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COAL—AN ALTERNATE FUEL FOR THERMAL ENHANCED OIL RECOVERY STEAM GENERATION

Final Report

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Prepared for the Department of Energy
Under Contract No. EX-77-C-01-2418

Date Published—August 1979

Arthur G. McKee & Company
Cleveland, Ohio

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COAL—AN ALTERNATE FUEL FOR THERMAL ENHANCED OIL RECOVERY STEAM GENERATION

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0.0 SUMMARY

This project evaluates technical and economic aspects of using coal to replace lower tier crude oil as fuel for thermal enhanced oil recovery steam generation in Kern County, California. Replacing this oil by an alternate fuel would increase the amount of oil available for sale by approximately 30 percent. The intent of the project is not to select a particular technology to be used, or to provide an economic comparison of the technologies selected for evaluation, but rather to present technical and economic aspects associated with each technology.

The primary consideration in the evaluation is compliance with emission limitations as proposed in the California Air Resources Board (CARB), Model Rule on April 26, 1978.

Three coal technologies were evaluated:

1. Coal-Oil Mixtures (COM)

2. Coal Gasification

3. Fluidized Bed Combustion (FBC)

Coal-Oil Mixtures and Coal Gasification were evaluated for retrofit application. Fluidized Bed Combustion was evaluated from the standpoint of providing replacement or future steam generating capacity.

The use of Coal Oil Mixtures would require the installation of pollution control equipment on the steam generators to reduce emissions of sulfur oxides and nitrogen oxides in order to meet the proposed CARB limitations. Particulate collection would also be required. Operation of the steam generators in conjunction with

this equipment to meet the proposed CARB limitations may present a potential problem. Removing ash from steam generators may present an additional problem, since steam generator design does not readily lend itself to modification to allow for this removal.

Uncontrolled emissions of sulfur oxides and nitrogen oxides resulting from the use of the gas, produced by the proposed Coal Gasification plant, to fire steam generators would be below the proposed CARB limitations. Therefore, no flue gas cleaning would be required.

The use of FBC steam generators would require pollution control equipment in order to meet the proposed CARB limitation for nitrogen oxide. Particulate collection would also be required. Uncontrolled emissions of sulfur oxides would be below the proposed CARB limitations and therefore scrubbing would not be required.

Implementation was based on a synthesized field four miles in diameter and containing one hundred-50 million Btu per hour steam generators. Economics were developed on the basis of providing fuel or steam generating capacity to satisfy this criteria. It is recognized that the economics associated with each technology are not competitive with current operations using lower tier crude oil. World oil pricing would most likely be required to make these technologies economical.

It must also be noted that the results of the economic analyses do not lend themselves to direct comparison. Whereas the Coal-Oil Mixture and Coal Gasification programs would utilize existing steam generators, and hence are evaluated in terms of additional production costs, the Fluidized Bed Combustion program entails replacement or addition of steam generators. In this case, the calculated production cost more closely represents a total steam production cost. In addition, it must be emphasized that Coal Gasification exhibits greater economy of scale than either Coal-Oil Mixtures or Fluidized Bed Combustion. On a scale smaller than the synthesized field, it is expected that the incremental cost per thousand pounds of steam associated with these two technologies would be relatively close to that of the synthesized field, whereas this cost for Coal Gasification would increase. On a scale larger than the synthesized field, it is anticipated that costs associated with all three technologies would decrease.

Based on the assumptions detailed in the Basis of Evaluation discussion, there is an additional cost of producing a thousand pounds of steam from implementation of new coal based systems. For Coal-Oil Mixtures this figure is approximately \$3.11. For the alternative of Coal Gasification the added cost is \$6.00 per thousand pounds of steam. Implementation of Fluidized Bed Combustion in the synthesized field yields a total steam production cost of approximately \$6.36 per thousand pounds of steam. The operating costs and fixed charges contributing to these costs are shown in Table 1. In terms of dollars per barrel of oil available for sale, these costs are equivalent to \$5.69, \$9.00, and \$9.54 respectively.

Continued operation of existing steam generators would require a capital expenditure for pollution control equipment in order to meet the proposed CARB limitations. In order to estimate the cost that would have to be incurred to bring an equivalent existing operation into environmental compliance, a Base Case was established. It is estimated that this cost would add approximately \$0.48 to the cost of producing a thousand pounds of steam. In terms of dollars per barrel of oil available for sale, this would represent an addition of \$1.03 to the selling price per barrel. Applying these figures as credits to the coal based system costs results in a net additional cost of producing a thousand pounds of steam of \$2.63 for Coal Oil Mixtures, \$5.52 for Coal Gasification and \$5.88 for Fluidized Bed Combustion. These costs are summarized in the following table:

Added Cost of Steam Production

Technology	Cost ($ per 1000 lbs Steam) Coal Based System	Base Case Environmental Control	Net
Coal-Oil Mixtures	3.11	0.48	2.63
Coal Gasification	6.00	0.48	5.52
Fluidized Bed Combustion	6.36	0.48	5.88

In terms of dollars per barrel of oil available for sale, these costs are equivalent to $4.66, $7.97 and $8.51 respectively. In all cases the added cost includes the cost of new fuel rather than its incremental cost over the existing fuel cost structure. In other words, no credit was taken for cost of fuel currently being used.

The results clearly indicate, however, that none of these technologies are economically competitive with the current regulated price in effect on oil produced from existing operations. An incentive in the form of a higher allowable price for oil produced from enhanced techniques would be required to make such ventures economically viable.

The use of coal in this application, however, could have far greater impact than only that associated with increasing the amount of oil available for sale. Coal Gasification and Fluidized Bed Combustion offer significant environmental advantages over the current operation, in terms of lower uncontrolled emissions. These reductions could provide significant emission off sets, which in turn would allow substantial expansion of EOR operations or even expansion in other industries. In this respect, the cost associated with these technologies would have to be evaluated in terms of the potential expansion in operations that could be allowed to occur as a result of their implementation.

ADDED OPERATING COSTS AND FIXED CHARGES

COST ($ PER 1000 LBS STEAM)

ITEM	COM	COAL GASIFICATION	FBC
Raw Materials (Not Incremental)	1.82	2.73	2.35
Other Operating Costs (Variable)	.46	.37	.50
Labor (Semi-Variable)	.08	.14	.22
Ad-Valorem Taxes and Insurance	.08	.30	.40
Total Incremental Operating Costs	2.44	3.54	3.47
Capital Charges	.67	2.46	2.89
Total Cost	3.11	6.00	6.36

TABLE 1

1.0 INTRODUCTION

1.1 Purpose of the Project

The decline of domestic petroleum production in the United States, along with the rising demand for petroleum products, will present significant future strains on the national economy. Dependence on imported oil continues to increase as domestic reserves diminish at a rate greater than can be compensated for by new expansions and discoveries.

One of the near term options for substantial new domestic energy is Enhanced Oil Recovery (EOR). It is estimated that in existing developed oil fields, approximately 70 percent of the original oil-in-place remains unrecovered after production by conventional methods has been exhausted. EOR will allow recovery of a portion of this reserve.

The near term recovery of this oil reserve is presently constrained by major economic, environmental and technical difficulties. Among these are price regulations, environmental regulations and the technical and economic feasibility of EOR techniques. In an effort to assist industry in eliminating some of these problems, the Department of Energy (DOE) has initiated an EOR Research and Development Program. The primary goal of this program is to significantly increase the national oil production from existing reservoirs through the application of EOR technology.

Presently, approximately 30 percent of the crude oil produced by thermal enhancement is consumed in generating steam required for the process. Replacement of all or a portion of this crude oil by an alternate fuel would have considerable impact on the 1985-1990 production goals. Consistent with the goals of the EOR R&D program, as well as the national energy policy, the present project is investigating the technical aspects of various coal technologies that could be used in this application, as well as the associated economics for implementation. It is recognized that relative to today's heavy oil market, along with current price regulations, the economics associated with the technologies being evaluated are not competitive with the present method of operation. Much of the oil produced by the steam flood EOR technique is presently classified as lower tier oil. Uncontrolled oil pricing would most likely be required to make these technologies economical. Therefore, one of the objectives of the economic evaluation will be to give an indication of the oil pricing incentive required to make these technologies economical.

1.2 Basis for Evaluation

Six items had to be evaluated in order to establish the bases for the project. These items were:

1. Location for evaluation
2. Emission Regulations
3. Applicable technologies
4. Fuel production or steam generating capacity required
5. Source and cost of raw materials
6. Basis of economic analysis

The major portion of the projected thermal recovery production will occur in California. Approximately 50 percent of this production is anticipated to take place in Kern County. In view of these statistics, Kern County, California was chosen as the location for evaluation.

The major problem facing the current thermal recovery projects, as well as future projects in Kern County, is compliance with environmental regulations. The California Air Resources Board (CARB) has designated Kern County as a non-attainment area for sulfur dioxide and total suspended particulate matter. CARB has proposed a model rule for controlling emissions of sulfur oxides and oxides of nitrogen from fossil fired steam generators having a heat input greater than five million Btu's per hour.

The proposed rule would impose the following emission limitations on affected sources:

1. After September 1, 1980, owners and operators of steam generators which were constructed prior to September 1, 1978 would be required to limit sulfur oxide and nitrgoen oxide emissions from such sources to 200 parts per million and 150 parts per million by volume, respectively. Sources subject to these provisions would be required to comply with specified increments of progress.

2. Sources which are constructed after August 31, 1978 would be required to limit sulfur oxide and nitrogen oxide emissions to 60 parts per million and 100 parts per million by volume, respectively, beginning on the date the steam generator(s) commence operation.

These proposed regulations were used as the applicable emissions standards for the project.

Two philosophies for coal utilization in this application were investigated. The first philosophy was to examine technologies that could be retrofitted to existing steam generators. The second philosophy was to examine technologies that could be used for replacement or future expansion capacity.

Three coal technologies were selected for evaluation:

1. Coal-Oil Mixtures (COM)
2. Coal Gasification
3. Fluidized Bed Combustion (FBC)

Coal-Oil Mixtures and Coal Gasification were evaluated for retrofit application. Fluidized Bed Combustion was evaluated from the standpoint of providing a direct coal fired steam generator which could be used to provide replacement or future expansion capacity. The primary consideration in evaluating these technologies was the emissions resulting from their application. The technology must be able to meet the proposed CARB standards either uncontrolled or through application of available control technologies.

In order to determine a fuel production or a steam generating requirement, a common basis for evaluation was established. This was done by creating a synthesized field. This field encompasses an area four miles in diameter and contains one hundred-50 million Btu per hour steam generators. When operated at 80% utilization, this would represent an annual production capability of approximately 3.5×10^{10} pounds of steam per year.

Coal reserves are available in Colorado, Wyoming, Utah and New Mexico. These reserves are in a radius of 500 to 1000 miles of Kern County. Rail facilities are available to deliver coal from these areas to Kern County. Delivered coal costs were developed on this basis. A Utah coal was used for process evaluation. The analysis of this coal is shown in Figure 1.

Cost of limestone to be used in the Fluidized Bed Combustion section of project was developed on the basis of delivery to Kern County by rail from near Las Vegas, Nevada. The analysis of this limestone is shown in Figure 2.

UTAH COAL ANALYSIS

	PERCENT
Moisture	8.5
Carbon	69.0
Hydrogen	5.1
Nitrogen	1.3
Sulfur	.5
Ash	8.5
Oxygen	7.1
	100.0

Btu/lb 11,570

Ash Fusion Temperature:

I.D.: 2300°F

Fluid: 2435°F

Figure 1

FLINTKOTE LIMESTONE ANALYSIS

	PERCENT
Calcium Carbonate	92.0
Magnesium Oxide	1.7
Aluminum Oxide	.3
Iron Oxide	.1
Acid Insol.	1.7

Figure 2

The Economic Analysis is based on the project internal rate of return (IRR). IRR is determined by the discount factors for uniform compounding of the project cash flow over each year of the study period. The study period starts at the time of release of the engineering contract and formulation of the financing plan. It continues through the construction period and ends in 2001. The cash flow used is for the hypothetical case of total equity financing (no debt or interest payments). The IRR is that interest rate which, when applied to the future cash flows to discount them to their present value (1979), makes the cumulative discounted cash flow equal to zero at the end of the study period (2001). In this method of analysis, the retention value of the business (book value of fixed assets plus working capital) at the close of the final year of the study period is taken as a cash inflow to close the analysis.

In the evaluation of each of the three alternate fuel options for enhanced oil recovery, the IRR is assumed to be 15 percent after tax, while the future annual cash flows are unknown. These flows can be developed however, by calculating the product (i.e., steam) selling price, given the production volume. Since this price represents a cost that must be added to the present cost of producing steam an incremental analysis is implied. However, the fuel cost for the present operating system is not known. Therefore, in the Economic Analyses that follow the incremental steam price covers the fuel cost, all incremental operating costs and incremental capital charges, which yield an after-tax IRR of 15 percent on the incremental investment.

Each analysis is performed in constant 1978 dollars. No inflation or escalation is applied to succeeding years.

The McKee Venture Analysis computer program is used to develop cash flows for the years included in the study period (1979-2001).

The program output includes a sensitivity matrix of the internal rate of return (IRR) for changes in individual input variables as discussed later.

The evaluation of the three alternate fuel options are based on certain criteria, some of which result from McKee's experience in the study and engineering of similar type operations. The major assumptions are listed below. Unless noted otherwise, these are common to all cases.

a. The Economic Analysis is based on certain incremental and non-incremental capital and operating costs; i.e., costs that are over and above those experienced in the operation of the 100 existing oil-fired steam generators. All figures are given in 1978 dollars and are accurate to within ±25 percent. The following costs were excluded from the capital cost estimates:

   - Site preparation.

   - Property.

   - Cost of investment capital.

   - Railroad track work.

   - Rolling stock such as railroad tank cars, coal cars and tank trucks.

   - Royalties and commissions.

   - Permits, leases, titles, concessions, licenses, easements, and rights-of-way necessary for the performance of the work.

- Operator's training and start-up.

b. A cost credit is not taken for the cost of fuel oil burned in the existing boilers, as the true cost per barrel of this oil is not known.

c. Pollution abatement equipment when required for SO_x, NO_x and/or particulate control for the 100 steam generators, is included in the capital cost estimates.

d. Continued operation of existing steam generators would require capital and operating expenditures for installation and operation of emission control equipment in order to meet proposed CARB limitations. The cost associated with the equipment required to bring an existing operation equivalent to the synthesized field into environmental compliance for SO_x and NO_x is estimated as follows:

	Capital Cost	Operating Cost ($/Yr)
SO_x	$26,400,000	3,200,000
NO_x	$ 5,800,000	4,750,000
Total	32,200,000	7,950,000

Capital and operating costs are based on using double alkali scrubbing for SO_x control and noncatalytic ammonia injection for NO_x control. These costs add $0.48 to the cost of producing a thousand pounds of steam. In terms of dollars per barrel of oil available for sale, this would represent an addition of $1.03 to the selling price per barrel.

e. The synthesized field generates approximately 3.504×10^7 thousand pounds of steam per year. Based on an 80 percent utilization rate, this amount of steam would annually produce

23,360,000 barrels of oil. This figure was arrived at by calculating the amount of oil that would be required to generate 3.504×10^7 thousand pounds of steam. Since the current method of operation produces approximately 3.3 barrels of oil per barrel burned, this fuel consumption was multiplied by a factor of 3.3 in order to estimate the volume of oil for sale.

f. Implementation of either Coal Gasification or Fluidized Bed Combustion would result in making the entire oil production volume of 23,360,000 barrels per year available for sale. Implementation of Coal-Oil Mixtures would save approximately 40 percent of the oil presently being used to fire steam generators. Therefore, use of COM would yield 19,155,200 barrels per year for sale.

g. Incremental operating costs are calculated from the following unit prices for labor and raw materials:

o	Coal:	\$40/ton; delivered
o	Lower Tier Oil:	\$5.00/bbl; delivered (for COM plant use only)
o	Limestone:	\$15.00/ton; delivered
o	Electricity:	\$.035/KWH
o	Labor (nonconstruction):	\$11.00/hr

h. The sum of the years digits accelerated depreciation schedule used on major plant and equipment items are as follows:

Pollution Control Equipment	5 years (straight line method)
COM Plant	15 years
Other Plant and Equipment	20 years
Pipeline Distribution System	20 years

Steam Generators	25 years
Buildings	40 years (straight line method)

i. Each option qualifies for a 20 percent investment tax credit applied to 80 percent of the total equipment capital cost.

j. Property taxes for the Bakersfield area are equivalent to 3.2 percent of the total plant cost.

k. Land can be acquired at zero capital cost.

l. A total of two man-years of technical assistance are provided during the first two years of plant operation.

m. In the analysis of Coal Gasification and Coal-Oil Mixtures, the existing steam generators are assumed to last through the study period (1979-2001).

