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Pulsed Atmospheric Fluidized Bed Combustion

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EXECUTIVE SUMMARY

Coal-fueled technology appears to have a great potential for oil and gas replacement in small-scale installations of less than 6.3 kg/s (50,000 lb/hr) steam equivalent. These smaller units can meet the needs of process heat, hot water, steam, and space heating in the residential, commercial, and industrial sectors. Currently, oil- and natural gas-fired equipment are being used almost exclusively for these applications. Due to the large difference between the prices of these fuels and coal, a coal-fueled technology engineered for small-scale applications has the potential of becoming very competitive under economic conditions in which the price differential overcomes the initial capital cost of the coal-based system. From performance, emissions, and economics standpoints, atmospheric fluidized bed combustion (AFBC) technology has emerged as a leading candidate for utilizing high sulfur fuels. A successful coal-fueled FBC system cannot only be more economical, but can also reduce the nation's dependence on foreign oil and open up new markets for domestic coal and the coal-fueled fluid-bed technologies.

A market analysis indicated that a coal-based system that provides competitive levels of capital and O&M cost, performance, and reliability at the 0.126 to 1.26 kg/s (1,000 to 10,000 lb/hr) steam-generation rate can displace as much as 2.64 EJ (2.5 quad Btu) of gas and oil within the residential, commercial, and light industrial sectors. In the industrial sector, systems from 1.26 to 6.3 kg/s (10,000 to 50,000 lb/hr) steam can displace another 1.16 EJ (1.1 quad Btu) of energy per year.

Simply scaling-down existing large AFBC systems to the size range suitable for small end-use sectors of interest will result in complex and expensive systems that will not be competitive with presently available oil- and gas-fired equipment. New innovative approaches are needed to reduce cost and enhance performance. Specifically, the new system should possess the following attributes:

- High combustion efficiency;
- High SO_2 capture capacity;

- Low NO_x emissions;
- Reliability, maintainability and safety of operations equivalent to oil- and gas-fired packaged systems;
- Cost competitiveness with gas- and oil-fired systems;
- Simple and inexpensive controls;
- Rapid start-up and load-following capability;
- Clean, aesthetic, and compact design; and
- Minimal operator attention.

ThermoChem, under contract to the Department of Energy, conducted extensive research, development and demonstration work on a Pulsed Atmospheric Fluidized Bed Combustor (PAFBC) to confirm that this advanced technology can meet these performance objectives.

The ThermoChem/MTCI PAFBC system integrates a pulse combustor with an atmospheric bubbling-bed type fluidized bed combustor (BFBC). In this modular configuration, the pulse combustor burns the fuel fines (typically less than 30 sieve or 600 microns) and the fluidized bed combusts the coarse fuel particles.

Since the ThermoChem/MTCI PAFBC employs both the pulse combustor and the AFBC technologies, it can handle the full-size range of coarse and fines. The oscillating flow field in the pulse combustor provides for high interphase and intraparticle mass transfer rates. Therefore, the fuel fines essentially burn under kinetic control. Due to the reasonably high temperature (> 1093°C but less than the temperature for ash fusion to prevent slagging), combustion of fuel fines is substantially complete at the exit of the pulse combustor. The additional residence time of 1 to 2 seconds in the freeboard of the PAFBC unit then ensures high carbon conversion and, in turn, high combustion efficiency.

A laboratory unit was successfully designed, constructed and tested for over 600 hours to confirm that the PAFBC technology could meet the performance objectives. Subsequently, a 50,000 lb/hr PAFBC demonstration steam boiler was designed, constructed and tested at Clemson University in Clemson, South Carolina.

This Final Report presents the detailed results of this extensive and successful PAFBC research, development and demonstration project

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SECTION 1.0

INTRODUCTION TO PAFBC TECHNOLOGY

Many technologies have been developed and/or demonstrated for utilizing high-sulfur fuels in general and coals in particular. From performance, emissions, and economics standpoints, fluidized bed combustion (FBC) technology has emerged as a leading candidate for utilizing high sulfur fuels. Many FBC designs are available and are at various stages of commercialization. FBCs can be classified in terms of operating pressure (atmospheric or pressurized) and fluidization mode (bubbling or circulating). All the FBC designs possess attributes such as in-situ sulfur capture, no slagging or fouling of heat transfer surfaces, high heat transfer rates to heat exchange surfaces, near uniform temperature in combustion zone, and fuel flexibility. These features have made it possible for FBC technology to compete successfully for the large industrial boiler market (6.3-37.8 kg/s or 50,000 to 300,000 lb/hr steam). Large-scale (70 to 150 MWe) field demonstration projects are in progress to enable FBC commercialization in the utility sector. The potential of FBC technology, and specifically, atmospheric fluidized bed combustion (AFBC) for small-scale (<6.3 kg/s or 50,000 lb/hr steam equivalent) applications has, however, not been explored seriously until recently.[1]

An AFBC-based technology appears to have a great potential for oil and gas replacement in small-scale installations of less than 6.3 kg/s (50,000 lb/hr) steam equivalent. These smaller units can meet the needs of process heat, hot water, steam, and space heating in the residential, commercial, and industrial sectors. Currently, oil- and natural gas-fired equipment are being used almost exclusively for these applications. Due to the large difference between the prices of these fuels and coal, a coal-fueled AFBC technology engineered for small-scale applications has the potential of becoming very competitive under economic conditions in which the price differential overcomes the initial capital cost of the coal-based system. A successful coal-fueled AFBC system cannot only be more economical, but can also reduce the nation's dependence on foreign oil and open up new markets for domestic coal and the coal-fueled fluid-bed technologies.

A market analysis indicated that a coal-based system that provides competitive levels of capital and O&M cost, performance, and reliability at the 0.126 to 1.26 kg/s (1,000 to 10,000 lb/hr) steam-generation rate can displace as much as 2.64 EJ (2.5 quad Btu) of gas and oil within the residential, commercial, and light industrial sectors. In the industrial sector, systems from 1.26 to 6.3 kg/s (10,000 to 50,000 lb/hr) steam can displace another 1.16 EJ (1.1 quad Btu) of energy per year.

As pointed out earlier, the AFBC systems can be classified into bubbling-bed (BFBC) and circulating-bed (CFBC) FBC systems. In BFBC, it is critical to control the extent of fines (elutriable particles) in the coal and sorbent feed in order to limit particle carryover and its adverse effect on combustion and sulfur capture performance, emissions, and the size of solids collection equipment. Additionally, the higher Ca/S feed ratios typically required in BFBC applications tend to increase sorbent and waste disposal costs. Furthermore, the turndown capability of the BFBC is rather limited. As regards the CFBC, it exhibits higher combustion efficiency and sorbent utilization, lower NO_x emissions due to multiple air staging, and greater fuel flexibility and turndown as compared to BFBC. However, the CFBC system requires a tall combustor to accommodate sufficient heat exchange surface. This makes it both impractical and expensive to scale-down CFBC to sizes significantly smaller than 12.6 kg/s (100,000 lbs) steam equivalent.

Fluid beds in general tend to have large thermal inertia. Start-up of large fluid-bed systems requires a considerable amount of time and auxiliary subsystems to preheat the beds in a controlled manner. This adds to overall system cost and complexity. Concepts which provide for a simple compact design for fast start-up with low-cost hardware and also have simple operational characteristics are a must for small-scale applications. Thermal inertia of fluid beds also affects load following to some extent and this has also been a serious shortcoming for scale-down to small end-use applications. System designs must provide fast response to load changes, particularly through auxiliary firing subsystems and methods of bed heating. Such a design should not require additional hardware and control systems if the system capital cost is to be maintained sufficiently low to compete favorably with the existing oil and gas equipment. In addition, new designs capable of higher throughput for given combustor size will contribute to a reduction in capital cost per kJ/hr (Btu/hr) of fuel fired. This must be achieved, however, without compromising the pollution control performance of equipment intended to meet stringent requirements in some of these end-use applications.

Simply scaling-down existing large AFBC systems to the size range suitable for small end-use sectors of interest will result in complex and expensive systems that will not be competitive with presently available oil- and gas-fired equipment. New innovative approaches are needed to reduce cost and enhance performance. Specifically, the new system should possess the following attributes:

- High combustion efficiency;
- High SO_2 capture capacity;
- Low NO_x emissions;
- Reliability, maintainability and safety of operations equivalent to oil- and gas-fired packaged systems;
- Cost competitiveness with gas- and oil-fired systems;
- Simple and inexpensive controls;
- Rapid start-up and load-following capability;
- Clean, aesthetic, and compact design; and
- Minimal operator attention.

The ThermoChem/MTCI PAFBC system integrates a pulse combustor with an atmospheric bubbling-bed type fluidized bed combustor (BFBC) as shown in *Figure 1-1*. In this modular configuration, the pulse combustor burns the fuel fines (typically less than 30 sieve or 600 microns) and the fluidized bed combusts the coarse fuel particles. For the sake of readers unfamiliar with pulse combustion, a brief discussion of this concept is provided following a description of the PAFBC system.

