better defined.

In general, dry FGD systems require a higher stoichiometric ratio of sorbent to entering SO_2 to achieve the desired removal efficiency than do conventional limestone wet scrubbing systems. In addition, the reagents used in spray drying and dry injection systems (soda ash, lime, commercial and naturally occurring sodium carbonates and bicarbonates, such as nahcolite and trona) are significantly more expensive than limestone. Consequently, limestone wet scrubbing systems will have an advantage with regard to both reagent utilization and sorbent-related operating costs.

Several vendors claim that dry systems will have lower maintenance requirements than comparable wet systems. Dry systems require less equipment than wet systems becuase the thickeners, centrifuges, vacuum filters and mixers required to handle the wet sludge waste product from wet systems are eliminated. In addition, slurry pumping requirements are much lower for spray drying and are eliminated in dry injection and combustion of coal/limestone fuel systems. This is important because wet systems have reported high maintenance requirements associated with large slurry circulation equipment. Finally, the scaling potential in limestone wet systems requires extra effort to maintain proper scrubber operation and possibly makes dry systems somewhat more flexible as far as their ability to adjust process operations to respond to variations in inlet SO_2 concentrations and flue gas flow rates.

With regard to energy requirements, dry FGD systems appear to have a significant advantage over wet systems due to savings in reheat and pumping requirements. Spray dryer systems are usually designed to achieve required SO_2 removal while maintaining a 30° to 50°F approach to the adiabatic saturation temperature of the gas at the outlet of the spray dryer. Some systems are designed with warm gas (downstream of the air preheater) or hot gas (upstream of the air preheater) bypass. A small energy penalty is associated with the use of hot gas bypass and reduced overall SO_2 removal results from the use of warm or hot gas bypass.

Energy savings from reduced pumping requirements result from the fact that wet scrubbers may require liquid to gas ratios (L/G) pumping rates of up to 100 gallons per 1000 acfm whereas the L/G for spray drying systems ranges from 0.2 to 0.3.

One of the major driving forces for development of dry SO_2 removal systems is the opportunity for reduction in both capital and operating costs. Although costs are quite site specific, the three types of dry FGD technologies considered here offer several potential possibilities for cost savings. This is due to the reduction in equipment and operation and maintenance requirements relative to conventional wet lime/limestone systems, especially in utility applications. Basin Electric evaluated the costs of the two spray drying systems they have purchased (Antelope Valley and Laramie River Stations) to be about 20 to 30 percent less over the 35-year life of the plant than comparable wet systems. However, it should also be noted that these economics are based on pilot scale data and should be better determined after the operation of commercial systems has begun. The minimal equipment and operating requirements for dry injection systems make the process economically attractive as far as capital costs are concerned, but high sorbent requirements and uncertainties in sorbent availability and cost are slowing further development of the technology on a commercial scale. Capital costs for both the pellet and low-NO_x burner coal/limestone fuel mixture systems should also be low since they will consist mainly of the equipment needed to produce the mixtures. However, since these systems have the potential for impacting the design and/or operation of the boiler, more information on the overall operability of these systems is needed before total operating costs can be estimated.

Technical and Economic Limitations of Dry FGD

Spray Drying

The application of spray dryer technology to higher sulfur coals may be subject more to economic rather than technological limitations. Higher stoichiometries are required for higher sulfur applications. Consequently, the reagent cost differential between lime and limestone may alone make a spray dryer based system uneconomical.

Dry Injection

Major restraints to the development of dry injection technology have been uncertainty in sorbent (nacholite) availability, waste-disposal problems associated with sodium based salts and the relatively high flue gas temperatures required to achieve high SO_2 removals with calcium based sorbents.

Combustion of Coal/Limestone Fuel Mixture

To date, only preliminary data exist for the SO_2 control effectiveness and operation of boilers firing either coal/limestone pellets or a pulverized coal/limestone fuel mixture. Further research on a larger scale for both systems is needed to determine the effects of firing a coal/limestone fuel mixture on boiler operation and maintenance. Effects of the increased particulate loading, and the degree of SO_2 removal achievable also need to be investigated further.

TABLE 3

CURRENT DRY INJECTION PROGRAMS

Vendor/Agency	Location	Size	Comments
Buell Envirotech	Colorado Springs - Martin Drake Station	3000 acfm	Testing in progress, EPA funded
DOE/Grand Forks Energy Technology Center	GFETC Labs	200 acfm	Testing complete Report expected in July 1980
DOE/Pittsburgh Energy Technology Center	PETC Labs	500 lb coal/ hr furnace	Testing in progress
EPRI	Public Service Company of Colorado- Cameo Station	20 MW_e	Tests underway

GUIDELINES FOR COPING WITH THE POWERPLANT AND INDUSTRIAL FUEL USE ACT OF 1978

Jan Benes Vlcek

GUIDELINES FOR COPING WITH THE POWERPLANT AND INDUSTRIAL FUEL USE ACT OF 1978

Jan Benes Vlcek
Gardner, Carton & Douglas
1875 Eye Street, N.W.
Washington, D.C. 20006
United States of America