2.0 COAL-OIL MIXTURES

2.1 Technical Issues

The application of Coal-Oil Mixtures, COM, in enhanced oil recovery steam generation was evaluated from the standpoint of producing a fuel that could be used to fire existing steam generators.

The COM concept is based on preparing a liquid fuel, containing an appreciable fraction of pulverized coal, which can be burned in existing combustors with a minimum of modification. The resulting fuel would have characteristics which are weighted averages of the constituent fuels, for most properties. Consequently, COM would have a heating value, ash content and sulfur content which are between those of the coal and oil of which it is composed. Applying this concept results in reduced oil consumption. Using COM containing a coal concentration of 50 percent by weight would reduce oil consumption by approximately 40 percent.

The major issues that must be considered in the use of a COM fuel concern stability, handling, preparation and storage, combustion characteristics and selection of process equipment. The term "stability," as applied to COM, refers to the rate at which coal settles out of the mixture. A stable COM would be one that exhibits minimum settling over a given period of time. The level of stability and methods used for stabilization can vary considerably depending upon the application. Long term stability would be required in applications that require long storage periods prior to consumption. Shorter term stability could be tolerated in applications where consumption keeps pace with production.

Stability can be accomplished through the use of additives, by very fine pulverization of coal particles or by continuous agitation. Additives fall into the categories of emulsifying agents, gelling agents or surfactive agents. Stabilization by the use of additives is achieved in two manners. Coal is either prevented from settling out of the mixture or, coal is allowed to settle out of the mixture but is prevented from hard packing. In the latter case, the COM must be remixed or agitated to uniformity prior to combustion.

Stabilization by fine pulverization requires pulverizing coal to particle sizes of less than 15 microns. Pulverizing to this size requires increased horsepower, which increases cost. Commercial coal grinding equipment can reduce coal particle sizes in the range of-200 mesh to -325 mesh. The generally accepted practice for grinding coal to be used in COM fuels is to reduce coal to particle sizes of approximately 80 percent -200 mesh. Stabilization by fine pulverization requires pulverizing coal to particle sizes of less than 15 micron, which is approximately five times finer than 200 mesh. Pulverizing to this size requires increased horsepower, which increases costs. The economics of grinding beyond 80 percent -200 mesh, relative to the gain in stability however, have not yet been determined.

Stability can also be accomplished by continuous agitation. This method keeps coal in suspension by continuous remixing. When this approach is used, a system must be provided to flush fuel lines with oil during periods of shutdown. This is done in order to prevent coal from settling in the lines.

Stability is very important in that it can effect handling and combustion. Settling of coal in pipe lines could cause plugging problems. Non-uniform COM could cause incomplete combustion. Due to its critical nature, a good engineering approach to insure stability may be to use additives in conjunction with continuous agitation.

The most economical approach to a COM fuel is to achieve the highest coal concentration that is technically feasible. This concentration is limited by the viscosity of the prepared fuel. As the coal concentration is increased, the relative viscosity of the COM increases to a point where pumping would become difficult. This factor limits maximum coal concentration to approximately 50 percent by weight.

Viscosity is also a function of temperature. COM exhibits a temperature profile very similar to that of its base oil. As temperature increases, viscosity decreases. In order to decrease handling problems, COM should be heated during storage. Additional heating may be required prior to combustion. The temperatures required depend on the characteristic of the base oil. Using a No. 6 oil as a base would require storage temperatures in the range of 120°F to 140°F, with heating to approximately 160°F to 180°F prior to combustion.

Tests conducted on a package oil fired boiler have shown that COM can be burned using conventional combustion equipment. Combustion characteristics for COM composed of 30% to 50% coal and No. 6 fuel oil were very similar to those of oil alone. The flame stability and envelop were as good as that of oil alone. Emissivity was equal or higher than that of oil, and high coal carbon burnout was achieved. Air atomized burners performed better than steam atomized burners. Steam atomized burners had a tendency to plug due to their smaller ports. No evidence of slagging or ash buildup was noted. It was estimated that 99% of the ash input to the boiler went out the stack. Nitrogen oxide emissions were higher than when using oil. This was partially attributed to the higher nitrogen level of the coal. Particulate emissions were also higher due to the ash in the coal. No continuous high load operation was witnessed, and therefore the effect on boiler efficiency was not determined. This testing covered a period of approximately 750 hours. Although results were positive, longer test periods would be required to determine the effect on boiler efficiency and potential long term ash problems.

A major consideration in the selection of process equipment for preparation and utilization of COM is the abrasive nature of slurry fuels. To date, only short duration tests have been performed using standard equipment. Burner nozzle erosion may be a long term maintenance problem. Internal atomizing nozzles, where atomization takes place before the fuel and atomizing media leave the nozzle, are highly susceptible to erosion. External atomizing nozzles appear to provide some relief in this area. Pumps, valves, piping and instrumentation are also subject to the same erosion problem. As previously indicated, test programs have been of short duration. The effect that erosion will have on equipment reliability will have to be determined by longer term test programs. The Department of Energy is presently involved in several Coal-Oil Mixture demonstration projects. The objectives of these projects are to demonstrate COM in industrial and utility applications and to address the technical issues associated with its preparation and use. When these projects reach the actual demonstration phases, many questions should be answered relative to long term reliability of equipment. At present, however, erosion should be recognized as a potential problem and therefore should be a major consideration in the design of piping systems and selection of process equipment.

2.2 Retrofitting Steam Generators

The two major items that must be considered when retrofitting steam generators to burn COM concern (1) emission control requirements and (2) modification to allow for ash removal.

Estimates of emissions resulting from combustion of a COM containing 50% by weight Utah coal and 50% by weight lease crude oil with analysis shown in Figure 3, indicate that sulfur oxide and nitrogen oxide emissions would have to be reduced by approximately 73% and 67% respectively to meet the CARB limitations for existing steam generators. Particulate collection would also be required.

The reduction required for sulfur oxide could be achieved by the use of scrubbers. Both double alkali and single alkali scrubbers can achieve removal efficiencies greater than the 73% required in this application. Inasmuch as both types have the same capability, selection of which approach to use would have to be based on capital and operating costs, along with operating conditions. In general, the following comparisons can be made:

1. Capital costs for double alkali scrubbers are at least twice the cost of single alkali scrubbers for comparable sizes.

2. Water requirements for double alkali scrubbers are considerably less than those of single alkali scrubbers.

3. Waste disposal is less of a problem with double alkali scrubbers, since the greatest percentage of it is water insoluble. Accordingly, waste disposal costs should be lower.

It is conceivable that the higher capital cost for double alkali scrubbers could be offset by lower operating costs when a comparison is made. For the purpose of making an economic evaluation of the overall project, we have elected to use double alkali scrubbing. This was based primarily on lower water requirements and easier disposal of solid waste.

The generally accepted methods of reducing NO_x emissions are through combustion modifications or flue gas treatment. Reduction by combustion modification is accomplished by reducing flame temperature or oxygen availability. These techniques are most effective

LEASE CRUDE FUEL OIL ANALYSIS

	PERCENT
Moisture	0.05
Carbon	86.40
Hydrogen	11.00
Nitrogen	.67
Sulfur	1.30
Ash	.07
Btu/lb	18,300
Btu/gal	151,000
Gravity, API @ 60°F	11.7

Figure 3

in reducing NO_x generated thermally, but do not significantly reduce NO_x formed by nitrogen in the fuel. The only modification technique that could be used in this application would be minimum excess oxygen firing. Since the major portion of the NO_x generated in this application would come from nitrogen in fuel, it is estimated that this modification would only reduce NO_x emissions by approximately 15 percent.

Catalytic and non-catalytic ammonia injection processes are available to treat flue gas in order to reduce NO_x emissions. Both approaches react ammonia with NO_x to produce nitrogen and water. Both approaches also require a narrow and critical temperature range for optimum effectiveness. The catalytic process requires a temperature range of approximately 600°F to 800°F, while the non-catalytic process requires a temperature range of approximately 1700°F to 1800°F.

Catalytic ammonia injection does not readily lend itself to steam generator application. It would be difficult to provide the temperature range required by this process. In addition, it is anticipated that the concentration of SO_x and particulate in the flue gas would have adverse effects on the catalyst.

Non-catalytic ammonia injection is more applicable to steam generator design. Tests conducted on a 50 million Btu per hour steam generator indicate that temperatures at the junction of the radiant and convection sections were favorable for this process. NO_x reductions averaging 70% were achieved. It must be emphasized that this test was of short duration, under controlled conditions and on a specific unit. It is not known if these results could be duplicated on other units under actual operating conditions. In general, non-catalytic ammonia injection has reduced NO_x by over 90% under laboratory conditions, but reductions in commercial applications have been less than 70 percent.

Using combustion modifications and ammonia injection, it would appear that technically, under controlled conditions, reductions of NO_x emissions in excess of the 67% required in this application could be achieved. These reductions, however, were achieved under relatively controlled conditions. Due to the nature of the operation, steam generators in oil field use require broad operating margins. This must be taken into consideration when attempting to apply results from other applications to this application.

The second major item that must be considered is modification to the steam generators to allow ash removal. Duplication of the previously described test results in this application would result in accumulation of approximately forty pounds of ash per day in the steam generator. Even with this result, modifications would have to be made to allow for ash removal. It is likely that modification could be made to the convection section of the steam generator to accommodate soot blowing equipment, but the design of the steam generator radiant section does not readily lend itself to this modification. One approach that may be used would be the installation of a lance type soot blower in the bottom of the radiant section which would traverse the length of the steam generator. The objective would be re-entrain deposited ash or move the ash to one end of the steam generator for removal. A collection hopper would have to be installed at one end of the steam generator to receive the ash. The feasibility and success of this approach is speculative.

As can be seen from the previous discussion, burning COM in an environmentally acceptable manner would require scrubbers, non-catalytic ammonia injection and fly ash collection. Installation of this equipment would have considerable impact on the current method of operation. Slagging, ash removal and potential steam generator derating may present additional problems. In order to make an economic evaluation, we have assumed that emission limitations could be achieved by the methods discussed and capital and operating costs were developed accordingly. In addition we have assumed that a steam generator could be modified to allow ash removal.

2.3 Implementation

The plan for implementing COM in the synthesized field calls for one preparation plant to service the field. This plant will be located on the perimeter of the field. It is assumed that COM would be distributed to the steam generators using an existing oil distribution system.

The COM preparation plant for the project is a continuous process designed to produce approximately 27,900 barrels per day (405.6 million gallons per year) of stable COM with a 50% by weight coal concentration. This production requires approximately 3000 tons of coal per day, 15,800 barrels of crude oil per day and 155 barrels of stabilizer/ additive per day. The analysis of the coal and oil used in producing this COM are shown in Figures 1 and 3 respectively.

The material flow diagram and equipment requirements for the preparation plant are shown in drawing 4598-A-081478-1. Drawing 4598-A-081478-2 shows plot plans and elevations for the preparation plant.

Run of mine coal is delivered to the preparation plant site by rail and is discharged to below grade track receiving hoppers. It is discharged from these hoppers by vibrating feeders and transported to the storage yard by belt conveyors. Crude oil is delivered to the site and pumped to oil storage tanks. Stabilizer/additive is delivered to the site by truck and pumped to additive storage tanks.

Coal is reclaimed from live storage by two underpile hoppers equipped with vibrating feeders. Reclaimed coal is transported by belt conveyor to the preparation area. This area consists of two identical COM preparation trains. Coal received in this area is discharged into two storage silos. It is withdrawn from the storage silos by vibrating feeders and conveyed to roller mills where it is subsequently pulverized to approximately 80% - 200 mesh. Pulverized coal is then classified by a cyclone collector. Classified pulverized coal is discharged to surge bins by means of rotary feeders. Coal, oil and additive are then metered, in the correct proportions, into mixing tanks. These tanks are equipped with agitators which mix the coal, oil and additive to the desired specification. COM is then pumped from the mixing tanks to storage tanks. These tanks are also equipped with agitators in order to insure stability during storage.

2.4 Economics

This section contains the results of an economic analysis of implementing Coal-Oil Mixtures (COM) in the synthesized field. This field would generate approximately 3.504×10^7 thousand pounds of steam per year.

The Economic Analysis, as explained in the Introduction, is based upon capital costs (including pollution control equipment) which are over and above those of the existing facilities. Similarly, the operating costs used are not the total costs for the production of steam but rather, the cost to cover only those items related to the new facilities. This includes the total fuel cost, however.

2.4.1 Conclusions

Using delivered costs of $40.00 per ton of coal and $5.00 per barrel of lower tier oil at the COM plant, an added price of $3.11 per thousand pounds of steam is necessary in order for the COM operation to obtain an after-tax IRR of 15 percent. Assuming a net 19,155,200 barrels of oil annually available for sale from the enhanced recovery program, the above price equates to $5.69 added to the cost of producing one barrel of oil for sale. This price is based upon a plant investment of $72.9 million and a working capital requirement of $12.6 million.

The sensitivity of the IRR to a change in the value of a specific input variable is measured by the index of sensitivity. For the variables listed below, the indices of sensitivity are as follows:

Variable	Index of Sensitivity (Percentage Point Change in IRR Per 1 Percent Change in Value of Input Variable)
Incremental Price	.51
Incremental Investment	.09
Incremental Variable Costs	.34

The internal rate of return is, therefore, most sensitive to the incremental price of steam. Its index value indicates that a one percent change in steam price effects a change of 0.51 percentage point in the IRR.

2.4.2 Input Variables

The following paragraphs detail the values that are used as inputs to the computer program for the COM implementation analysis.

The COM plant is assumed to attain its expected production level approximately two years after start-up. Costs and other data relating to operation at rated capacity are as follows:

2.4.2a Production Volume and Operating Schedule

Production costs are based on an annual steam production of 3.504×10^7 thousand pounds of steam. The plant operates 347 days per year.

2.4.2b Added Oper ting Costs

The estimated annual additional operating costs associated with the COM plant and related pollution control equipment on the existing steam generators are based on the labor and raw material costs detailed in the Basis of Evaluation discussion. These estimates are considered to be attainable at full production.

Added Operating Costs

Variable Costs	($1000)	Cost ($ per 1000 lbs. Steam; Approx.)
Raw Materials at COM Plant (Not Incremental)	63,908	1.82
Coal $35,666		
Oil 23,117		
Additive 5,125		
Other Operating Costs (Variable) (Includes repair and maintenance, utilities, supplies and consumables, ash handling (sludge) and NO_x, SO_x and particulate control)	16,027	.46
Total Variable Costs	79,935	2.28
Semi-Variable Costs		
Labor	2,650	.08
Semi-Variable Expenses		
Ad-valorem taxes and insurance	2,916	.08
TOTAL ADDED OPERATING COSTS	85,501	2.44

2.4.2c Capital Cost Estimates

The installed cost of the COM facilities and related pollution abatement equipment for the existing boilers is estimated to be $72.9 million in 1978. This estimate is accurate to within ±25 percent and is itemized as follows:

Year	-3	-2	-1	Start-Up 0	Total
	(Millions of Dollars)				
Buildings	-	1.9	1.9	-	3.8
Pollution Control	8.4	16.7	8.4	-	33.5
COM Plant & Equipment	8.9	17.8	8.9	-	35.6
	17.3	36.4	19.2		72.9

2.4.2d Incremental Working Capital

Working capital is provided to cover current cash requirements, accounts receivable, inventory of operating materials, as well as accounts payable.

Cash is sufficient for a four-month payroll, accounts receivable are 17 percent of sales (60 day payment), accounts payable are 17 percent of operating costs excluding labor (60 day payment), and inventory volumes are sufficient for annual needs.

Incremental Total Working Capital = 12 Percent of Incremental Sales.

2.4.3 Internal Rate of Return

The added price of steam generated from the COM-fired steam generators is $3.11 per thousand pounds of steam in order to obtain the assumed 15 percent after-tax IRR.

The individual components of that price are as follows:

Added Operating Costs and Fixed Charges

Item	($1000)	Cost ($ per 1000 lbs. Steam; Approx.)
Raw Materials (Not Incremental)	63,908	1.82
Other Operating Costs (Variable)	16,027	.46
Labor (Semi-Variable)	2,650	.08
Ad-valorem Taxes and Insurance	2,916	.08
Total Incremental Operating Costs	85,501	2.44
Capital Charges	23,473	.67
Total Cost	108,974	3.11

Annual capital charges noted above are required to obtain an IRR of 15 percent after tax on the added investment. These total approximately 32 percent of the total capital cost discussed earlier in this section.

2.4.4 Sensitivity Analysis

The following is a summary of variables including incremental steam price, investments, and costs as they affect the sensitivity of the internal rate of return for the proposed project.

2.4.4a Incremental Price

An incremental price of $3.11 per thousand pounds of steam was calculated to be the price "most likely" to yield an after-tax IRR of 15 percent.

To gauge sensitivity, this price was varied from a low of $2.47 per thousand pounds of steam to a high of $3.73 per thousand pounds of steam. The corresponding values of the IRR was -2.0 and 23.7 percent respectively.