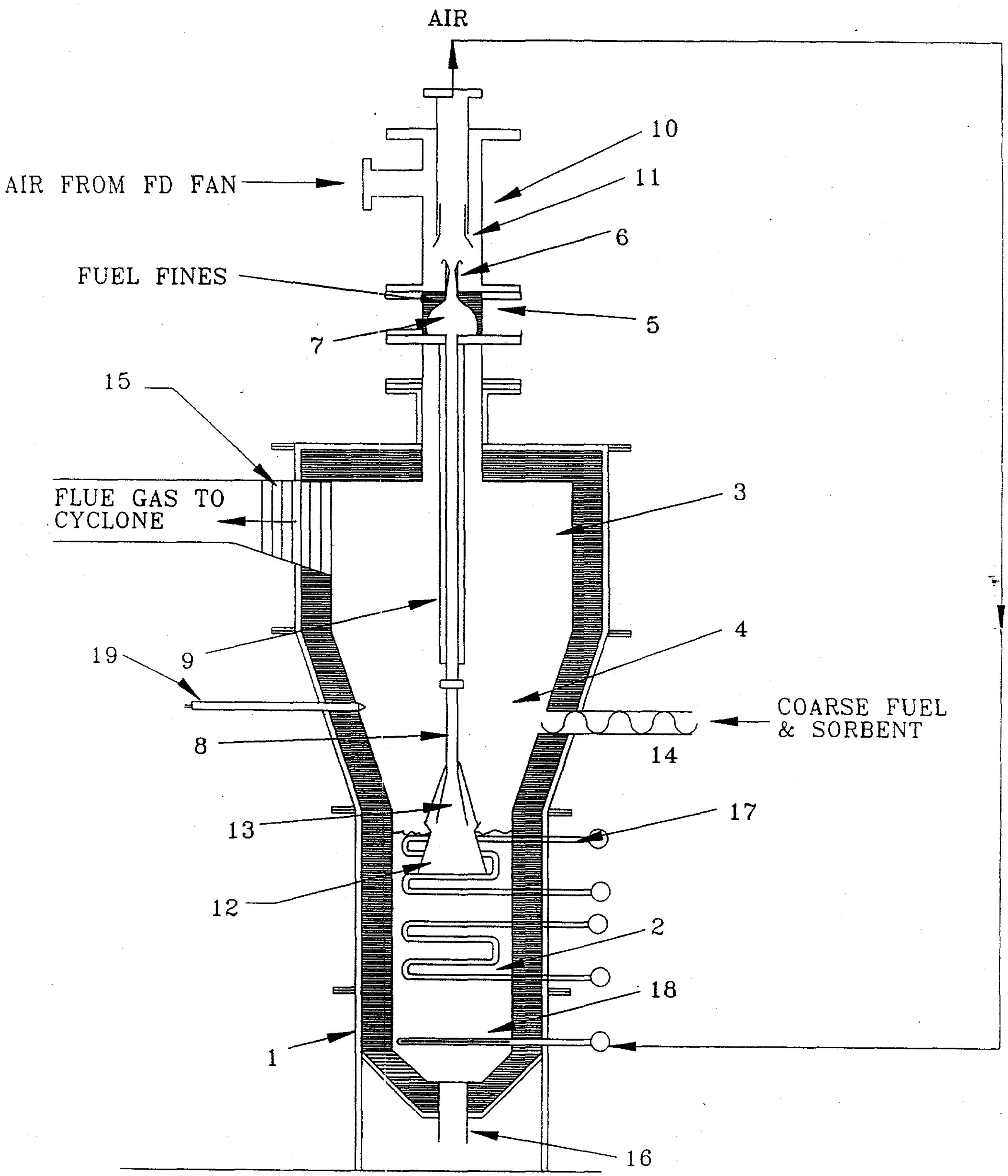


FIGURE 1-1: THERMOCHEM/MTCI PAFBC SYSTEM

As shown in Figure 1-1, the PAFBC comprises a refractory-lined fluidized bed (1) and a pulse combustor (5). The fluidized bed consists of a dense bed (2) and a freeboard (3). The dense bed operates in a bubbling/turbulent fluidization mode. The freeboard incorporates an expanded section (4) to decrease gas velocity, increase gas residence time and decrease elutriation. The pulse combustor (5) comprises an aerovalve (6), combustion chamber (7), tailpipe (8) with a water/steam jacket (9), air plenum (10), thrust augmentor (11), and a diffuser (12). The expansion section (13) at the end of the tailpipe (8) is to reduce the flue gas exit velocity and prevent channeling in the bed. After the flue gas from the pulse combustor exits the tailpipe, it enters a diffuser section which provides fines recirculation and increased particle residence time in the bed. The additional components of the PAFBC unit include overbed coarse fuel and sorbent screw feed system (14), a fuel classifier (not shown) for separating fuel feed into coarse fraction and fines, an inertial solids separator (15) for in-furnace solids disengagement, a bed drain (16) to facilitate rock removal, tube banks (17) immersed in-bed for steam generation and bed temperature control, an air distributor (18), and a natural gas-fired pilot burner (19) for operational safety during start-up.

A pulse combustor typically consists of a flow diode, combustion chamber and resonance tube. Fuel and air enter the combustion chamber. An ignition source detonates the explosive mixture during start-up. The sudden increase in volume, triggered by the rapid increase in temperature and evolution of combustion products, pressurizes the chamber. As the hot gas expands, the flow diode permits preferential flow in the direction of the resonance tube. Gases exiting the hot combustion chamber in the resonance tube possess significant momentum. A vacuum is created in the combustion chamber due to the inertia of the gases within the resonance tube. The inertia of the gases in the resonance tube permits only a small fraction of exhaust gases to return to the combustion chamber; the balance of the gas exits the resonance tube. Since the chamber pressure is below atmospheric pressure, air and fuel are drawn into the chamber where autoignition takes place. Again, the flow diode constrains reverse flow, and the cycle begins anew. Once the first cycle is initiated, engine operation is self-sustaining.

The flow diode utilized in many other pulse combustion concepts is a mechanical "flapper valve." The flapper valve is actually a check valve permitting flow from inlet to chamber, and constraining reverse flow by a mechanical seating arrangement. This served quite well for the purpose intended. MTCI's pulse combustor technology is designed for a much longer service life utilizing an aerodynamic valve without moving parts as an effective alternative to the flapper valve. During the exhaust stroke, a boundary layer builds in the aerodynamic valve. Turbulent eddies choke off much of the reverse flow. Moreover, the exhaust gases are of a much higher temperature than the inlet gases. Therefore, the viscosity of the gases is much higher and the reverse resistance of the inlet diameter, in turn, is much higher than that for forward flow through the same opening. These phenomena, along with the high inertia of the exhausting gases in the resonance tube, combine to yield preferential and mean flow from inlet to exhaust. Thus, the non-mechanical pulse combustor is a self-aspirating engine, drawing its own air and fuel into the combustion chamber and auto-ejecting combustion products.

The rapid pressure oscillations in the combustion chamber generate an intense oscillating flow field. In the case of coal combustion, the fluctuating flow field causes the products of combustion to be swept away from the reacting solid, thus providing access to oxygen with little or no diffusion limitation. Second, pulse combustors experience very high mass transfer and heat transfer rates within the combustion zone. While these combustors tend to have very high heat release rates (typically ten times those of conventional burners), the vigorous mass transfer and high heat transfer within the combustion region result in a more uniform temperature. Thus, peak temperatures attained are much lower than in the case of conventional systems. This results in a significant reduction in nitrogen oxide (NO_x) formation. The high heat release rates also result in a smaller combustor size for a given firing rate and a reduction in the residence time required.

The performance of AFBC is affected by the rate of combustion of coal, which in turn is affected by coal properties (devolatilization, swelling, fragmentation, and char combustion), feed particle size range, feed system and combustion-enhanced mechanical attrition, heat and mass transfer rates, and unit operating conditions.(2) In AFBC, the carbon carryover into the primary particle separator is generally high due to limited residence time of fuel fines in the combustor.

The acoustic field radiated into the fluid bed enhances the mass transfer rate and in turn increases the reaction rate between the sorbent and SO_2. This a priori sulfur release, acoustic enhancement in the fluid-bed mass transfer process, and the fines recirculation as a consequence of the draft tube design help achieve high sulfur capture efficiency at low Ca/S molar feed ratio, which leads to lower limestone and waste disposal costs.

Pulse combustors are inherently low NO_x devices.[3] Keller and Hongo[4] investigated the mechanisms of NO_x production in pulse combustion and concluded that several complementary mechanisms are responsible. The rate of heat transfer in the pulsating flow is higher than that in conventional steady flow and helps create lower overall temperature in the combustion chamber. Also, the high rates of mixing between the hot combustion products and the cooler residual products from the previous cycle and the incoming cold reactants create a short residence time at high temperature quenching the NO_x production. These complementary mechanisms create an environment which approximates a well-stirred tank at relatively low temperature and result in low NO_x production. The dense fluid bed, due to operation at low temperature (~ 843°C) and with coarse fuel particles, enjoys a lower NO_x production as well. Consequently, the NO_x emissions from PAFBC are likely to be lower than that of AFBC.

The overall heat transfer coefficient in the water-jacketed pulse combustor tailpipe is of the same order as that for tubes immersed in the dense fluidized bed. The replacement of the inefficient heat exchanger in the freeboard of a conventional BFBC by the water-jacketed pulse combustor tailpipe significantly decreases the heat transfer surface area requirement and cost.

1.1 BACKGROUND

Bubbling atmospheric fluid-bed combustors have a number of technical attributes that hold a great deal of promise in combustion of high sulfur, high ash coals having fuel-bound nitrogen, with inherent pollution control capabilities. The most relevant technical attributes of bubbling atmospheric fluid-bed combustors and a discussion of the degree to which the technology realized its potential is discussed below.

The fundamental concept of a bubbling atmospheric fluid-bed combustor is to burn crushed coal at a moderate temperature in a bed comprised of particles that are made of a sulfur sorbent. The ability to burn the coal at a moderate temperature (1550°F to 1650°F) in a fluid bed is aimed at control of NO_x formation from fuel-bound nitrogen and thermal sources. Limestone, generally used as the bed material, calcines at the operating temperature providing a calcium oxide sorbent that captures sulfur oxides produced from burning the sulfur-containing coal fuels.

Heat removal from the bed to control the bed temperature to the desired range is indeed opportune with bubbling fluid beds due to the generally high heat transfer available between the bed material and surfaces immersed in the bed. This provides for relatively small heat transfer surface to extract useful heat from the combustion process.

As the development of atmospheric fluidized beds proceeded, it became evident that coal fines tend to quickly elutriate from the fluid bed that is normally operating at a fluidization air velocity suitable for the bulk of the crushed coal particle size (say, 3/8" x 28 mesh) and the corresponding selection of limestone particle size. Generally, the higher the fluidization velocity, the higher the heat release rate per square foot of bed area. The higher heat release rates are desirable to reduce the specific capital cost of the equipment per Btu/hr of firing rate. Nevertheless, limitation on fluidization velocity was encountered because of residence time, bed height, particle size distribution and related bed stability considerations. Practical considerations and developmental experience gave rise to bubbling fluid beds that operate within the range of 3 to 5 feet in height with a large freeboard required above the bubbling bed.