Introduction

1. Coverage
   (a) The Powerplant and Industrial Fuel Use Act ("FUA") applies to new and existing powerplants and new and existing Major Fuel-Burning Installations ("MFBI's")

   (b) Powerplants and MFBI's are defined as boilers, internal combustion engines (MFBI's only), gas turbines, or combined cycle units

2. Use of petroleum or natural gas in new powerplants or new MFBI's is prohibited unless an exemption is granted.

3. Use of natural gas is prohibited in existing powerplants after January 1, 1990 unless an exemption is granted.

4. The Secretary of the Department of Energy ("the Secretary") may, at any time, prohibit use of natural gas or petroleum in existing powerplants on a case-by-case or categorical basis unless such plant qualifies for an exemption.

5. The Secretary may, at any time, prohibit use of petroleum or natural gas in an existing MFBI that is capable of using coal or an alternate fuel, unless such installation qualifies for an exemption.

NOTE: Final regulations issued in June 6, 1980 *Federal Register* generally cover all Parts except those pertaining to existing MFBI's and powerplants. Cites for those Parts are to interim regulations contained in July 23, 1979 *Federal Register*.

<u>MAJOR REGULATORY PROVISIONS IMPLEMENTING THE POWERPLANT AND INDUSTRIAL FUEL USE ACT OF 1978 WITH RESPECT TO MAJOR FUEL-BURNING INSTALLATIONS</u>

1. Definition of an MFBI - (Subpart 500.2)
   (a) stationary unit consisting of:
      (1) boiler
      (2) gas turbine unit
      (3) combined cycle unit (combination of steam and combustion turbine with input to steam turbine provided by exhaust from the combustion turbine) or
      (4) internal combustion engine

   (b) size of units covered - according to FUA: (Subpart 500.5)
      (1) individual units of 100 million Btu's/hr heat input rate or larger
      (2) 50 million Btu's/hr units located at same site in combinations equalling 250 million Btu's/hr or larger

2. Units are considered to be "in combination" if:
   (a) contributing to same product
   (b) part of same plant or facility
   (c) physically connected to common energy or power system
   (d) any other factors exist indicating functional integration

3. For purposes of the 250 million Btu threshold count:
   (a) all existing units at a site over 50 million Btu's/hr
   (b) all new units at a site over 50 million Btu's/hr
   (c) if combination of new and existing units at a site, only existing units over 100 million Btu's/hr and new units over 50 million Btu's.

4. New or Existing - Transitional Units

   "New" - Construction or acquisition after the date of enactment of FUA, November 9, 1978, and includes installations on which construction or acquisition began after April 30, 1977, unless they have proceeded to the point where they could not comply with applicable requirements of FUA without incurring significant operational or finanical detriment.

"Existing" - Any units not covered by above provisions.

"Reconstruction" - An existing MFBI that is reconstructed will become new if expenditures for current year and preceding two calendar years equal or exceed 50% of the expenditure for an "equivalent replacement unit." Refurbishment or modifications of a unit which does not increase useful life or annual fuel consumption and is solely for the purpose of increasing fuel burning efficiency will not constitute reconstruction.

5. Primary Energy Source -
   (a) means the fuel used by any facility except minimum amounts of fuel, not to exceed 15%, unless otherwise demonstrated, of the total annual Btu heat input of the unit for ignition, startup, testing flame stabilization or control
   (b) minimum amounts necessary to prevent equipment outages or emergencies

Prohibitions

1. New:
   (a) <u>MFBI's consisting of boilers</u> - petroleum or natural gas as a primary energy source (Subpart 503.3(a))
   (b) <u>Other MFBI's</u> - prohibited from using natural gas or petroleum as a primary energy source if Economic Regulatory Administration ("ERA") issues categorical rules or case-by-case orders (Subpart 503.3(b))
   (c) <u>Exclusions</u> - alternate fuels (coal, solar, petroleum coke, industrial wastes, wood, renewable and geothermal sources), unmarketable gas (including low Btu gas), marginal well gas, natural gas and petroleum up to 5% of total annual Btu output for unit ignition, startup, control and flame stabilization.

2. Existing (Final Regulations not issued) coverage limited to "coal capable" units. ERA may prohibit the use of natural gas or petroleum by an MFBI (or "category" of MFBI's), if ERA finds:

   (a) MFBI has the technical capability (elements physically necessary to sustain combustion and maintain heat transfer) to use coal or other alternate fuel (electricity or any fuel other than natural gas and petroleum) or

   (b) MFBI previously had such capability and could again have such capability without substantial physical modification or reduction in design capacity (additions of appurtenances such as pollution control and fuel handling equipment are not usually considered in the determination of such technical capability) and

   (c) coal or alternate fuel use is financially feasible

3. Existing - Mixture Capability

If technically and financially feasible for MFBI's to use mixtures of alternate fuel with natural gas or petroleum, ERA may prohibit use of natural gas or petroleum in excess of minimum requirements for reliability and efficiency. (FUA maximum requirement for alternate fuel use in such mixtures is 75%).