Incremental Price

$/1000 lbs. Steam	IRR, %
2.47	-2.0
3.11	15.0
3.73	23.7

2.4.4b Incremental Investment

The incremental plant and equipment investment for the proposed COM alternative has a "most likely" total of $72.9 million spent over an approximate three-year period. This total was varied by 25 percent on the high and low side to yield investment totals from $54.7 million to $91.9 million. The resulting IRR values were 18.7 and 12.8 percent respectively.

Incremental Investment

Plant Capital Cost	IRR, %
$54,700,000	18.7
72,900,000	15.0
91,100,000	12.8

2.4.4c Incremental Variable Cost

Variable cost includes all items noted in the discussion of operating costs, with the exception of the semi-variable components. A 25 percent factor was applied for the high and low incremental variable cost values, resulting in a range of 1.71 to 2.85 per thousand pounds of steam. The corresponding IRR values were 23.5 and 1.2 respectively.

Incremental Variable Cost

Cost per 1000 lbs. Steam	IRR, %
$1.71	23.5
$2.28	15.0
$2.85	1.2

As can be seen from the table of Added Operating Costs and Fixed Charges, the major portion of the variable cost is attributed to the cost of raw materials and in particular coal. Figure 4 presents the sensitivity of steam cost as a function of the delivered price of coal. Varying the price of coal by 25 percent above and below the value of $1.72 per million Btu used in the study yields costs per thousand pound of steam of $3.37 and $2.86 respectively.

Coal Cost

Delivered Price ($ per Million Btu)	Steam Cost ($ per 1000 lbs)
1.29	2.86
1.72	3.11
2.15	3.37

COAL–OIL MIXTURES
SENSITIVITY TO COAL COST

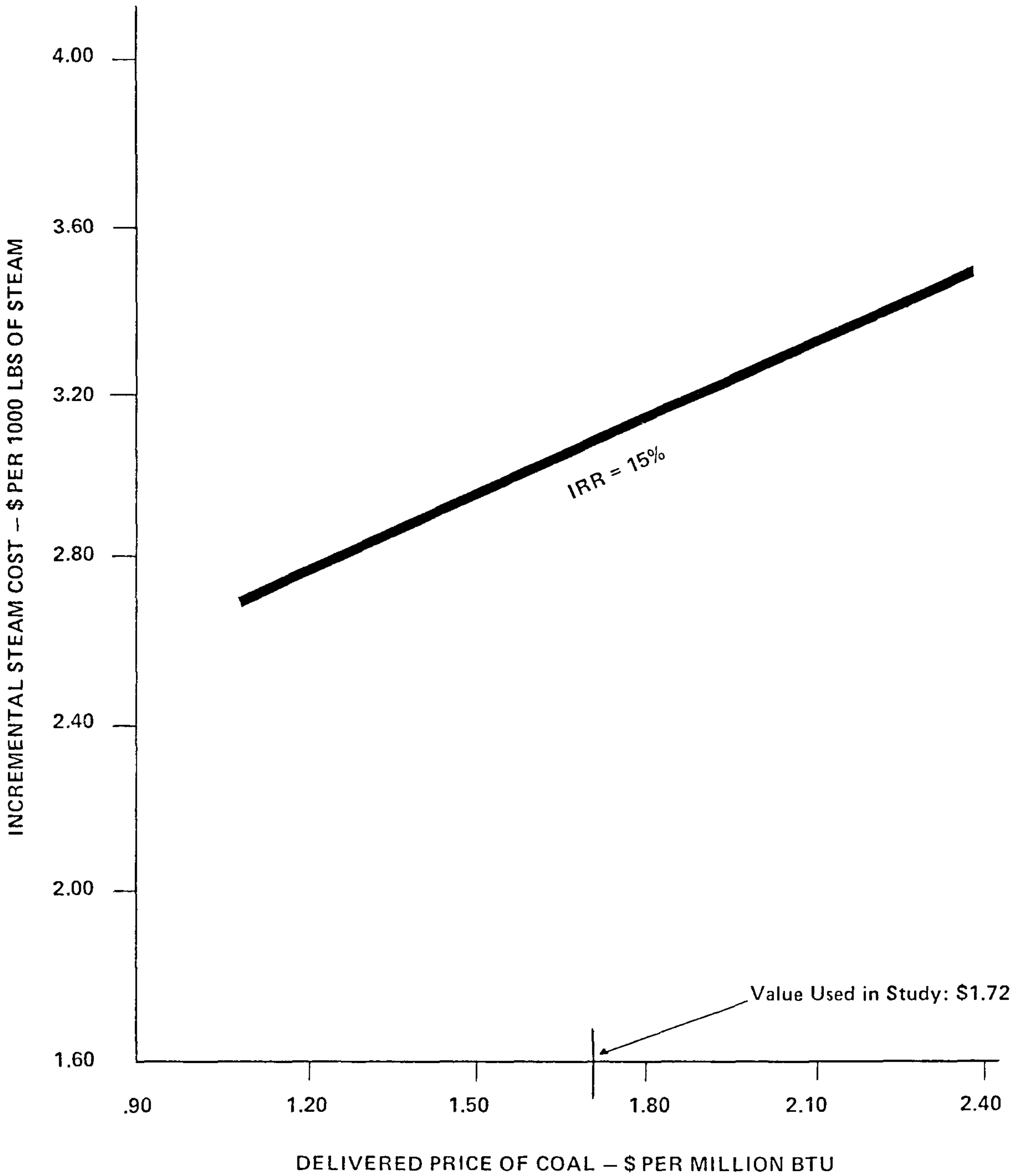

FIGURE 4

2.4.4d Index of Sensitivity

The sensitivity analysis can be summarized by noting that an index of sensitivity can be calculated for each of the variables discussed earlier. The larger this index number, the more sensitive the IRR is to a percent change in that variable.

Mathematically, the index of sensitivity can be expressed as:

$$I = \frac{IRR}{\left(\frac{V_2 - V_1}{V_1}\right) \times 100}$$

IRR = Absolute change in IRR obtained by varying a specific input variable (e.g., selling price)

V_1 = Initial input variable value

V_2 = Final input variable value

The value "I" gives the change in IRR units per percentage change in the specific variable.

Using this equation results in the following indices of sensitivity:

Variable	Index of Sensitivity, I
Incremental Price	.36
Incremental Investments	.09
Incremental Variable Costs	.22

The Internal Rate of Return is, therefore, most sensitive to the incremental price of steam. A one percent change in that price will result in a change of 0.36 in the IRR.

2.5 Demonstration

Demonstration of Coal-Oil Mixtures could be accomplished by using COM to fire one steam generator. Demonstration in this manner would require construction of a COM preparation plant and installation of emission control equipment on the steam generator.

The following is an estimate of the cost of this program based on constructing a preparation plant designed to produce a 50/50 COM to fire one 50 million Btu per hour steam generator. Capital costs include costs such as engineering, equipment, installation and construction of the COM preparation plant and emission control equipment. Direct operating costs were developed on the basis of one year's operation at 80% utilization. This cost covers items such as raw materials (coal, oil, additives and chemicals), labor, utilities, repair and maintenace and waste disposal.

Capital Cost	$2,320,000
Direct Operating Cost	1,260,000
TOTAL DEMONSTRATION COST	$3,580,000

3.0 COAL GASIFICATION

3.1 Technical Issues

The application of Coal Gasification in enhanced oil recovery steam generation was evaluated from the standpoint of producing a low-Btu or medium-Btu gas which could be utilized to fire existing steam generators.

The selection of a gasifier technology was based on the ability of the technology to satisfy the following three basic criteria.

1. The gasifier must be able to accept coal fines.
2. The gasifier could be operated without producing by-product tars.
3. The technology must be demonstrated and commercially available.

After evaluating the various gasifiers that are commercially available, the Winkler process, offered by Davy Powergas Inc., was chosen for the project. In addition to satisfying the three base criteria, the Winkler process provides the option of utilizing either air or oxygen in the gasification reaction. With this option, it would be possible to demonstrate the technology on a small scale using air, and subsequently switch to oxygen for larger scale implementation. This allows implementation of the technology in a cost efficient step wise manner. When air is utilized a low-Btu gas with a heating value of approximately 135 Btu/scf is produced. When oxygen is used, a medium-Btu gas with a heating value of approximately 290 Btu/scf is produced. The gas compositions, based on the Utah coal analysis shown in Figure 1, are shown in Figure 5.

GAS COMPOSITION

Component	Low Btu Volume %	Medium Btu Volume %
Hydrogen	13.7	37.4
Carbon Monoxide	22.4	45.8
Carbon Dioxide	7.2	12.9
Methane	2.0	2.7
Nitrogen	54.6	1.2
Hydrogen Sulfide	0.1	-
	100.0	100.0
Heating Value	137 Btu/scf	296 Btu/scf

Figure 5

Other attractive features offered by the Winkler process are:

- A wide variety of coals can be handled by the system.
- It is not necessary to dry the coal feed if it is in a free flowing state.
- The system is tolerant of both high and variable coal ash content.
- The system has a high turn down ratio and can be "banked" for periods when gas production is not required.

Process Description

The Winkler process gasifies coal in a fluidized bed. Figure 6 shows a typical gasifier train consisting of a feed system, gasifier, waste heat recovery boiler, cyclone and venturi scrubber.

Run of mine coal is crushed to a 3/8" x 0" size and fed to the gasifier through lock hoppers and screw conveyors. The bed fluidization is maintained at the specified operating temperature by the injection of a mixture of steam and air or oxygen at several points within the bed. Air is utilized for production of low-Btu gas, while oxygen is used for production of medium-Btu gas. A portion of the steam is fed as a pure component into the bottom of the bed to cool the larger ash particles that fall to the bottom of the bed. These particles are discharged from the bottom of the gasifier through lock hoppers and screw conveyors.

The operating temperature and pressure of the gasifier are extremely important to the process economics. Operation at pressures above atmospheric allows the production of greater quantities of gas from a given gasifier. It also allows savings by eliminating the need for fuel gas compression, because the volume of feed air or oxygen is substantially less than the product gas.

The operating temperature of the bed is also very important in that it effects the carbon conversion efficiency of the process. The Winkler system is capable of operating, if desired, above the ash softening temperature of the coal. This is achieved by injecting a portion of the air or oxygen above the dense phase of the bed. The softened entrained ash is cooled and resolidified by the use of a radiant boiler above the bed, out of the line of gas flow. This cooling of the ash avoids refractory damage that can result from ash sintering. Operation at temperature above 1700°F eliminates the formation of tars and phenols that can create a disposal problem in the gas cleaning section.

After leaving the gasifier, the gas is cooled in a waste heat boiler. By recovering waste heat in the radiant and waste heat boilers it is possible to produce steam in excess of the process requirements. This excess steam can either be exported, or utilized to supply part of the horsepower necessary to run the plant by using turbine driven pumps or compressors. Additional char particles are collected at the base of the waste heat boiler.

Gas leaving the waste heat boiler is passed through a cyclone for the removal of additional char particulates. Char collected from the gasifier, waste heat boiler, and cyclone is collected and can be used as fuel in a solid fuel boiler. The use of the recovered char as a supplemental boiler fuel improves the thermal efficiency of the process. The remaining char particulate matter is removed in a venturi scrubber to a level of less than ten grains per thousand standard cubic feet. Additional particulate will be removed in the sulfur removal process. Because the process does not produce phenols and tars, the blowdown stream from the venturi scrubber presents minimal waste-water treatment problems.

Essentially all the sulfur in the coal feed appears in the product gas as either H_2S or COS. Approximately 85 percent is H_2S. Following particulate removal, the gas can be treated for sulfur removal. The removal process consists of a COS hydrolysis step to H_2S, followed by an H_2S absorption step. The absorbed H_2S is then stripped from the absorption solution and converted to elemental sulfur.

The final step in gas production is to remove a portion of the moisture content to prevent condensation in distribution lines. This is done by refrigerating a portion of the gas, removing the condensed moisture, and then reheating the total gas stream above its dew point.

3.2 Retrofitting Steam Generators

The major modification required to retrofit existing steam generators to gas firing would be the installation of a fuel supply train and proportioning system designed for gaseous fuels. The existing oil burner could be replaced by a dual fuel burner (oil or gas). Leaving the existing oil supply and proportioning system intact would give the capability of returning to oil firing if so desired. The design of this burner would differ from that used for natural gas, due to the increased gas volume required for the same Btu input, along with the specific combustion characteristics associated with these lower Btu gases. Although considerable work has been done on combustion of lower Btu gaseous fuels, it is anticipated that work would be required to develop a burner which would provide a flame geometry compatible with the combustion chamber of a steam generator.

One factor associated with the use of either medium-Btu or low-Btu is steam generator derating. It is estimated that approximately a five percent derating would result when using medium-Btu gas, while derating would be in the neighborhood of ten percent if low-Btu gas were utilized.

Estimates of emissions of SO_2 resulting from the combustion of either medium-Btu or low-Btu gas indicate that both gases would be in compliance with the CARB standard for existing steam generators, as well as the standard for new steam generators. The use of the proposed medium-Btu gas, with a H_2S content of 100 ppm, would generate approximately 30 ppm of SO_2 in the flue gas of a 50 million Btu per hour steam generator when fired at rated capacity. This is well below the CARB limit of 60 ppm for new steam generators.

Estimates of NO_x emissions indicate that both low-Btu gas and medium-Btu gas would meet the CARB standard for existing steam generators. Low-Btu gas would also meet the standard for new steam generators, however medium-Btu gas may not achieve this standard. The difference results from a higher flame temperature produced by medium-Btu gas. The adiabatic flame temperature of medium-Btu gas is approximately 3550°F, compared to approximately 2935°F for low-Btu gas. At this higher temperature, oxidation of a greater portion of the nitrogen, available in the combustion air, could result in higher NO_x emissions. In view of this it may be necessary to use low NO_x burner technology to reduce formation of NO_x. Laboratory test work would be necessary to resolve this issue.

Based on the above discussion it is believed that the need for flue gas cleaning devices to meet the CARB standards for sulfur oxide and nitrogen oxide, would not be required. Also, particulate control would not be required.

3.3 Implementation

The plan for implementing Coal Gasification in the synthesized field calls for one gasifier complex to service the field. This complex will be located on the perimeter of the field, with gas being distributed throughout the field by pipeline. In order to estimate the cost of the gas distribution system, a hypothetical piping system was established. The major design criteria for this system was to provide gas at 10 psig at each group of steam generators. The maximum distribution distance would be four miles.

The selection of low-Btu gas or medium-Btu gas for the implementation portion of the project was based on economic considerations. Work done by Davy Powergas Inc. indicates that the economics are favorable for medium-Btu gas over low-Btu gas at the scale required for our synthesized field. Production of equivalent Btu capacity of low-Btu gas would require more gasifiers and material handling equipment. The cost of this equipment would be greater than the cost of oxygen plants needed to produce medium Btu gas while using fewer gasifiers. At this scale the equipment savings, along with improvement in thermal efficiency associated with the production of medium-Btu gas, justify the cost of oxygen production. It is estimated however that these economics would change in favor of low-Btu gas at a requirement of approxiamately 30 percent of that needed for our synthesized field.

For purposes of this project, we have elected to evaluate the use of medium-Btu gas because of its indicated economic advantage. The gasifier complex designed for this project calls for the production 17.8 million SCFH of 296 Btu/scf gas. The gas composition, based on the Utah coal selected for the project, is shown in Figure 5. The production of this gas requires approximately 6800 tons of coal per day, 4500 tons of oxygen per day and 305,000 pounds of steam per hour. An additional 540 tons of coal per day is required to produce process steam for the plant. The use of coal to generate additional steam to supply operating horsepower, allows the use of the recovered char as a supplemental fuel and improves the thermal efficiency of the project.

The material flow diagram and equipment requirements for the gasifier complex are shown in Drawing 4598-A-081478-3. Drawing 4598-A-081478-4 shows plot plans and elevations for the gasifier complex. Facilities required for the implementation plan are:

- Coal unloading, storage and handling facilities
- Six Winkler gasifier trains
- Three 1500 T/D oxygen plants
- Two 250,000 lb/hr solid fuel boilers
- Three gas treatment plants (COS hydrolysis, H_2S absorption, drying)
- One Claus Plant
- One raw water treatment plant

- One boiler feed water treatment plant
- One waste water treatment plant

Coal, delivered to the gasifier site by rail, is discharged to below grade track receiving hoppers. It is discharged from these hoppers by vibrating feeders and transported by belt conveyors to the coal storage yard.