The freeboard was provided to permit sufficient residence time at a suitable temperature for the combustion of fines that are found with crushed coal as received at the fluid-bed combustor fuel inlet. Coal screening and classification became necessary to remove excessive amounts of fines prior to feeding the fuel to a fluid-bed combustor so as to maintain the operation of the equipment within the environmental performance constraints. This, however, tends to compromise the economics of the operation significantly. Most, if not all, bubbling atmospheric fluid bed manufacturers only warrantee performance with fuel fines content (28 mesh x 0) less than 7 to 10 percent of the total feed.

While bubbling fluid-bed combustors are designed with an appropriate freeboard height and volume with suitable operating temperatures, practical problems still accrue due to the presence of elutriated fines in the freeboard.

The carbon burnout of the fines is not the primary problem because such fines can be captured in a cyclone and/or a baghouse and recycled into the bed, preferably educted with air in an underbed feed, to improve the ultimate carbon burnout. Nevertheless, production of excessive amounts of CO can be experienced from fines burning in the freeboard having insufficiently high gas temperature. Thus, heat removal from the freeboard must be mindful of such considerations. This is particularly important because *gas flow is only once-through the entire system.*

Should the rate of elutriation of fines increase, however, temperatures in the freeboard could rise sufficiently to levels that promote NO_x production in the flue gas even though under such conditions the CO content in the flue tends to decrease. Further increase in the rate of fines elutriation could once more promote production of CO with a modest reduction in NO_x production from combustion in the freeboard.

In addition to the above, fines burning in the freeboard release sulfur oxides *downstream* of the sorbent bed. Some sulfur capture by sorbent fines elutriating form the bed is encountered but generally not sufficient. This is particularly the case when the coal being fed to the combustor contains excessive amounts of sulfur-laden fines.

Therefore, the design of the freeboard height and volume as well as the allowable heat loss (or removal) from the freeboard must be mindful of the fines combustion considerations as well as the need for a sufficient particle disengagement height for the return of larger particles ejected by the bubbling action at the surface of the bed.

Practical experience with bubbling fluid beds further indicated that gas bubble growth in the bed does occur and ultimately causes channeling of the gas flow through the bed. This tends to compromise both the combustion process efficiency (high excess air detected as high O_2 concentration in the flue) and low sorbent utilization efficiency with sulfur oxides breaking through.

In response to the above practical problem, with what is otherwise a very promising technology, much of the industry moved to pursue the circulating fluid bed. Circulating fluid beds operate at high fluidization velocities with much of the solids elutriating from the bed and later captured and circulated back to the bed. In this case there is no freeboard per se, but there is coal and sorbent material, with heat removal practiced to alleviate some of the above issues, throughout the entire height of the combustor vessel.

Heat transfer surface is typically installed on the walls of the CFB combustor. This surface is much less effective than surfaces immersed in bubbling beds, therefore, CFRs require as much as three to five times the surface area than bubbling beds of the same steam-generating capacity.

With high fluidization velocities, however, came the excessive height of the circulating fluid bed as dictated by the need for reasonable residence times and the need to provide sufficient heat transfer area. Circulating fluid beds are often 100 feet high or even higher which in turn caused the circulating fluid bed to be high in capital cost.

The Department of Energy recognized the need for advanced technology concepts and as a result it sought new advanced fluid-bed combustion technology so as to provide alternatives to the above technologies (AFB and CFB) ameliorating the technical and cost issues. The DOE program also challenged the community to provide innovative technology that can even be scaled down to smaller commercial and light industrial applications (below 50,000 lb/hr of steam) while maintaining their economic viability and environmental compliance.

Under this DOE program, which was underway in 1987, three technologies were pursued by DOE/METC, namely, a) the multi-solids fluidized bed combustor (MSFBC), b) the two-stage circulating fluid-bed combined with cyclone combustor (vortex fluidized bed), and c) the MTCI Pulse Stabilized Atmospheric Fluidized Bed Combustor.

The MSFBC technology is depicted in *Figure 1-2*. The combustion system is comprised of a combined bubbling-bed/circulating bed for combustion and a bubbling bed with in-bed heat exchangers for heat removal.

The combined bubbling bed/circulating bed combustor section employs an entrained bed of fine ash and limestone particles superimposed on a fluidized dense bed of large particles, which are not circulated. The lower dense bed of large particles is intended to promote mixing and increase the residence time of the circulating fine material, include fuel fines and fine limestone particles, within the combustor. The combustor section operates at high superficial gas velocity of 25 to 30 ft/sec. In this technology the heat transfer and combustion processes are affected in two separate vessels. Most of the heat removal occurs in an external heat exchanger (EHE) and a convective boiler.

The EHE is comprised of a conventional fluidized bed of fines that are captured by the circulating fluid-bed cyclone with heat exchanger tubes buried in the bed. This bed is fluidized at a low fluidization velocity (1 to 2 ft/ sec). The cooled fines are then reintroduced into the combined bubbling-bed/circulating bed combustor.

The two-stage circulating fluid-bed combined with cyclone combustor or the Vortex Fluidized Bed is depicted in *Figure 1-3*. In this technology a circulating fluid-bed combustor is provided and is operated with approximately 25 percent of the combustion air required at full load, thus maintaining the CFB combustor section with a relatively low cross-sectional area. The balance of the heat release (75%) is affected in the cyclone through tangential air injection in the cyclone barrel. Heat removal is practiced in the two-side bubbling beds with heat exchanger tubes