Coal is reclaimed from live storage by two underpile hoppers equipped with vibrating feeders. Reclaimed coal is delivered by belt conveyor to a vibrating screen for sizing. Undersize coal, -3/8", passes through the screen, while oversize coal, +3/8", passes over the screen and is subsequently crushed to size by a roll crusher. Undersize coal, from the vibrating screen, and crushed coal, from the roll crusher, are then discharged on to a common belt conveyor. The sized coal is then conveyed to either the gasifier day bins or to the boiler coal bunker. Coal delivered to the gasifier day bins is withdrawn from these bins by belt feeders and delivered to the individual gasifier trains by six feed conveyors.

The gasifier section of the plant consists of six parallel trains. In this design the gasifiers are assumed to operate at 70 psia and a maximum temperature of 2100°F. Operating the gasifiers at 70 psia, versus atmospheric pressure, provides a significant increase in the quantity of gas that can be produced in each gasifier and eliminates the need for fuel gas compression for distribution. The 2100°F operating temperature provides an overall carbon conversion efficiency of 95%. Each train consists of a gasifier, waste heat boiler, cyclone and venturi scrubber as shown in Figure 6. The thermal efficiency of the entire plant is estimated to be 75%.

Oxygen for the gasification reaction is provided by three 1500 ton per day plants. Oxygen from these plants is supplied to the gasifier at 90 psia. By-product nitrogen is used to seal the lock hoppers and to convey and cool the hot char recovered for use as a supplemental boiler fuel. The primary source of power for the oxygen plant is supplied by high pressure steam produced from the gasifier, and two coal/char fired boilers. Less than 5% of the total plant horsepower is supplied by electricity.

As indicated, steam is produced both from the gasifiers and two 250,000 lb/hr solid fuel boilers. The boilers are necessary for both power and start-up steam. Steam is produced at 1050 psig and superheated to 900°F. Low pressure steam is supplied from back pressure turbines. A portion of the steam used to supply the large power requirement of the oxygen plant utilizes surface condensers.

As indicated, the two solid fuel boilers have the capability of burning char as well as coal. Approximately 5% of the carbon content of the coal is converted to char. Ninety percent of this is recovered for use as fuel. All char collected ahead of the venturi scrubbers is collected for use as fuel. Potential SO_2 emissions from the coal fired boilers are reduced by more than 90% by use of the Wellman-Lord flue gas desulfurization process. This process uses a buffered sodium sulfite/bisulfite system to absorb SO_2 from the flue gas. The absorption solution is regenerated in an evaporator crystallizer which produces a concentrated stream of SO_2. This SO_2 can be converted into either sulfuric acid or elemental sulfur. For this project, it has been assumed that the SO_2 is converted to elemental sulfur in a Claus plant, utilizing H_2S recovered from the product gas. The Claus plant tail gas is also treated with the Wellman-Lord process, with the recovered SO_2 being recycled to the Claus plant. Approximately 40 tons per day of elemental sulfur are recovered by the Claus plant for sale.

After particulate removal in the venturi scrubbers, the product gas is sent to a gas treatment plant where sulfur compounds are removed. Essentially all the sulfur in the coal is converted to H_2S and COS. Approximately 85% of the sulfur is in the form of H_2S and 15% as COS. The first step is to convert COS to H_2S in a hydrolysis unit. This is followed by an Alkazid H_2S absorption unit. This system selectively absorbs H_2S so that the product contains less than 100 ppm of H_2S. The absorbed H_2S is stripped and sent to the Claus plant for conversion to elemental sulfur.

The final gas treatment step is to reduce the moisture content of the gas to avoid condensation in the distribution system. This is done by refrigerating the gas, removing the condensed water vapor, and by reheating the gas. Steam is used to power the refrigeration compressor and to reheat the gas. The treated dry gas is then discharged into the gas distribution system at approximately 50 psia.

3.4 Economics

This section contains the results of an economic analysis of implementing coal gasification in the synthesized field. This field would generate approximately 3.504 x 10^7 thousand pounds of steam per year.

The Economic Analysis, as explained in a preceding section, is based upon capital costs (including pollution control equipment) which are over and above those of the existing facilities. Similarly, the operating costs used are not the total costs for the production of steam but rather the cost to cover only those items related to the new facilities. This includes the total fuel cost, however.

3.4.1 Conclusions

Using a delivered cost of $40.00 per ton of coal and a credit of $30.00 per ton of sulfur, an added price of $6.00 per thousand pounds of steam is necessary in order for the coal gasification operation to obtain an after-tax IRR of 15 percent. Assuming 23,360,000 barrels of oil annually available for sale from the enhanced recovery program, the above price equates to $9.00 added to the cost of producing one barrel of oil for sale. This price is predicated upon a plant investment of $261.4 million and a working capital requirements of $29.4 million.

The sensitivity of the IRR to a change in the value of a specific input variable is measured by the index of sensitivity. For the variables listed below, the indices of sensitivity are as follows:

Variable	Index of Sensitivity (Percentage Point Change in IRR Per 1 Percent Change in Value of Input Variable
Incremental Price	.21
Incremental Investment	.09
Incremental Variable Costs	.11

The internal rate of return is, therefore, most sensitive to the incremental price of steam. Its index value indicates that a one percent change in steam price effects a change of .21 percentage points in the IRR.

3.4.2 Input Variables

The following paragraphs detail the values that are used as inputs to the computer program for the Coal Gasification analysis.

The gasification plant is assumed to attain its expected production level approximately two years after start-up. Costs and other data relating to operation at rated capacity are as follows:

3.4.2a Production Volume and Operating Schedule

Production costs are based on an annual steam production of 3.504×10^7 thousand pounds of steam. The plant operates 328 days per year.

3.4.2b Added Operating Costs

The estimated annual additional operating costs associated with the coal gasification plant are based on the labor and raw material costs detailed in the Basis of Evaluation discussion. These estimates are considered to be attainable at full production.

Added Operating Costs

Variable Costs	($1000)	Cost ($ per 1000 lbs. Steam)
Raw Materials (Not Incremental)	95,749	2.73
Coal $96,160		
Sulfur (credit) (411)		
Other Operating Costs (Variable)		
(Includes repair and maintenance, utilities, supplies and consummables, ash handling (sludge) and NO_x, SO_x and particulate control)	12,951	.37
Total Variable Costs	108,700	3.10
Semi-Variable Costs		
Labor	4,900	.14
Semi-Variable Expenses		
Ad-valorem taxes and insurance	10,456	.30
TOTAL ADDED OPERATING COSTS	124,056	3.54

3.4.2c Capital Cost Estimates

The installed cost of the coal gasification facilities and related pollution abatement equipment for the existing steam generators is estimated to be \$261.4 million in 1978 dollars. This estimate is accurate to within ±25 percent and is itemized as follows:

	-4	-3	-2	-1	Start-Up 0	Total
	(Millions of Dollars)					
Buildings	.86	2.15	3.87	1.72	-	8.6
Distrib. Pipeline	.50	1.25	2.25	1.00	-	5.0
Plant Equipment	24.78	61.95	111.51	49.56	-	247.8
	26.14	65.35	117.63	52.28		261.4

3.4.2d Incremental Working Capital

Working capital is provided to cover current cash requirements, accounts receivable, inventory of operating materials, as well as accounts payable.

Cash is sufficient for a four-month payroll, accounts receivable are 18 percent of sales (60 day payment), accounts payable are 18 percent of operating costs excluding labor (60 day payment), and inventory volumes are sufficient for annual needs.

Incremental Total Working Capital = 14 Percent of Incremental Sales.

3.4.3 Internal Rate of Return

The added price of steam generated by using gas to fire steam generators is \$6.00 per thousand pounds of steam in order to obtain the assumed 15 percent after-tax IRR.

The individual components of that price are as follows:

Added Operating Costs and Fixed Charges

Item	($1000)	Cost ($ per 1000 lbs. Steam)
Raw Materials (Not Incremental)	95,749	2.73
Other Operating Costs (Variable)	12,951	.37
Labor (Semi-Variable)	4,900	.14
Ad-valorem Taxes and Insurance	10,456	.30
Total Incremental Operating Costs	124,056	3.54
Capital Charges	86,184	2.46
TOTAL COST	210,240	6.00

Annual capital charges noted above are required to obtain an IRR of 15 percent after tax on the added investment. These total approximately 33 percent of the total capital cost discussed earlier in this section.

3.4.4 Sensitivity Analysis

The following is a summary of variables including incremental steam price, investments, and costs as they affect the sensitivity of the internal rate of return for the proposed project.

3.4.4a Incremental Price

An incremental price of $6.00 per thousand pounds of steam was calculated to be the price "most likely" to yield an after-tax IRR of 15 percent.

To gauge sensitivity, this price was varied from a low of $4.50 per thousand pounds of steam to a high of $7.50 per thousand pounds of steam. The corresponding values of the IRR were 6.7 and 20.5 percent respectively.

Incremental Price

$/1000 lbs. Steam	IRR, %
4.50	6.7
6.00	15.0
7.50	20.5

3.4.4b Incremental Investment

The incremental plant and equipment investment for the proposed coal gasification alternative has a "most likely" total of $261.4 million spent over an approximate four-year period. This total was varied by 25 percent on the high and low side to yield investment totals from $196.1 million to $326.8 million. The resulting IRR values were 18.4 and 12.5 percent respectively.

Incremental Investment

Plant Capital Cost	IRR, %
$196,100,000	18.4
261,400,000	15.0
326,800,000	12.5

3.4.4c Incremental Variable Cost

Variable cost includes all items noted in the discussion of operating costs, with the exception of the semi-variable components. A 25 percent factor was applied for the high and low incremental variable cost values, resulting in a range of $2.33 to $3.88 per thousand pounds of steam. The corresponding IRR values were 18.1 and 11.0 respectively.

Incremental Variable Cost

Cost per 1000 lbs. Steam	IRR, %
$2.33	18.1
$3.10	15.0
$3.88	11.0

As can be seen from the table of Added Operating Costs and Fixed Charges, the major portion of the variable cost is attributed to the cost of raw materials and in particular coal. Figure 7 presents the sensitivity of steam cost as a function of the delivered price of coal. Varying the price of coal by 25 percent above and below the value of $1.72 per million Btu used in the study yields costs per thousand pound of steam of $6.70 and $5.22 respectively.

Coal Cost

Delivered Price ($ per Million Btu)	Steam Cost ($ per 1000 lbs)
1.29	5.22
1.72	6.00
2.15	6.70

COAL GASIFICATION
SENSITIVITY TO COAL COST

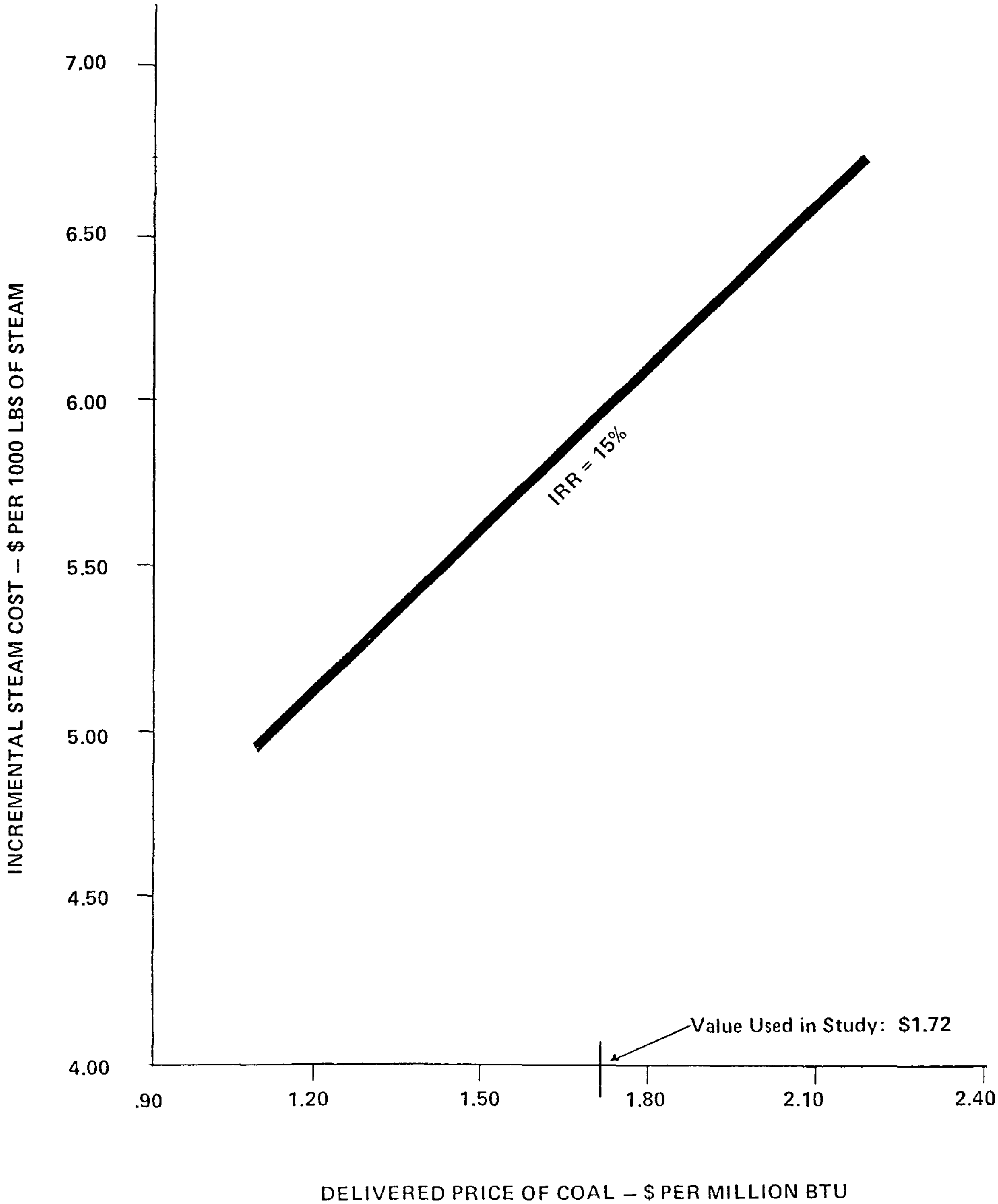

FIGURE 7

3.4.4d <u>Index of Sensitivity</u>

The sensitivity analysis can be summarized by noting that an index of sensitivity can be calculated for each of the variables discussed earlier. The larger this index number, the more sensitive the IRR is to a percent change in that variable.

Mathematically, the index of sensitivity can be expressed as:

$$I = \frac{IRR}{\left(\frac{V_2 - V_1}{V_1}\right) \times 100}$$

IRR = Absolute change in IRR obtained by varying a specific input variable (e.g., selling price)

V_1 = Initial input variable value

V_2 = Final input variable value

The value "I" gives the change in IRR units per percentage change in the specific variable.

Using this equation results in the following indices of sensitivity:

Variable	Index of Sensitivity, I
Incremental Price	.21
Incremental Investments	.09
Incremental Variable Costs	.11

The Internal Rate of Return is, therefore, most sensitive to the incremental price of steam. A one percent change in that price will result in a change of 0.21 in the IRR.

3.5 Demonstration

The primary objectives of a Coal Gasification demonstration program would be to evaluate the operational and environmental impact of low-Btu or medium-Btu gas on an oil-fired steam generator. Demonstration by construction and operation of a gasifier plant in the field is not required. Considerable work has already been done to demonstrate gasification technology. The Winkler process, along with several other processes, are proven technologies that are commercially available. The cost of demonstration in the field, even at ten percent the scale of the synthesized field, would be prohibitive. In addition, almost 100 percent of this cost would be directed at demonstrating proven technology.

The objectives of the demonstration program could be achieved in a more cost effective manner through subscale combustion testing. Synthesized gas could be used for this purpose. This testing would determine flame characteristics and emission levels. It would also determine if any burner development work would be required. Test results could then be compared with oil firing to determine the impact on the steam generator. It is anticipated that this work could be accomplished for under $500,000. Commercially available gasifier technology could then be used for implementation, if so desired.

4.0 FLUIDIZED BED COMBUSTION

4.1 Technical Issues

After examining various direct coal fired technologies, the one most suited to enhanced oil recovery steam generation, and therefore selected for evaluation, was Fluidized Bed Combustion (FBC). This direct fired technology offers several technical advantages, over stoker or pulverized coal firing methods, which lend themselves to this application. The primary advantage is its capablility to control sulfur oxide emissions by addition of a limestone sorbent into the bed of the combustor. This feature eliminates the need for expensive scrubbing equipment.

In coal combustion, sulfur in the coal is oxidized to gaseous sulfur dioxide. With FBC, coal is burned in a bed of limestone. The limestone is calcined to produced calcium oxide, as shown by Equation 1.

$$(1)\quad CaCO_3 + \text{Heat} \longrightarrow CaO + CO_2$$

The calcium oxide then reacts with the sulfur dioxide, as shown by Equation 2, to produce calcium sulfate.

$$(2)\quad CaO + 1/2\ O_2 + SO_2 \longrightarrow CaSO_4 + \text{Heat}$$

The calcium sulfate is then drained from the bed along with the coal ash. It has been demonstrated that with calcium to sulfur mole ratios of 3:1 or more, sulfur retention in the neighborhood of 90 percent can be achieved.