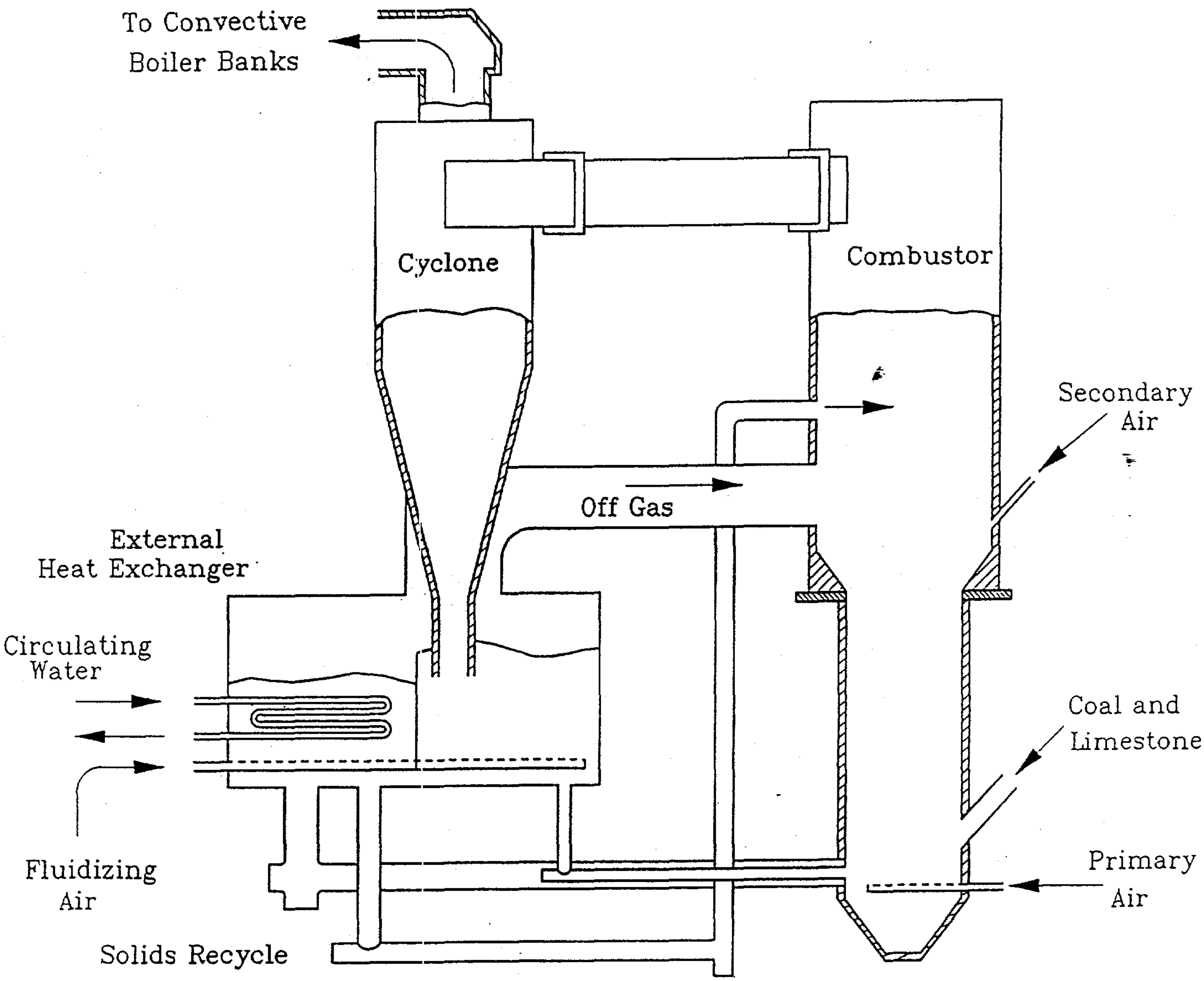


FIGURE 1-2: THE MSFBC TECHNOLOGY

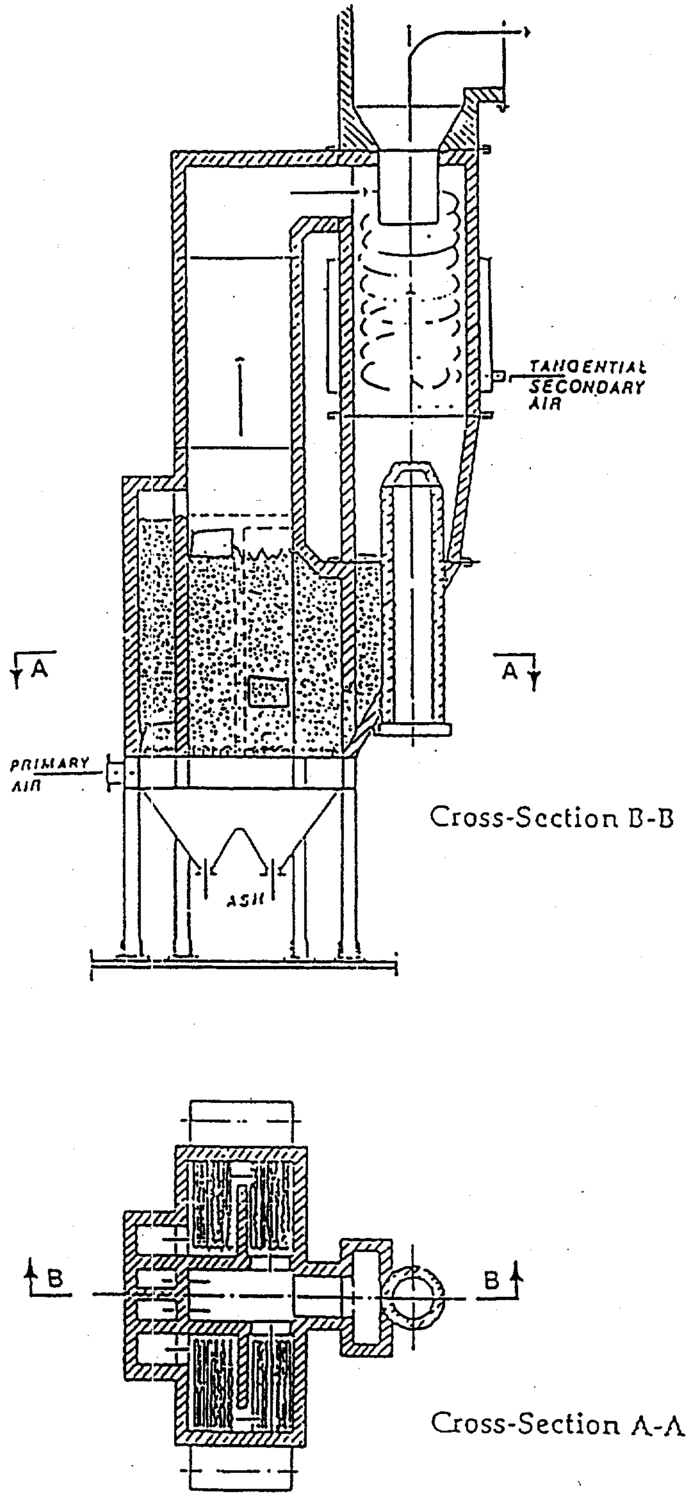


FIGURE 1-3: VORTEX FLUIDIZED BED

buried in the two-side bubbling beds. An elaborate scheme to circulate the bed material to and from the bubbling beds back into the CFB combustor is provided to control heat removal and combustor temperature. *Figure 1-4* provides an illustration of this solids reinjection scheme.

With the above review of the technology background and the relevant state of the art, we now present the Pulse Atmospheric Fluidized Bed Combustor Technology and a discussion of its technical attributes as they relate to the technology background and the state of the art.

The PAFBC technology was formulated on the basis of a set of performance and economic criteria that were established by an intense study of the fluid-bed technology including the current state of the art. This was undertaken to ensure that the resulting technology can in fact meet the technical, economic and operational goals set forth in the DOE advanced fluid-bed program objectives. The criteria for the PAFBC technology were as follows:

- Heat transfer in bubbling beds is one of the good features of the fluid-bed technology and this feature should be retained.
- Having to go to a circulating fluid bed to deal with fines in bubbling bed combustion adds too much to the capital cost and should be avoided.
- Fines combustion should be upstream of the bubbling bed (not downstream of the sorbent) to maximize sulfur capture efficiency of the system and sorbent utilization since gas flow through the system is once through.
- A means should be found that would minimize fines combustion in the freeboard to, in turn, minimize SO_2, NO_x and CO emissions in the freeboard.
- A means should be found to allow the bed to operate at reasonably high fluidization velocity without the formation of large bubbles and channeling flow. This is to enhance throughput without compromising combustion efficiency, controlling gas bypass, and enhancing sorbent utilization.

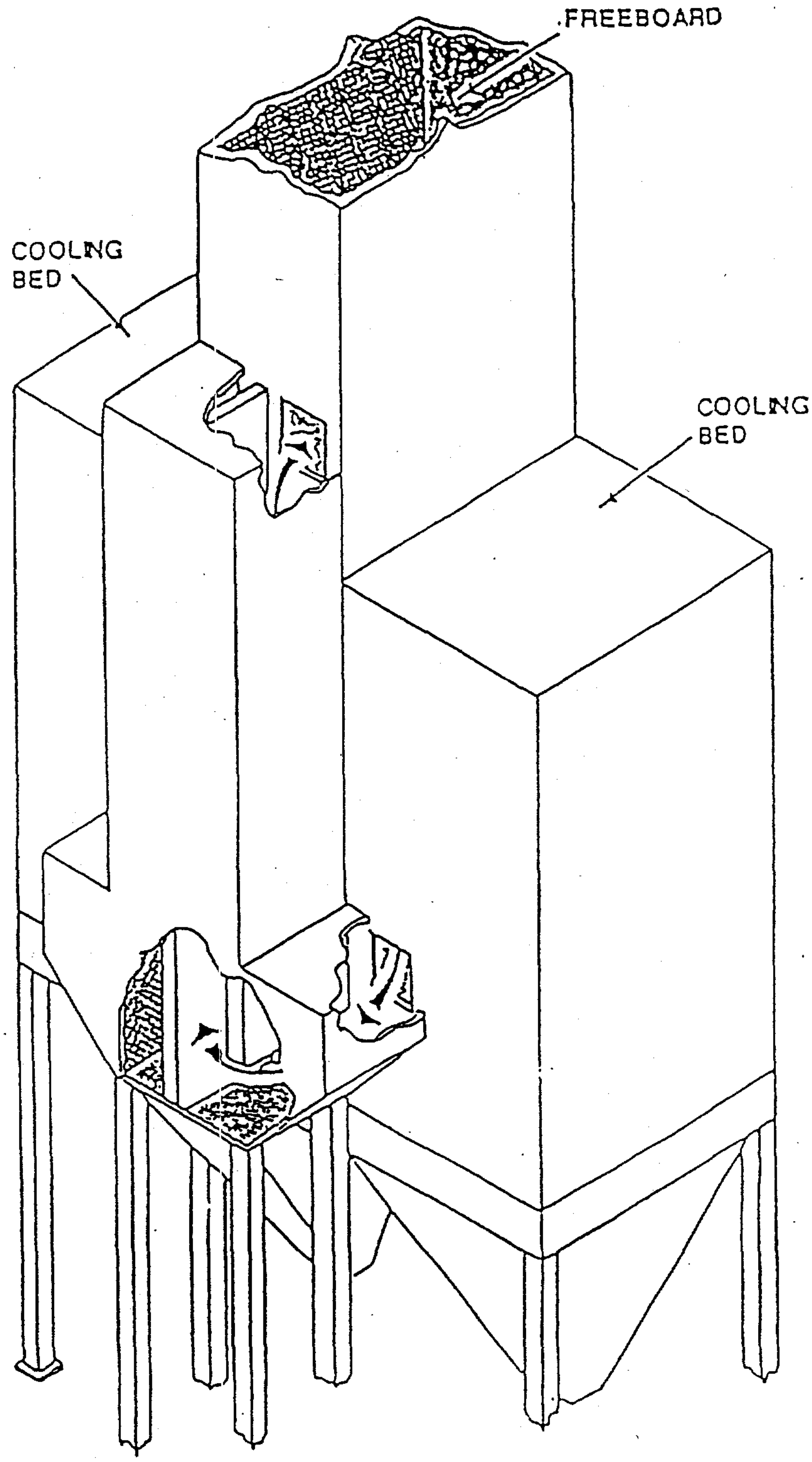


FIGURE 1-4: ISOMETRIC VIEW OF THE SOLID REINJECTION SYSTEM

- The entire system design must achieve compactness, low-capital cost and reliability of operation without elaborate hot solids handling and complex control systems.

- The system should be capable of modularization for adaptation to both industrial and utility needs.

In order to achieve the above, MTCI introduced the PAFBC technology which is a hybrid combustion system that employs both a pulse coal combustor and a bubbling fluid bed. The hybrid combustor system configuration is depicted in *Figure 1-5*.

The oscillating jet flow leaving the tailpipe, of the down-fired pulse combustor, can have a velocity in the range of 150 to 1600 ft/sec by design. A divergent tailpipe (not shown) can be employed to obtain slower exit velocities from the resonance tube. The kinetic energy in this jet of flue gas is employed to affect pressure recovery in a coupler to couple the pulse combustor to the fluid bed.

The dynamic pressure oscillations emanating from the coupler impart a forced oscillation to the fluid bed which is typically at frequencies in the range from 35 to 60 Hz. This forced oscillation causes large bubbles normally formed in a bubbling bed to break up into smaller bubbles, minimizing gas breakthrough and higher sulfur capture and combustion efficiency in the bed at higher fluidization velocities (and hence higher throughput).

The combustion air for the pulse combustor enters through the pulse combustor's air plenum. The aerovalve oscillatory flow is employed by a thrust augmentor device to provide a jet pump action giving rise to an increase in air pressure to the wind box. Typical pressure boost achieved is in the range of 8 to inches of water which subsidizes a significant portion of the pressure drop across the distributor plate at the bottom of the fluid bed.

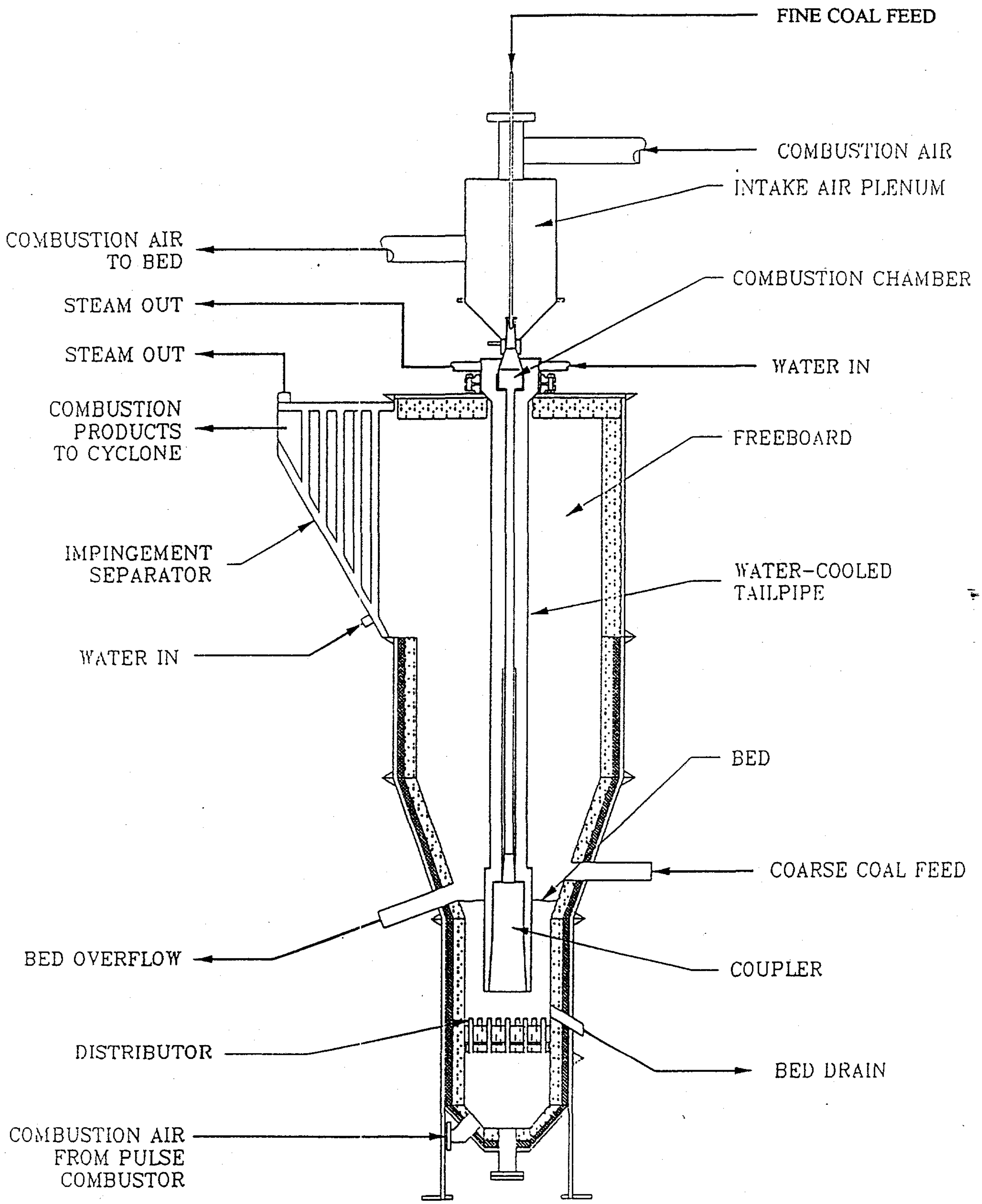


FIGURE 1-5: PULSED ATMOSPHERIC FLUIDIZED BED COMBUSTOR (PAFBC) SYSTEM

In addition to the above, the fluid-bed combustor design employs an expanded freeboard cross-sectional area to permit shorter freeboard height but with adequate freeboard volume. The considerations that entered into this portion of the design are as follows.

First, the top of the expanded fluid bed is maintained at the start of the cross-section expansion zone by the location of the overflow drain. This is aimed at slowing down the mean superficial velocity as the upward gas flow approach the top of the expanded bed. This in turn is intended to further reduce the vigor by which the bed ejects solids upward in the freeboard as it operates.

Second, the pulsed bed tends to form smaller bubbles, which was experienced with both directly and indirectly coupled beds, therefore, the disengagement height need not be very high.

Third, the superficial velocity in the expanded cross-section freeboard would be lower than that in a freeboard without the expansion which would promote the return of ejected solids, within reasonable limits, back to the bed.

Fourth, the expanded bed would have lower surface to volume ratio, hence, a modest reduction in equipment cost.

The cross-section of the exit opening in the freeboard is made sufficiently large to avoid entrainment of ejected solids with the departing flue gas leaving the freeboard of the fluid-bed combustor. The cross-section, however, is reduced progressively as the flow approaches the exit to increasingly remove both finer particles and heat from the flue with the minimum pressure drop possible. The lower portion of this solid separator is inclined at a sufficiently steep angle to ensure the return of the separated solids back to the bed.

Reflecting on the criteria and the technology review (including the state-of-the-art), the following discussion is devoted to the PAFBC technology design and operation attributes with the state-of-the-art technology as a basis for comparison.

In the PAFBC technology a CFB process was avoided and instead a compact high heat release pulse coal combustor was employed to burn the fines *upstream* of a bubbling fluid-bed. This is to ensure that NO_x, SO_x or even CO that may be produced in the pulse combustor will pass through a reburn zone with in-situ sorbent at the appropriate operating temperature.

The capital cost of a pulse combustor which is made of carbon steel is simply dwarfed by any CFB. The pulse combustor is cooled by raising steam from its resonance tube(s) which exhibit very high heat transfer.

The PAFBC hybrid technology provides a low capital cost, simple design, promoting high reliability, efficiency and environmental control unparalleled without a CFB but retaining and, in fact, enhancing the fundamental benefits of a bubbling fluid-bed system.

SECTION 2.0

DESIGN AND FABRICATION OF TEST UNITS

2.1 PRELIMINARY DESIGN UNIT TEST PROGRAM

During a Preliminary Design Unit Test program, a laboratory-scale model of the PAFBC was operated according to site-specific specifications for both adiabatic (drying) and non-adiabatic (steam) conditions. Technical and economic evaluations and a market analysis were also completed. A preliminary design of a proof-of-concept (POC) PAFBC facility based on the test results was completed.

The laboratory PAFBC test facility was characterized over a range of coal types and feed rates. Coals tested included two coal preparation plant refuse streams and three coal products representative of the fuels that will be used in the proof-of-concept system test. The performance of the PAFBC exceeded that of conventional bubbling fluidized bed combustors in all cases and at least matched the performance of circulating fluidized bed combustors.

Overall performance of the PAFBC exceeded design expectations; throughput of the system was nearly double that anticipated. Enhancement of heat transfer was demonstrated to be sufficient to permit incorporation of all required heat transfer surface as extended surfaces on the water-cooled eductor. Emissions from the PAFBC were experimentally determined to exceed all new source performance standards.

A cross-section drawing of the laboratory PAFBC test facility is shown in *Figure 2-1*.

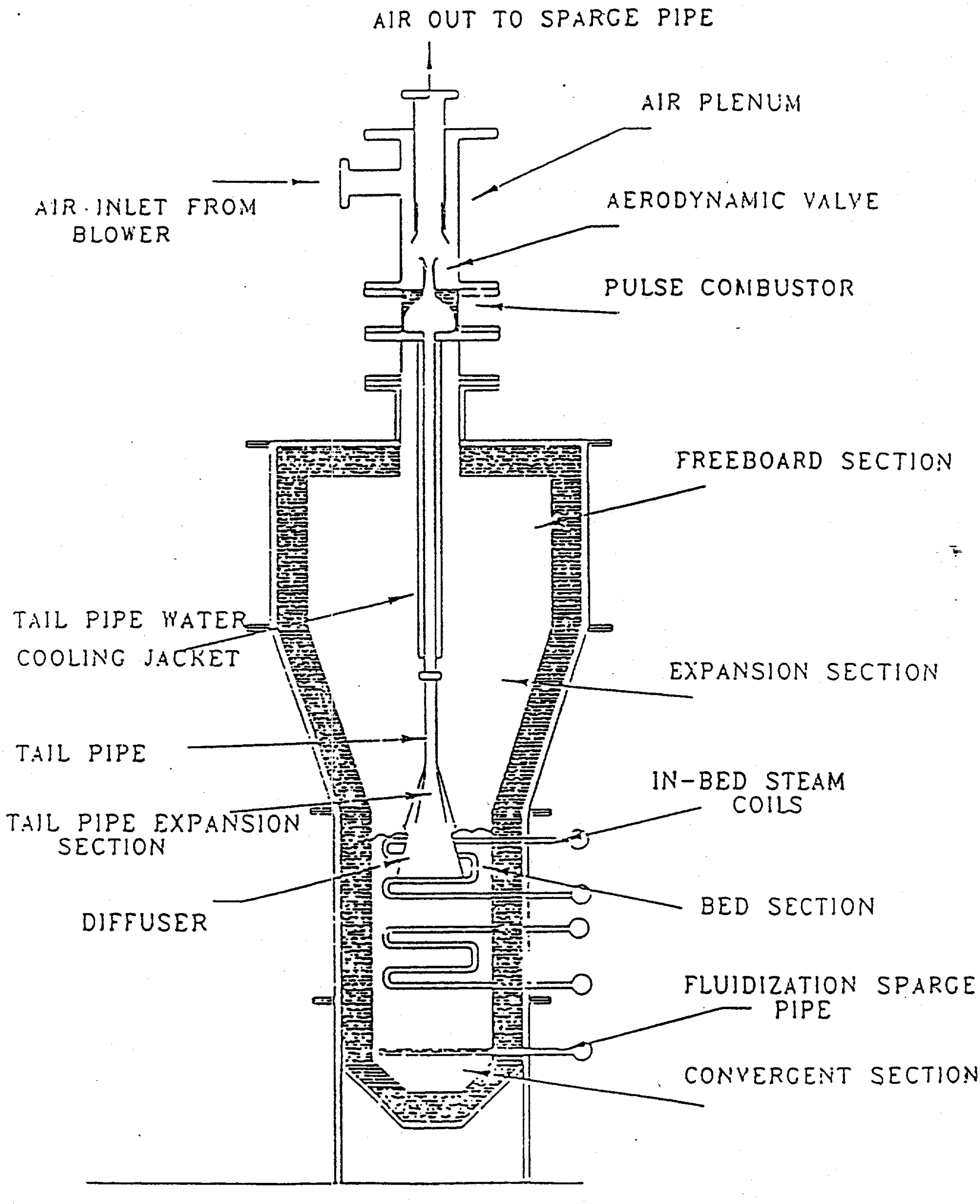


FIGURE 2-1: PULSED ATMOSPHERIC FLUIDIZED BED COMBUSTOR (PAFBC) PRELIMINARY DESIGN UNIT

2.1.1 Laboratory-Scale Design Summary

Pulse Combustor Design

Figure 2-2 is a drawing of the water-cooled pulse combustor and tailpipe assembly. The combustion chamber and tailpipe are designed for a 60 Hz resonant mode with the tailpipe length corresponding to 1/4-acoustic wave length. The thermal capacity of the combustion chamber under these conditions is nominally 500,000 Btu/hr. Non-slagging operation of the pulse combustor requires that the maximum combustion temperature be maintained below 2200°F. This condition is met when the excess air to the combustor is about 70 percent. Since both the combustion chamber and tailpipe are water-jacketed, substantial cooling of the flue gas occurs before exit from the tailpipe. A design tailpipe exit temperature of 1685°F has been derived on the basis of MTCI's pulse combustor design calculations. The maximum allowable flue gas temperature in the combustion chamber determines the required excess air to chamber, which in turn determines aerovalve sizing. The design thermal input to the pulse combustor together with the excess air fraction to maintain non-slagging combustion temperature, determines the total flue gas mass flow rate from the combustor. The design 1685°F tailpipe exit temperature then completely characterizes the heat transferred as steam/water in the jacket around the combustor/ tailpipe assembly.