Waste disposal is also less of a problem than with other conventional coal firing methods. With bed temperatures of 1500°F to 1600°F, well below the ash fusion temperature of coal, the spent waste comes out as an ash-like sand with no clinkers. Studies currently underway to investigate the possibility of using this spent waste as a soil supplement, aggregate for concrete block, and many other applications, are very encouraging as to the potential use of this by-product material.

In addition to limiting sulfur oxide emissions, FBC offers several other benefits. A coal fired FBC unit would be smaller in size than a conventional coal fired steam generator. This is largely due to the ability of the FBC unit to work at higher heat release rates, approximately 100,000 Btu/hr-ft^3, compared to those of approximately 20,000 Btu/hr-ft^3 for conventional units. In addition high heat transfer coefficients to in-bed tubes, where approximately sixty-five percent of heat absorption takes place, have been experienced. Because of these advantages, it should be possible to design a forced circulation, once through steam generator which would use the same untreated water that is presently being used in existing steam generators to produce low quality steam.

Another benefit associated with FBC is reduction in the emission of oxides of nitrogen. The relatively low combustion temperature of an FBC unit, compared to conventional coal fired units, limits the production of oxides of nitrogen. Existing test data indicate that uncontrolled emissions of NO_x from FBC units are below existing federal regulations. This level however, as will be discussed later in the report, exceeds the limitation for new steam generators of 100 ppm as proposed in the California Air Resources Board model rule. Therefore, equipment to reduce NO_x emissions to an acceptable level will have to be included for this application. This still provides an advantage over other conventional coal firing techniques however, in that the amount of reduction required will be less.

4.1.1 Process Description

Combustion in the FBC process is accomplished within a bed of noncombustible, granular material. Coal is burned in a limestone bed above a grid plate, through which an evenly distributed air stream is passed. This air serves two functions in that it provides the air required for combustion as well as that needed for fluidization. As the air velocity is increased, the bed will expand, or fluidize, creating a fluid-like mass which is held in suspension by the air stream. Typical fluidizing velocities are in the range of 3 to 12 fps. This fluid-like turbulent bed causes rapid mixing of the particles to occur, which is ideal for good combustion. The heat produced by this combustion is then transferred to water tubes in the bed for steam generation.

Figure 8 shows a typical FBC flow diagram which may be used for enhanced oil recovery steam generation. This design shows sized coal and limestone, along with recycled ash, being blended prior to injection into the combustor. Typical feed size for coal is approximately 1/4" x 0, while the size requirement for limestone is approximately 1/8" x 30 mesh. Although this approach shows the fuel/sorbent mixture being blended prior to entry to the combustor, these materials can be fed independently.

Two approaches for feeding material into the combustor are available. Mechanical methods, such as screw conveyors or spreader stoker feeders can be used, or feeding can be accomplished pneumatically. Mechanical methods are used for overbed feeding, while feeding from below the bed or in the bed is generally accomplished pneumatically.

Fuel and limestone injected into the bed are fluidized by air passing through a distributor plate in the combustor. The distributor plate serves two functions in that it causes even distribution of air across the bed, and also supports the bed when in a non-fluidized, slumped condition.

The fluidizing air is supplied by a forced draft fan, through a plenum located below the distributor plate. This fan, in conjunction with an induced draft fan located after the combustor, operates as a balanced draft system to maintain slightly greater than atmospheric pressure in the bed.

Temperature in the bed of the combustor is approximately 1500°F to 1600°F. Sulfur dioxide formed during combustion is absorbed by the limestone in the bed. Sulfur capture will limit SO_2 emissions to a level below the limitation of 60 ppm proposed in the CARB model rule and therefore, flue gas scrubbing for SO_2 reduction is not required.

Spent bed materials are removed from the combustor by a gravity drain system. These materials are fed into an ash cooler, where their temperature is reduced to a level suitable for handling. After cooling, this material is discharged to an ash silo for subsequent disposal.

Flue gas leaving the combustor passes through a mechanical cyclone collector where approximately 85 to 90 percent of the suspended flyash and unburned carbon are removed. The material collected by the cyclone is then reinjected into the combustor. This recycling increases combustion efficiency.

Cyclone discharge gases then pass through an economizer section, where temperature is reduced to approximately 350°F. After leaving the economizer, the gases pass through a baghouse dust collector, where final particulate removal is accomplished.

Also shown in Figure 8 is provision for NO_x removal. Test data from laboratory scale test units indicate NO_x concentrations in the flue gas range between 300 and 500 ppm. Although this is below federal standards, it is higher than the limitation of 100 ppm proposed in the CARB model rule. After evaluating data from several test reports, an uncontrolled NO_x emission of 450 ppm was chosen in order to estimate the cost of NO_x emission control. Based on this figure, a reduction of approximately 78 percent would be required to meet the CARB limitation.

At the present time two processes, non-catalytic and catalytic, are available for controlling NO_x emissions. Both processes use ammonia injection and require a narrow and critical temperature range for optimum effectiveness. The non-catalytic process reacts ammonia with NO_x to form nitrogen and water. This process operates in a temperature range of approximately 1700°F to 1800°F and is capable of NO_x reductions of up to 70 percent. The catalytic process also reacts ammonia with NO_x to form nitrogen and water. This process, however, operates in a lower temperature range of 600°F to 800°F and is capable of NO_x reductions in excess of 90 percent.

Typical flue gas temperatures in an FBC unit are well below 1700°F. Flue gas reheating would be required in order to achieve the temperature range required for the non-catalytic process. This temperature range can be shifted to a lower level, however, but requires the use of hydrogen. The magnitude of this shift is mainly a function of the amount of hydrogen injected relative to ammonia. It should be noted, however, that this hydrogen addition does not widen the temperature range, but merely shifts this temperature window to a lower level.

Comparison of the two technologies would have to be made on a case by case basis. In this application, the cost of reheating the flue gas or the cost of hydrogen would have to be evaluated versus catalyst cost. In addition, the level of reduction required versus the capabilities of the two processes must be considered. Based on estimated uncontrolled NO_x emissions of 450 ppm, a reduction of approximately 80 percent is required. Since this level is beyond the capability of the non-catalytic process, we have elected to evaluate a catalytic process in order to estimate the cost associated with NO_x control. Should the reduction requirement fall within the capabilities of the non-catalytic process, then the cost of implementing that process versus the catalytic approach could become comparable.

4.1.2 NO_x Emission Control

Catalytic processes have been developed by a number of vendors. Among these are the Hitachi Ltd. process, the Kobe Steel process, the J.G.G. Paranox process, the Kurabo Knorca process and the Hitachi Zosen process. The characteristics and performances of these various processes are very similar. The technology selected for evaluation in the project is the Hitachi Zosen process offered by Envirotech/Chemic. This process is commercially available in the United States and has recently been selected for demonstration on a coal fired boiler by the Environmental Protection Agency. Cost data utilized was supplied by the vendor and in the accuracy presented, is thought to be representative of the other catalytic processes.

The Hitachi Zosen process is a dry catalytic type process and is selective in that only one chemical species, in this case oxides of nitrogen, is reduced. The process reacts ammonia with oxides of nitrogen, in near stoichiometric quantities, to form nitrogen and water. The process operates at temperatures between 600° and 800°F and is capable of reductions in excess of 90 percent. The NO_x decomposition reactions are shown by Equations 3 and 4:

3) $6\ NO + 4\ NH_3 \longrightarrow 5\ N_2 + 6\ H_2O$

4) $6\ NO_2 + 8\ NH_3 \longrightarrow 7\ N_2 + 12\ H_2O$

The catalyst used in the process is an unactivated metal, treated with aluminum and in turn an aqueous solution of acid or alkali. This produces a metal surface which is catalytically active in the reducing reaction between nitrogen oxides and ammonia.

The production of the catalyst used in the process involves plating an unactivated metal, is this case stainless steel, with aluminum.

The plating treatment of the aluminum layer on the surface of the stainless steel plate permits the stainless steel and aluminum to diffuse into each other in solid phase. The aluminum plated stainless steel panel is then immersed in an aqueous solution of sodium hydroxide where the aluminum is then selectively dissolved out into solution. The remaining stainless steel assumes porous surfaces and becomes catalytically active. The stainless steel when activated by this process then exhibits catalytic activity in the reaction for selectively reducing NO_x with ammonia. The active stainless steel plates are then formed into a honeycomb configuration for modular installation into a reactor.

The Hitachi Zosen process requires the installation of a catalytic reactor into the flue gas duct system. Figure 9 shows a simplified flow diagram of the process, as applied to a conventional boiler, not a steam generator, operation.

In this application, the catalytic reactor is located at a point where optimum process temperature conditions exist. Ammonia is injected into the flue gas stream ahead of the reactor. This is accomplished by means of a carrier gas, which can be either steam or air. Soot blowers are provided with the reactor to prevent blocking or adherence of dust on to the catalyst layer. Provision is also shown for flue gas reheating. This would become necessary in the event flue gas temperature would fall below 700° to 800°F range.

In as much as NO_x control would be required in order to meet the CARB limitation of 100 ppm, it is believed that hardware for this process could be incorporated into the design of the FBC unit. The catalytic reactor would be installed ahead of the economizer section. The inbed heat transfer surfaces would then have to be designed such that the resulting flue gas temperature leaving the bed and entering the catalytic reactor would fall into the 700°F to

HITACHI ZOSEN DRY CATALYTIC DE NOX PROCESS

SIMPLIFIED FLOW DIAGRAM

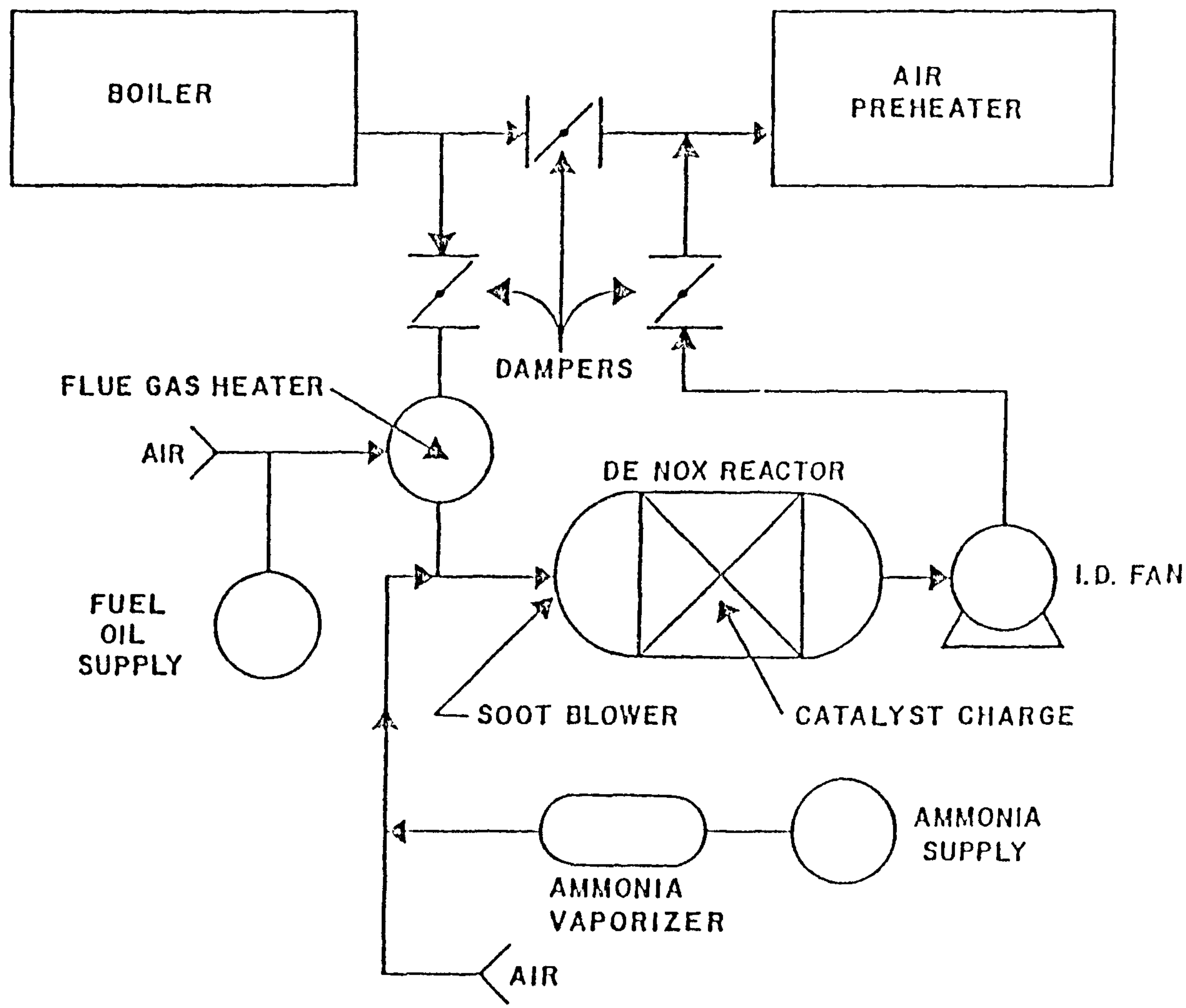

FIGURE 9

800°F range. This gas would then pass through the reactor and into the economizer section. Since no temperature drop takes place across the reactor, the performance of the economizer should not be effected. A pressure drop does occur however, and this would have to be compensated for by additional fan horsepower.

4.2 Implementation

The plan for implementing Fluidized Bed Combustion in the synthesized field calls for ten groups of FBC steam generators, with each group containing ten steam generators. Each unit will be rated at 50 million Btu per hour. Selection of this capacity was based on a unit size that could be shop assembled for shipment to the field as a package. This would minimize field erection and in turn lower overall cost.

Drawing 4598A-081478-5 shows the material flow diagram and equipment requirements for one group of ten FBC units. Drawing 4598A-081478-6 shows plot plans and elevations for the same ten units. This plan would be duplicated ten times in order to satisfy the requirement of the synthesized field.

The operation of each of the groups of ten FBC units requires approximately 730 tons of coal per day and 210 tons of limestone per day. Limestone analysis is shown in Figure 2. Total coal and limestone requirements are approximately 7300 and 2100 tons per day respectively.

Run of mine coal is delivered to the site by rail and is discharged to below grade track receiving hoppers. Coal is discharged from these hoppers by vibrating feeders and transported to the storage yard by belt conveyor. Sized limestone is delivered to the site by either rail or truck. Limestone is then pneumatically conveyed to the limestone storage silo.

Coal is reclaimed from storage by two under pile hoppers equipped with vibrating feeders. Reclaimed coal is delivered by belt conveyor to a vibrating screen for sizing. Undersize coal, -1/4", passes through the screen, while oversize coal, +1/4", passes over the screen and is subsequently crushed to size by a roll crusher. Undersize coal, from the vibrating screen, and crushed coal, from the roll crusher, are then discharged on to a common conveyor. This sized coal is then transported by belt conveyor to the coal storage silo. Sized coal is discharged from the coal storage silo by means of a vibrating bin bottom. It is then transported by belt conveyors to a diverter chute, where it is directed either to the coal storage bins of FBC units No. 1 through No. 5 or No. 6 through No. 10. Conveyor plows direct the coal to the individual FBC coal feed hoppers in a preselected sequence.

Limestone is discharged from the limestone storage silo by means of a screw conveyor. It is then pneumatically conveyed to the individual FBC limestone storage bins. This system is also automatically controlled in a preselected sequence.

Two baghouse dust collectors serve the ten FBC units. Flue gas from FBC Units 1 through 5 is manifolded into Baghouse #1. Flue gas for FBC Units 6 through 10 is manifolded to Baghouse #2.

Fly ash collected by the two baghouses, as well as bed drain ash from the individual FBC units, is collected and conveyed in a common pneumatic transport system. This system is automatically sequenced to accept ash on an individual basis, from each of the twelve collection points. The ash is then pneumatically conveyed to the ash storage silo. The discharge from the ash storage silo passes through a wetter muller, which moistens the ash to suppress dusting.

4.3 Economics

This section contains the results of an economic analysis of implementing Fluidized Bed Combustion in the synthesized field. The field would generate approximately 3.504×10^7 thousand pounds of steam per year.

The Economic Analysis, as explained in a preceding section, is based upon capital costs (including pollution control equipment) which are over and above those of the existing facilities. Similarly, the operating costs used are not the total costs for the production of steam but rather, the cost to cover only those items related to the new facilities. This includes the total fuel cost, however.