The pulse combustor also has a thrust augmentor which is a feedback of the high pressure of the tailpipe exit flue gas to the pulse combustor inlet. Recirculation of a small amount of the flue gas through the combustor promotes pulse combustion through improved fuel/air mixing within the combustor.

Fluidized Bed Design

The fluidized bed used in the laboratory-scale testing during this phase of the development project has inside dimensions at the bed level of 2' x 2'. The bed section is approximately three feet tall at which point the vessel walls begin a transition to the freeboard dimensions of 2' x 4'. This transition occurs over a height of 6 feet. A drawing of the fluidized bed showing the location of the pulse combustor extending down through the freeboard is shown in *Figure 2-3*.

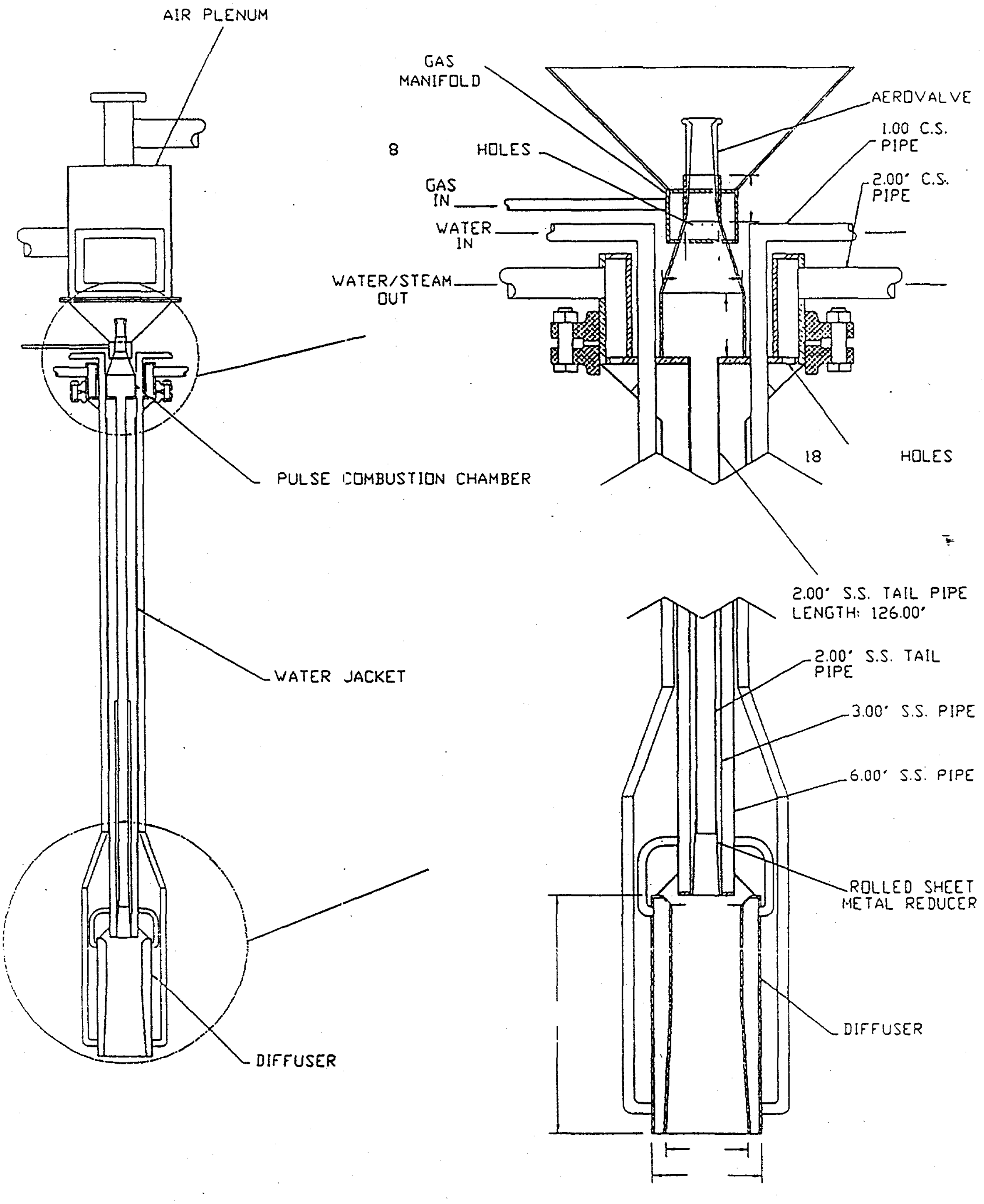


FIGURE 2-2: DESIGN OF PULSE COMBUSTOR UNIT

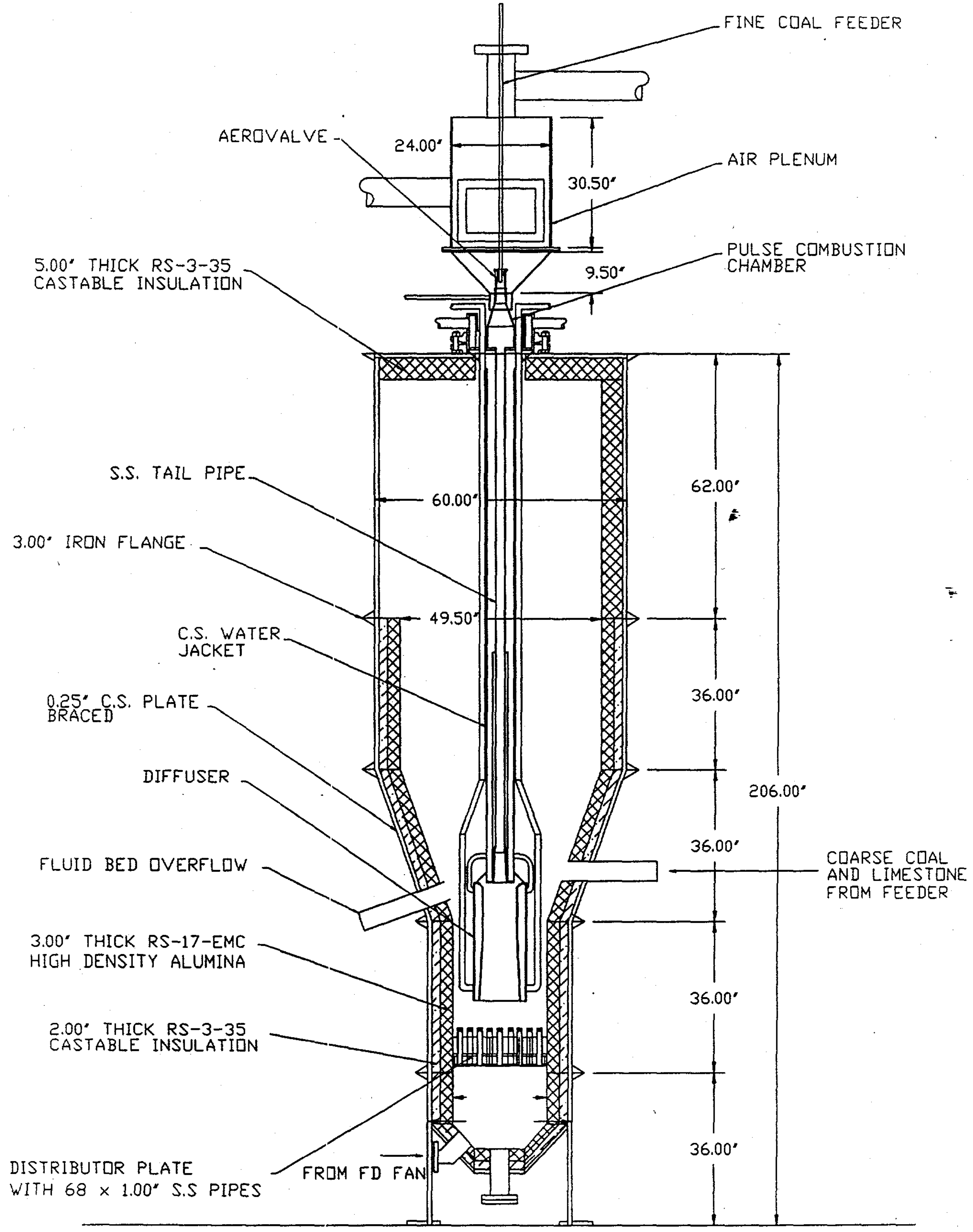


FIGURE 2-3: FLUID BED AND PULSE COMBUSTOR UNIT

The fluidized bed is designed to operate with a superficial velocity of 8 ft/sec. The design firing rate for the fluidized bed is approximately 1.5 MMBtu/hr. This rate of firing, added to the firing rate of the pulse combustor described above, provides a system design firing rate of about 2 MMBtu/hr in the near adiabatic mode of operation.

Particle Separator/Convective Heat Exchanger

This section of the laboratory-scale facility was designed to provide a smooth increase in gas velocity from the fluidized bed freeboard to the cyclone inlet. In addition, by installing vertical tubes in this section, the gas can be cooled prior to entering the cyclone and particles can be separated by impingement on the tubes.

At design conditions, gas leaving the freeboard enters the cross-flow heat exchanger at a temperature of 1550°F and is cooled to below 900°F before entering the cyclone. A pair of fins is attached along the length of each of the water-cooled staggered tubes so as to provide inertial capture of larger coal and fly ash particles as the flue gas flows around the tubes. This convective section is designed with an inclined base to promote the return of the inertially collected particulate to the fluidized bed, under gravity. The inclined base also allows a smooth transition from the 5 ft/s design gas velocity in the freeboard to a cyclone inlet gas velocity exceeding 50 ft/s. A sketch showing the fabrication of the particle separator is shown in *Figure 2-4*.

Cyclone

The design flue gas flow rate exiting the convective heat exchanger was used to size a high efficiency cyclone based on the classical Stairmand design. *Figure 2-5* shows the dimensions of the cyclone used in the laboratory unit. The design aerodynamic cut diameter (d_{50}) for this cyclone is 4 μm. The PAFBC cyclone is constructed with a water jacket around its outer wall. This jacket has a twofold purpose: (i) it allows construction from mild steel material, and (ii) it allows additional surface area for steam production. The latter is an advantage only in the case of non-adiabatic operation. Steam production needs to be minimized in adiabatic operation;

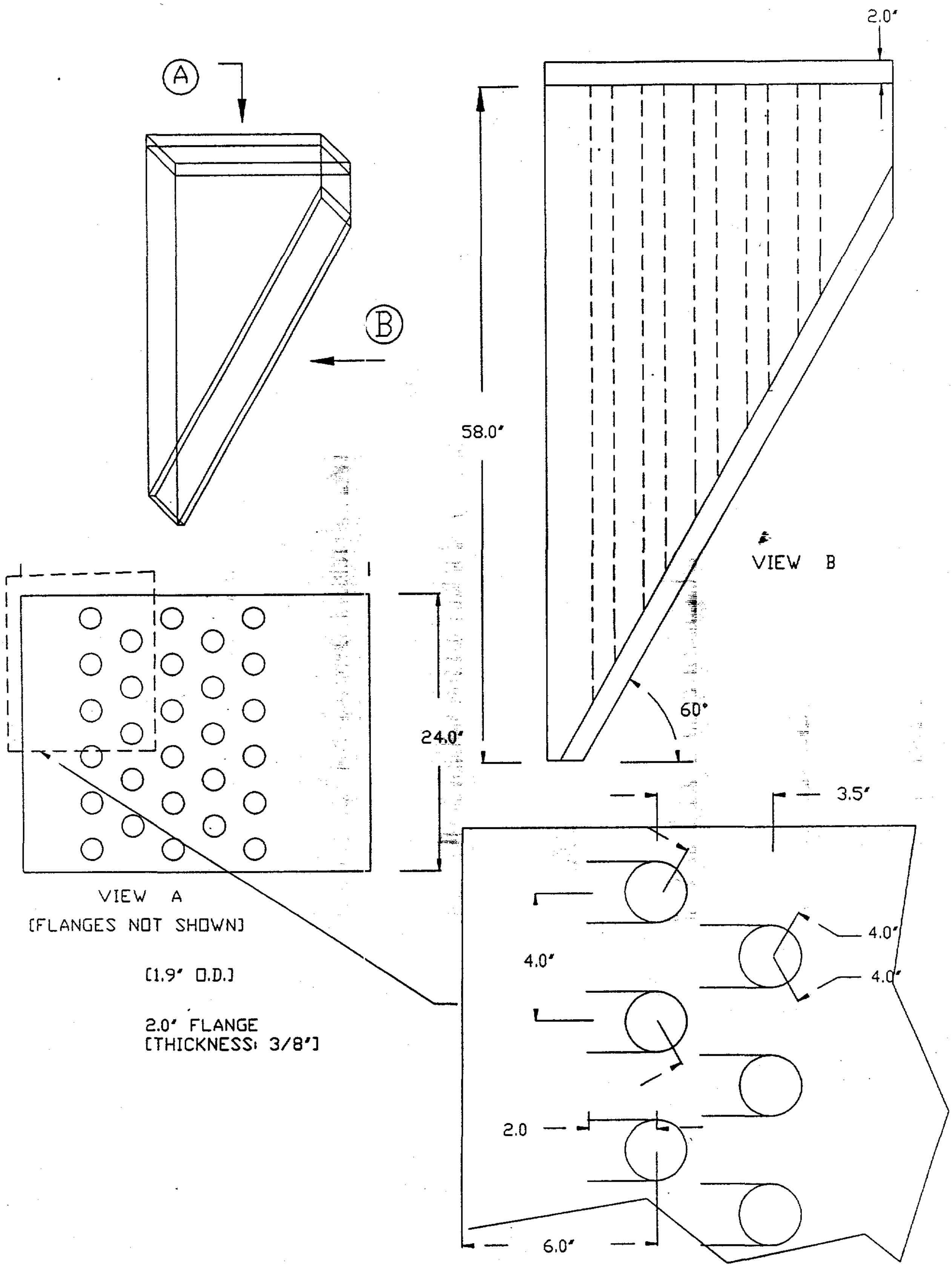


FIGURE 2-4: DESIGN OF CONVECTIVE HEAT EXCHANGER FOR PAFBC

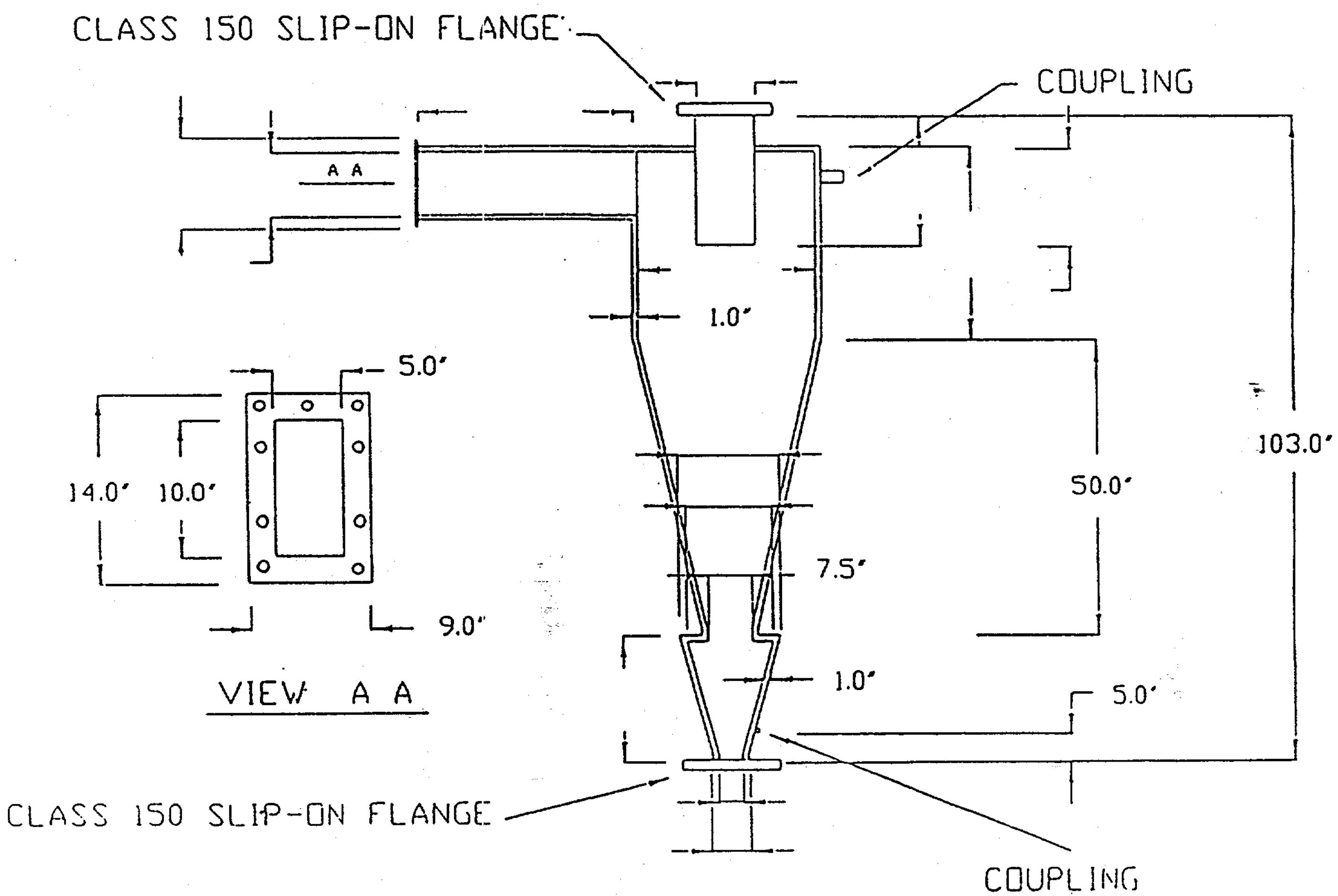


FIGURE 2-5: CYCLONE DESIGN FOR PAFBC

therefore, a refractory-lining will replace the water jacket and would still permit the use of mild steel construction.

2.1.2 Summary of Laboratory Results

The detailed results of laboratory operation on various coal feedstocks, including two waste streams from Island Creek Coal Company's Alpine Mine preparation plant and three production coals were discussed in previous reports. In this Final Report, a direct comparison of the laboratory unit performance operating in the PAFBC mode and as a conventional atmospheric bubbling fluidized bed combustor (AFBC) is presented.

The system was thoroughly tested for 600 hours of operation with a number of coal fuels including Maryland coals and both refuse and product coals from the Alpine plant of the Island Creek Coal Company. The Maryland coals employed were from sources anticipated for commercial coal supply to Baltimore Thermal. In order to establish the validity and robustness of the principal attributes of the PAFBC technology, two specific test runs were conducted to make the point clearly. The results of these tests are presented below. The coal employed was from the Island Creek Coal Alpine mine and the properties of the coal are provided in *Table 2-1*.