4.3.1 Conclusions

Using delivered costs of $40.00 per ton of coal and $15.00 per ton of limestone, an added price of $6.36 per thousand pounds of steam is necessary in order for the fluidized bed operation to obtain an after-tax IRR of 15 percent. Assuming 23,360,000 barrels of oil annually available for sale from the enhanced recovery program, the above price equates to $9.54 added to the cost of producing one barrel of oil for sale. This price is based upon a plant investment of $348 million and a working capital requirement of $28.9 million.

The sensitivity of the IRR to a change in the value of a specific input variable is measured by the index of sensitivity. For the variables listed below, the indices of sensitivity are as follows:

Variable	Index of Sensitivity (Percentage Point Change in IRR Per 1 Percent Change in Value of Input Variable
Incremental Price	.22
Incremental Investment	.10
Incremental Variable Costs	.10

The internal rate of return is, therefore, most sensitive to the incremental price of steam. Its index value indicates that a one percent change in steam price effects a change of 0.22 percentage point in the IRR.

4.3.2 Input Variables

The following paragraphs detail the values that are used as inputs to the computer program for the FBC analysis.

The 100 combustors are assumed to attain their total expected production level approximately two years after start-up. Costs and other data relating to operation at rated capacity are as follows:

4.3.2a Production Volume and Operating Schedule

Production costs are based on an annual steam production of 3.504×10^7 thousand pounds of steam.

4.3.2b Added Operating Costs

The estimated annual additional operating costs associated with the fluidized bed facilities and related pollution control equipment on the existing steam generators are based on the labor and raw material costs detailed in the Basis of Evaluation discussion. These estimates are considered to be attainable at full production.

Added Operating Costs

Variable Costs	($1000)	Cost ($ per 1000 lbs. Steam; Approx.)
Raw Materials (Not Incremental)	82,451	2.35
Coal $73,216		
Limestone 9,235		
Other Operating Costs (Variable)		
(Includes repair and maintenance, utilities, supplies and consumables, ash handling (sludge) and NO_x, SO_x and particulate control)	17,549	.50
Total Variable Costs	100,000	2.85
Semi-Variable Costs		
Labor	7,650	.22
Semi-Variable Expenses		
Ad-valorem Taxes and Insurance	13,920	.40
TOTAL ADDED OPERATING COSTS	121,570	3.47

4.3.2c Capital Cost Estimates

The installed cost of the fluidized bed facilities (including related pollution abatement equipment) is estimated to be $348

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million in 1978 dollars. This estimate is accurate to ±25 percent and is itemized as follows:

Year	-5	-4	-3	-2	-1	0	Total
			(Millions of Dollars)				
Buildings	.250	.750	2.00	1.50	.50	-	5.0
Pollution Control	1.415	4.245	11.32	8.49	2.83	-	28.3
FBC Plant Equipment	15.735	47.205	125.88	94.41	31.47	-	314.7
	17.400	52.200	139.20	104.40	34.80		348.0

4.3.2d Incremental Working Capital

Working capital is provided to cover current cash requirements, accounts receivable, inventory of operating materials, as well as accounts payable.

Cash is sufficient for a four-month payroll, accounts receivable are 17 percent of sales (60 day payment), accounts payable are 17 percent of operating costs excluding labor (60 day payment), and inventory volumes are sufficient for annual needs.

Incremental Total Working Capital = 13 Percent of Incremental Sales.

4.3.3 Internal Rate of Return

The added price of steam generated from the FBC steam generators is $6.36 per thousand pounds of steam in order to obtain the assumed 15 percent after-tax IRR.

The individual components of that price are as follows:

Added Operating Costs and Fixed Charges

Item	Cost ($1000)	Cost ($ per 1000 lbs. Steam)
Raw Materials (Not Incremental)	82,451	2.35
Other Operating Costs (Variable)	17,549	.50
Labor (Semi-Variable)	7,650	.22
Ad-valorem Taxes and Insurance	13,920	.40
Total Incremental Operating Costs	121,570	3.47
Capital Charges	101,284	2.89
TOTAL COST PER 1000 LBS. STEAM	228,854	6.36

Annual capital charges noted above are required to obtain an IRR of 15 percent after tax on the added investment. These total approximately 29 percent of the total capital cost discussed earlier in this section.

4.3.4 Sensitivity Analysis

The following is a summary of variables including incremental steam price, investments, and costs as they affect the sensitivity of the internal rate of return for the proposed project.

4.3.4a Incremental Price

An incremental price of $6.36 per thousand pounds of steam was calculated to be the price "most likely" to yield an after-tax IRR of 15 percent.

To gauge sensitivity, this price was varied from a low of $4.77 per thousand pounds of steam to a high of $7.95 per thousand pounds of steam. The corresponding values of the IRR were 6.9 and 21.3 percent respectively.

Incremental Price

$/1000 lbs. Steam	IRR, %
4.77	6.9
6.36	15.0
7.95	21.3

4.3.4b Incremental Investment

The incremental plant and equipment investment for the proposed fluidized bed alternative has a "most likely" total of $348 million spent over an approximate five-year period. This total was varied by 25 percent on the high and low side to yield investment totals from $261 million to $435 million. The resulting IRR values were 18.8 and 12.1 percent respectively.

Incremental Investment

Plant Capital Cost	IRR, %
$261,000,000	18.8
348,000,000	15.0
435,000,000	12.1

4.3.4c Incremental Variable Cost

Variable cost includes all items noted in the discussion of operating costs, with the exception of the semi-variable components. A 25 percent factor was applied for the high and low incremental variable cost values, resulting in a range of $2.14 to $3.56 per thousand pounds of steam. The corresponding IRR values were 18.0 and 11.4 respectively.

Incremental Variable Cost

Cost Per 1000 lbs. Steam	IRR, %
2.14	18.0
2.85	15.0
3.56	11.4

As can be seen from the table of Added Operating Costs and Fixed Charges, the major portion of the variable cost is attributed to the cost of raw materials and in particular coal. Figure 10 presents the sensitivity of steam cost as a function of the delivered price of coal. Varying the price of coal by 25 percent above and below the value of $1.72 per million Btu used in the study yields costs per thousand pound of steam of $6.88 and $5.83 respectively.

Coal Cost

Delivered Price ($ per Million Btu)	Steam Cost ($ per 1000 lbs)
1.29	5.83
1.72	6.36
2.15	6.88

4.3.4d Index of Sensitivity

The sensitivity analysis can be summarized by noting that an index of sensitivity can be calculated for each of the variables discussed earlier. The larger this index number, the more sensitive the IRR is to a percent change in that variable.

FLUIDIZED BED COMBUSTION
SENSITIVITY TO COAL COST

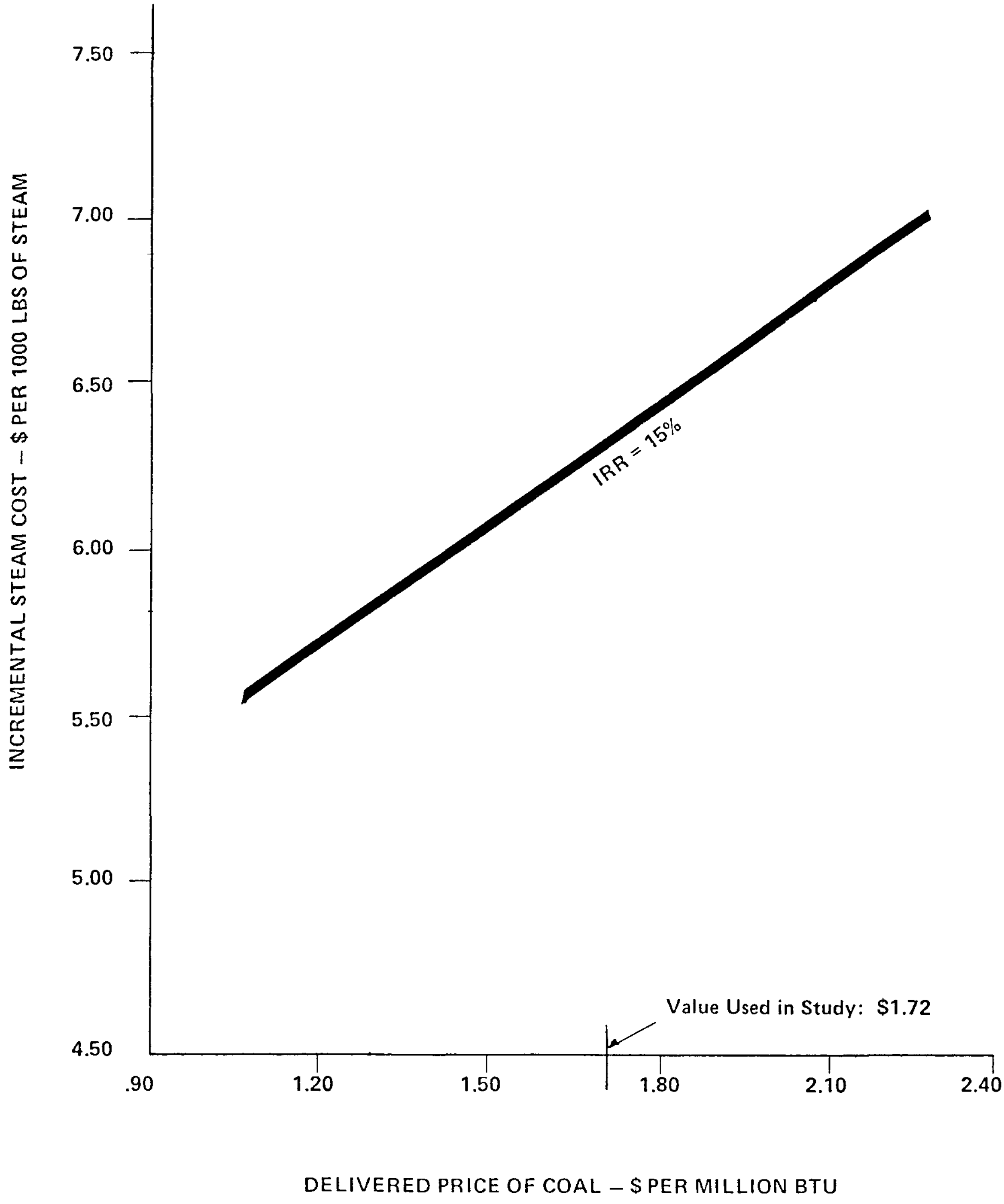


FIGURE 10

Mathematically, the index of sensitivity can be expressed as:

$$I = \frac{IRR}{\left(\frac{V_2 - V_1}{V_1}\right) \times 100}$$

IRR = Absolute change in IRR obtained by varying a specific input variable (e.g., selling price)

V_1 = Initial input variable value

V_2 = Final input variable value

The value "I" gives the change in IRR units per percentage change in the specific variable.

Using this equation results in the following indices of sensitivity:

Variable	Index of Sensitivity, I
Incremental Price	.22
Incremental Investments	.10
Incremental Variable Costs	.10

The Internal Rate of Return is, therefore, most sensitive to the incremental price of steam. A one percent change in that price will result in a change of 0.22 in the IRR.

4.4 Demonstration

Demonstration of Fluidized Bed Combustion could be accomplished through engineering, construction and operation of a single FBC steam generator designed specifically for oil field application. The following is an estimate of the cost of this program based on using a 50 million Btu per hour FBC steam generator for demonstration.

Capital costs include costs for items such as engineering, equipment, installation and construction. Equipment covered by this estimate includes coal, limestone and ash handling equipment, one Fluidized Bed Steam generator and emission control equipment for NO_x reduction and particulate collection. Direct operating costs were developed on the basis of one year's operation at 80% utilization. This cost covers items such as raw materials (coal, limestone and ammonia), labor, utilities, repair and maintenance and waste disposal.

Capital Cost	$5,500,000
Direct Operating Cost	1,735,000
TOTAL DEMONSTRATION COST	$7,235,000

Davy Powergas

WINKLER COAL GASIFIER PROCESS SCHEMATIC

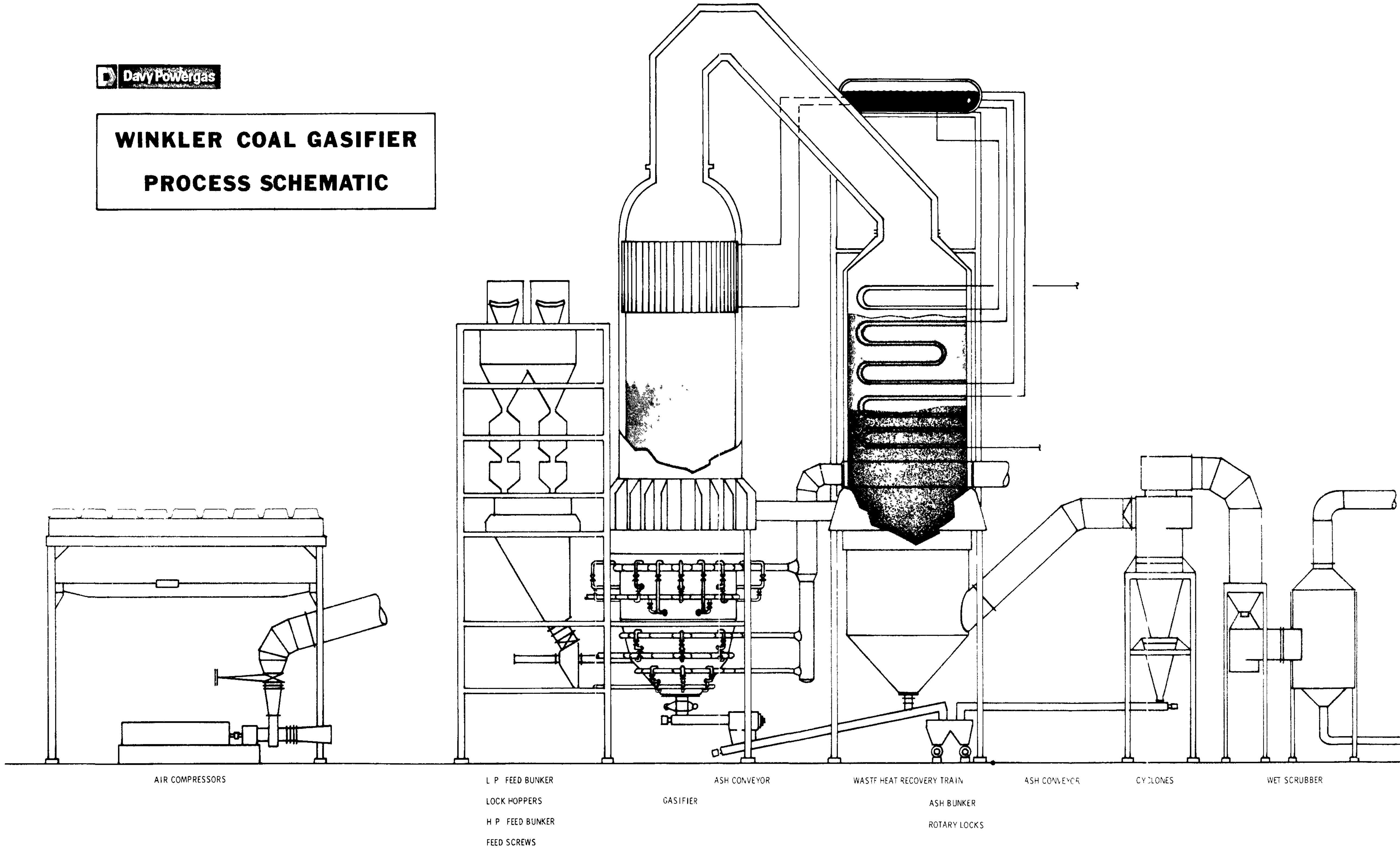


FIGURE 6

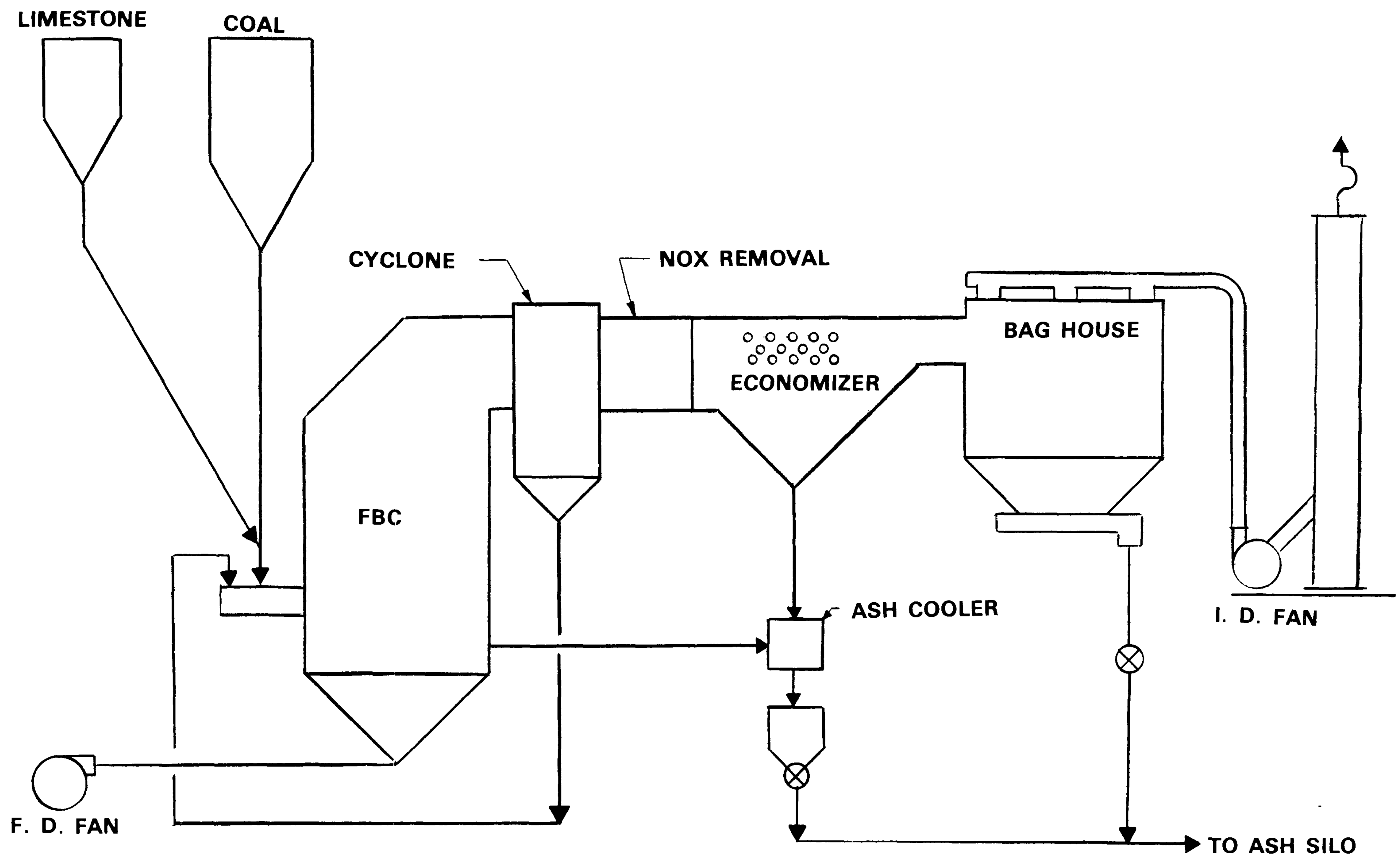
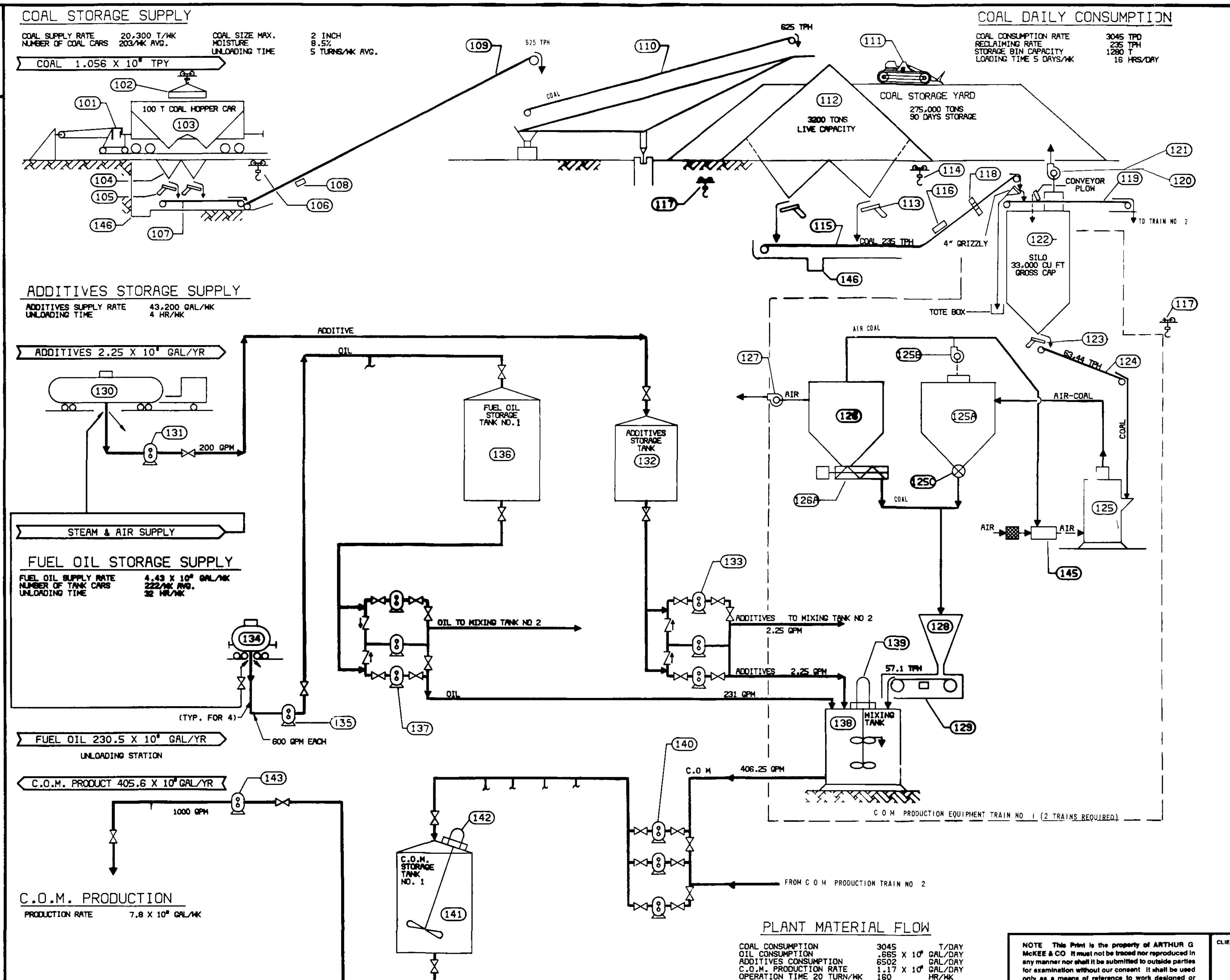

LIMESTONE
COAL
CYCLONE
NOX REMOVAL
BAG HOUSE
ECONOMIZER
FBC
ASH COOLER
I. D. FAN
F. D. FAN
TO ASH SILO

FIGURE 8



ITEM NO	QTY	DESCRIPTION
101	1	CAR POSIT ONER
102		CAR SHAKER
103		HOPPER OPEN TOP FREIGHT CARS
104	1	TRACK HOPPER
105	2	V BRATING FEEDER
106	1	HOIST
107	1	CONVEYOR
108	1	WEIGH SCALE
109	1	CONVEYOR
110	1	PIVOTAL STACKER
111	1	BULLDOZER
112	1	STORAGE YARD
113	2	VIBRATING FEEDER
114	1	HO ST
115	1	CONVEYOR
116	1	CROSSBELT MAGNET C PICKUP
117	2	HO ST
118	1	METAL DETECTOR
119	1	CONVEYOR
120	2	DUCT COLLECTOR
121	2	FAN
122	2	SILOS
123	2	VIBRATING FEEDERS
124	2	CONVEYOR
125	2	ROLLER M LL
125A	2	CYCLONE COLLECTORS
125B	2	FAN
125C	2	ROTARY FEEDER
126	2	DUST COLLECTORS
126A	2	FEED SCREW CONVEYOR
127	2	FAN
128	2	SURGE BIN
129	2	GRAVIMETRIC WEIGH FEEDER
130	15	TANK TRUCK TRAILER
131	1	ADDITIVES UNLOAD PUMP
132	1	ADDITIVES STORAGE TANK
133	3	ADDITIVES TRANSFER PUMPS
134		TANK CAR
135	4	O L PUMP
136	2	O L STORAGE TANK
137	3	OIL TRANSFER PUMP
138	2	MIXING TANK
139	2	AGITATOR
140	3	C O M PUMP
141	5	C O M STORAGE TANK
142	5	AG TATOR
143		C O M PUMP
144		
145	2	HEATER
146	2	SUMP PUMP

PLANT MATERIAL FLOW

COAL CONSUMPTION	3045	T/DAY
OIL CONSUMPTION	.665 X 10^6	GAL/DAY
ADDITIVES CONSUMPTION	6502	GAL/DAY
C.O.M. PRODUCTION RATE	1.17 X 10^6	GAL/DAY
OPERATION TIME 20 TURN/WK	160	HR/WK

CLIENT D. O. E.

McKee

FORM M-1542 9/77

TITLE 405 6 MILLION GAL YR COAL OIL MIXTURE PROCESS PLANT MATERIAL FLOW DIAGRAM

SCALE ______ E/R NO ______

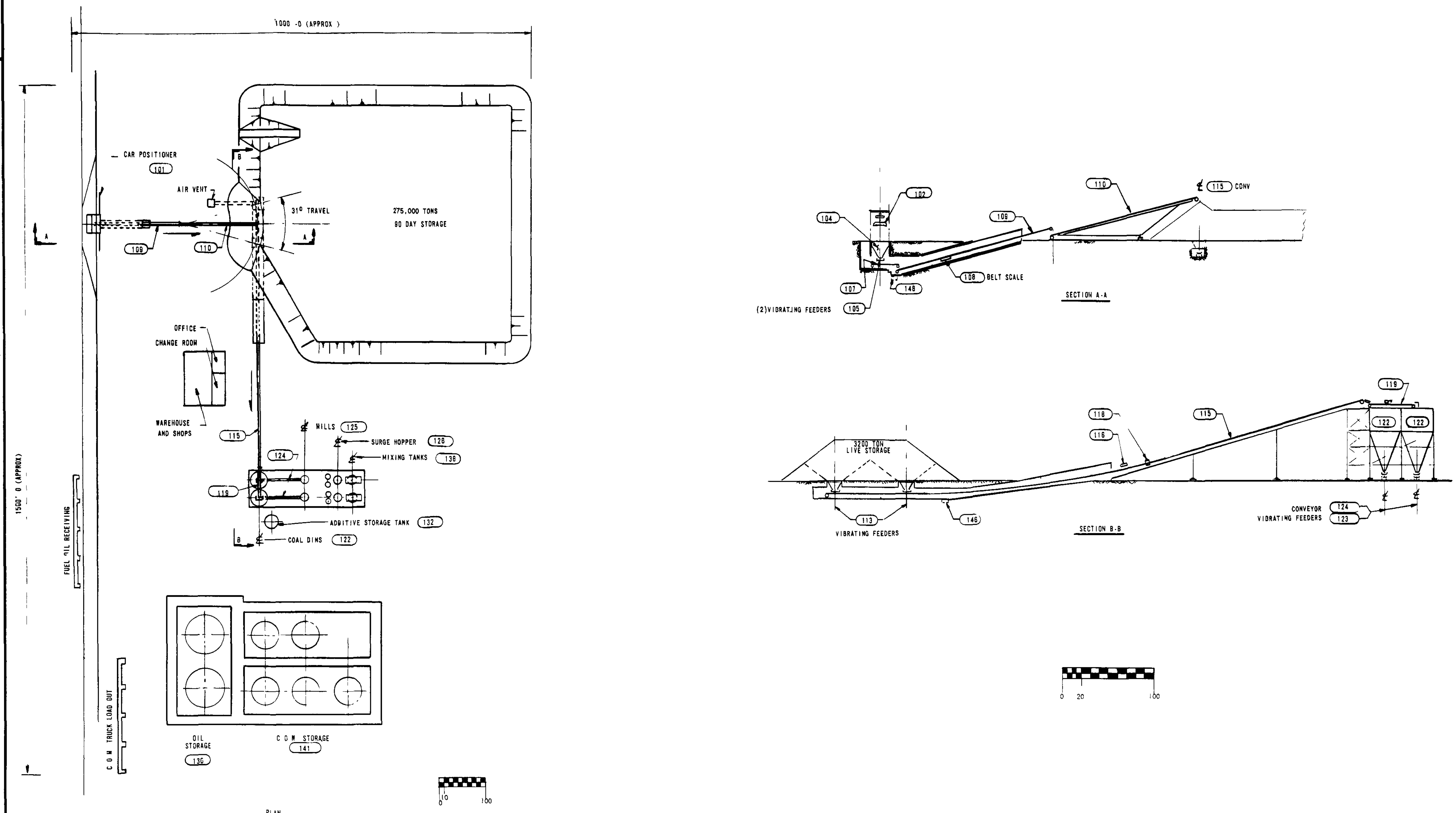
4598-A
081478-1

REVISION

NO	DESCRIPTION	BY	CH	APPROVED	DATE

DESIGNED	BY	DATE
DRAWN		
CHECKED		
APPROVED 1		
APPROVED 2		
APPROVED 3		

DATE TO: CLIENT, FIELD — A B C 0 1 2 3 4 5 6 7



NO	DESCRIPTION	BY	CH	APPROVED	DATE	NO	DESCRIPTION	BY	CH	APPROVED	DATE

REVISIONS | REVISIONS | REFERENCES

	BY	DATE
DESIGNED		
DRAWN		
CHECKED		
APPROVED 1		
APPROVED 2		
APPROVED 3		

DATE TO	A	B	C	0	1	2	3	4	5	6	7
CLIENT											
FIELD											

CLIENT

D. O. E.

McKee

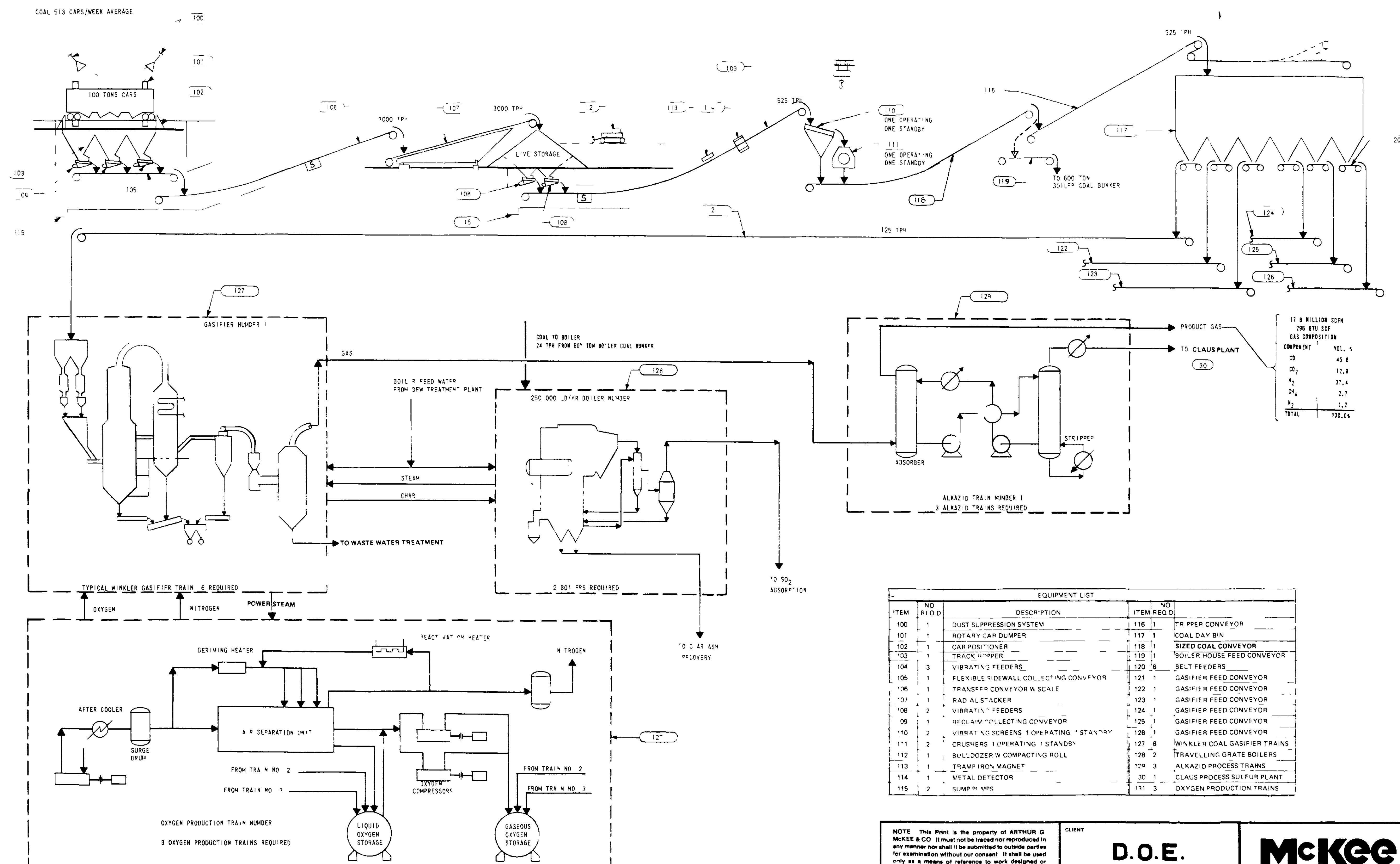
FORM M-1542 9/77

TITLE 405.6 MILLION GAL/YR
COAL-OIL MIXTURE PROCESS PLANT
PLAN AND ELEVATIONS

SCALE 1"=100' & 1"=50' E.R NO

4598-A
081478-2

REVISION



EQUIPMENT LIST

ITEM	NO REQ D	DESCRIPTION	ITEM	NO REQ D	
100	1	DUST SUPPRESSION SYSTEM	116	1	TRIPPER CONVEYOR
101	1	ROTARY CAR DUMPER	117	1	COAL DAY BIN
102	1	CAR POSITIONER	118	1	SIZED COAL CONVEYOR
103	1	TRACK HOPPER	119	1	BOILER HOUSE FEED CONVEYOR
104	3	VIBRATING FEEDERS	120	6	BELT FEEDERS
105	1	FLEXIBLE SIDEWALL COLLECTING CONVEYOR	121	1	GASIFIER FEED CONVEYOR
106	1	TRANSFER CONVEYOR W SCALE	122	1	GASIFIER FEED CONVEYOR
107	1	RADIAL STACKER	123	1	GASIFIER FEED CONVEYOR
108	2	VIBRATING FEEDERS	124	1	GASIFIER FEED CONVEYOR
109	1	RECLAIM COLLECTING CONVEYOR	125	1	GASIFIER FEED CONVEYOR
110	2	VIBRATING SCREENS 1 OPERATING 1 STANDBY	126	1	GASIFIER FEED CONVEYOR
111	2	CRUSHERS 1 OPERATING 1 STANDBY	127	6	WINKLER COAL GASIFIER TRAINS
112	1	BULLDOZER W COMPACTING ROLL	128	2	TRAVELLING GRATE BOILERS
113	1	TRAMP IRON MAGNET	129	3	ALKAZID PROCESS TRAINS
114	1	METAL DETECTOR	130	1	CLAUS PROCESS SULFUR PLANT
115	2	SUMP PUMPS	131	3	OXYGEN PRODUCTION TRAINS

CLIENT D.O.E.

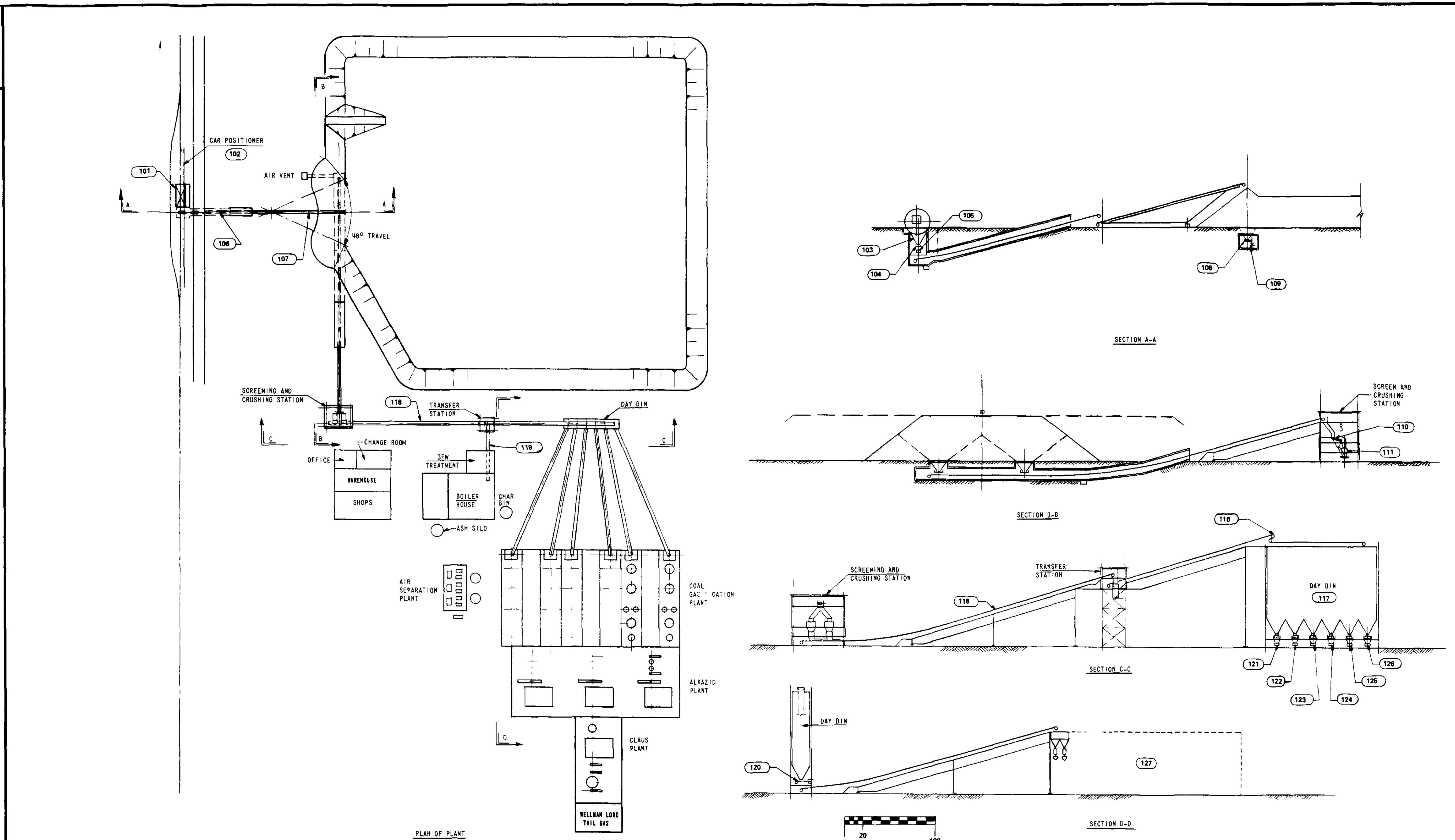
McKee

TITLE 17 8 MILLION SCFH COAL GASIFICATION PLANT FLOW DIAGRAM

SCALE E R NO

4598-A
081478-3

REVISION



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CLIENT: D. O. E.

McKee

FORM M 1542 9/77

TITLE: 17 8 MILLION SCFH COAL GASIFICATION PLANT PLANS & ELEVATIONS

SCALE 1"=100' & 1'=50' E.R NO

4598-A
081478-4

REVISION

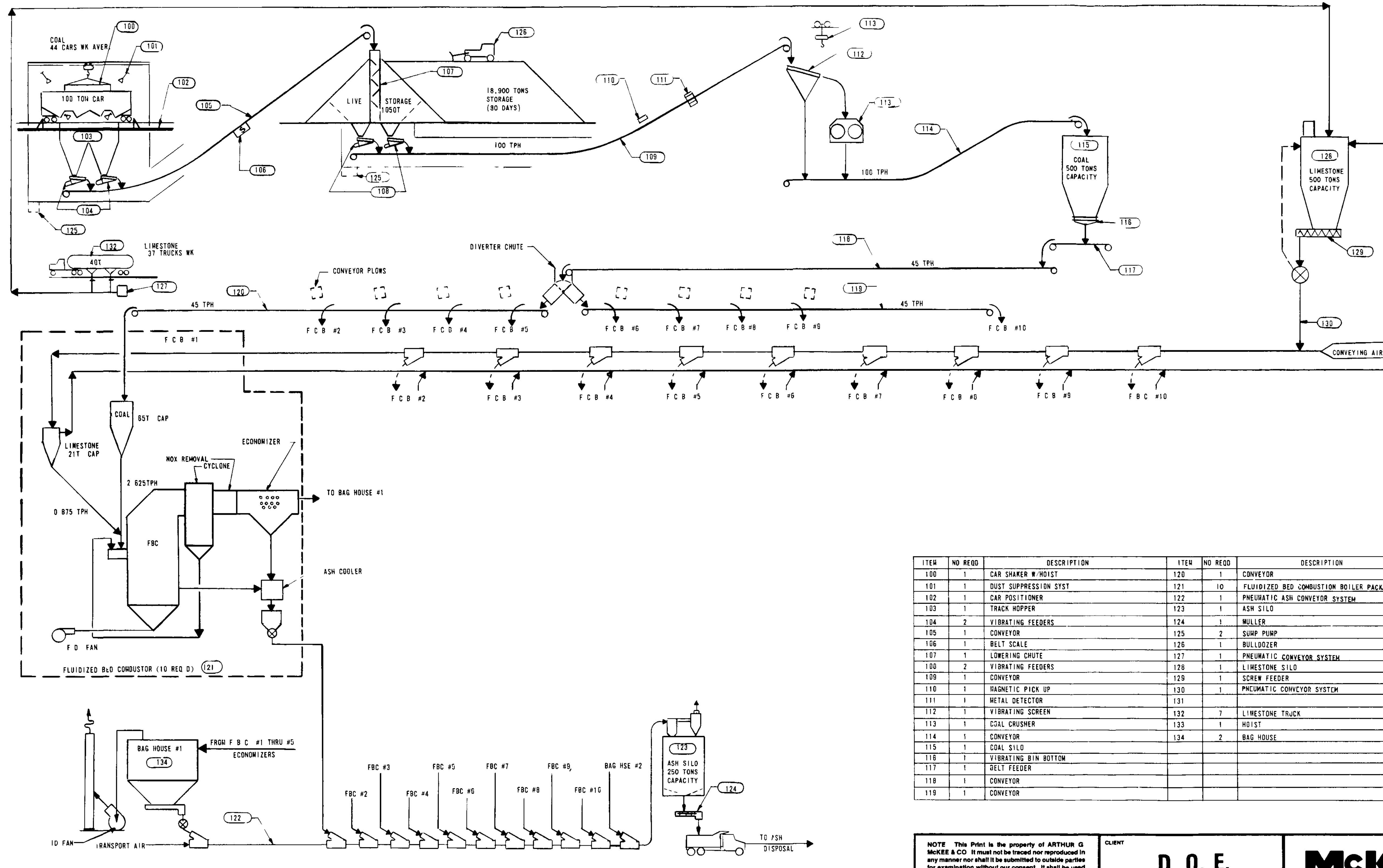
NO	DESCRIPTION	BY	CH	APPROVED	DATE

NO	DESCRIPTION	BY	CH	APPROVED	DATE

REFERENCES

	BY	DATE
DESIGNED		
DRAWN		
CHECKED		
APPROVED 1		
APPROVED 2		
APPROVED 3		

DATE TO	A	B	C	0	1	2	3	4	5	6	7
CLIENT											
FIELD											



ITEM	NO REQD	DESCRIPTION	ITEM	NO REQD	DESCRIPTION
100	1	CAR SHAKER W/HOIST	120	1	CONVEYOR
101	1	DUST SUPPRESSION SYST	121	10	FLUIDIZED BED COMBUSTION BOILER PACKAGE
102	1	CAR POSITIONER	122	1	PNEUMATIC ASH CONVEYOR SYSTEM
103	1	TRACK HOPPER	123	1	ASH SILO
104	2	VIBRATING FEEDERS	124	1	MULLER
105	1	CONVEYOR	125	2	SUMP PUMP
106	1	BELT SCALE	126	1	BULLDOZER
107	1	LOWERING CHUTE	127	1	PNEUMATIC CONVEYOR SYSTEM
108	2	VIBRATING FEEDERS	128	1	LIMESTONE SILO
109	1	CONVEYOR	129	1	SCREW FEEDER
110	1	MAGNETIC PICK UP	130	1	PNEUMATIC CONVEYOR SYSTEM
111	1	METAL DETECTOR	131		
112	1	VIBRATING SCREEN	132	7	LIMESTONE TRUCK
113	1	COAL CRUSHER	133	1	HOIST
114	1	CONVEYOR	134	2	BAG HOUSE
115	1	COAL SILO			
116	1	VIBRATING BIN BOTTOM			
117	1	BELT FEEDER			
118	1	CONVEYOR			
119	1	CONVEYOR			

CLIENT D. O. E.

McKee

FORM M-1542 9/77

TITLE 500M LB/HR FLUIDIZED BED COMBUSTION BOILER PLANT FLOW DIAGRAM

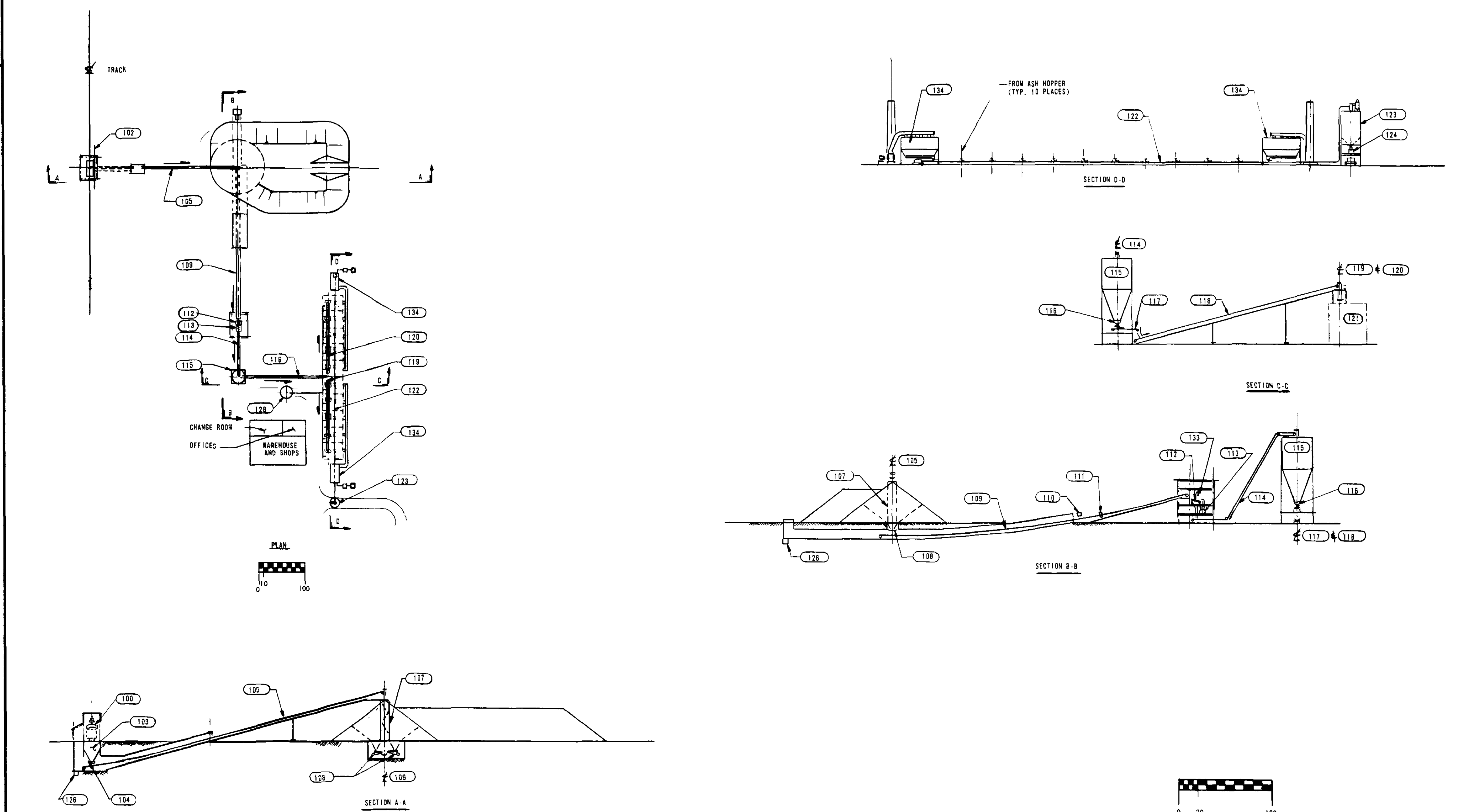
SCALE ______ E/R NO ______

4598-A
081478-5

REVISION

NO	DESCRIPTION	BY	CH	APPROVED	DATE	NO	DESCRIPTION	BY	CH	APPROVED	DATE

	BY	DATE	DATE TO	A	B	C	0	1	2	3	4	5	6	7
DESIGNED			CLIENT											
DRAWN			FIELD											
CHECKED														
APPROVED 1														
APPROVED 2														
APPROVED 3														



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CLIENT: D. O. E.

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FORM M-1542 9/77

TITLE: 500M LB/HR FLUIDIZED BED COMBUSTOR STEAM PLANS PLAN AND ELEVATIONS

SCALE 1"=100' & 1"=50' E/R NO

4598-A
081478-6

REVISION