TABLE 2-1:

CHEMICAL COAL ANALYSIS
(Percentages By Weight)

Coal Source:	*Island Creek*
Carbon	71%
Hydrogen	3.78%
Sulfur	1.95%
Nitrogen	1.18%
Oxygen	1.57%
Ash	11.92%
Moisture	8.08%
Btu/lb	12,032

The test conditions involved operating the system as a PAFBC combustor system and another time without the pulse combustor, as an AFBC, with all the coal fed to the fluid bed. Furthermore, in the PAFBC mode of operation, a high fluidization velocity was used to establish the ability of a pulsed bed to perform well at high throughput with both the bed pulsation and local recirculation of flue gas and solids minimizing the bubble growth (with gas breakthrough) and enhancing bed performance in sulfur capture, NO_x reduction, and combustion efficiency.

A summary of the test results is provided in *Table 2-2* which is also depicted in graphic form in *Figures 2-6*, *2-7*, and *2-8*. As can be readily understood from Figure 2-6, which depicts the operating conditions, in the PAFBC mode the bed temperature was slightly lower, the bed depth was also shallower, and the fluidization velocity was significantly higher.

TABLE 2-2:
PILOT PLANT TEST RESULTS

YEAR:	1989	1989	1991	1991	1991
Operating Conditions:	AFBC Steam	PAFBC Steam	AFBC Drying	PAFBC Drying	PAFBC Steam
System Firing Rate, MMBtu/hr	1.98	1.96	1.80	2.16	3.35
Bed Temperature, oF	1550	1550	1550	1500	1530
Bed Depth, inches	32	32	16	14	24
Superficial Velocity, ft/s	7.0	7.0	16.0	17.3	16.2
Ca/S Molar Ratio	2.7	2.5	3.0	3.0	-
Cyclone Catch Recycle Ratio	0	0	1:1	1:1	0
Coal Feed Rate to Bed, lb/hr	160	100	150	112	235
Coal Feed Rate to Pulse Combustor, lb/hr	0	60	0	68	20
Performance:					
Gas Analysis at Cyclone Exit:					
SO_2, ppmv	450	100	314	113	-
SO_2, lb/MMBtu	0.90	0.20	1.31	0.42	-
NO_x, ppmv	590	250	293	146	270
NO_x, lb/MMBtu	0.80	0.40	0.88	0.39	0.25
CO, ppmv	1100	260	200	40	475
CO, lb/MMBtu	1.00	0.20	0.36	0.007	0.42
O_2, %	3.8	4.0	11.2	11.2	3.1
Sulfur Capture, %	72	93	61	88	-
Carbon Conversion Efficiency, %	90.4	95.1	96.8	96.5	93.7
System Steam Rate, lb/hr	700	800	816	1,284	2,002

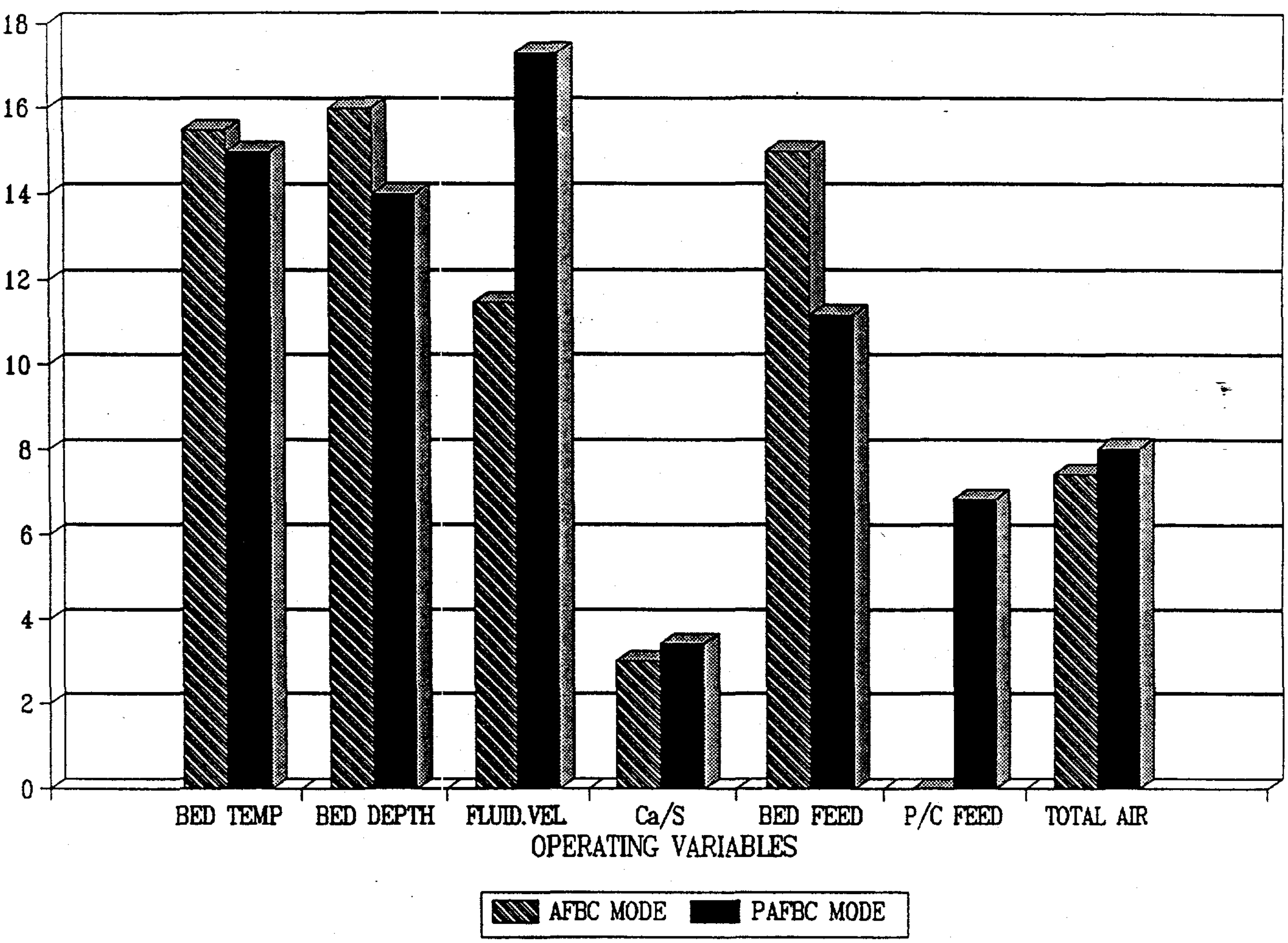


FIGURE 2-6: OPERATING CONDITIONS (AFBC & PAFBC MODES)

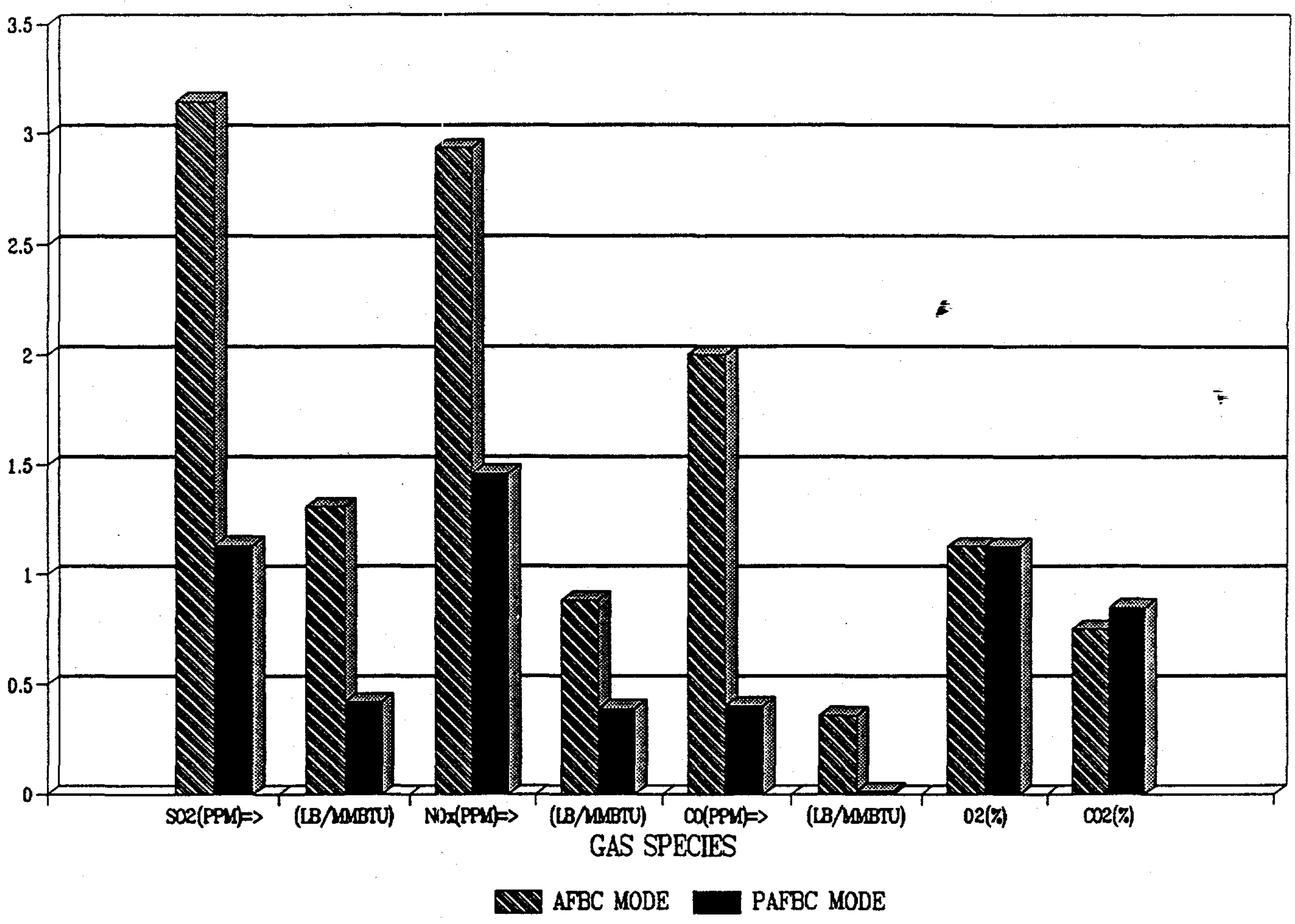


FIGURE 2-7: FLUE GAS ANALYSIS (AFBC & PAFBC MODES)

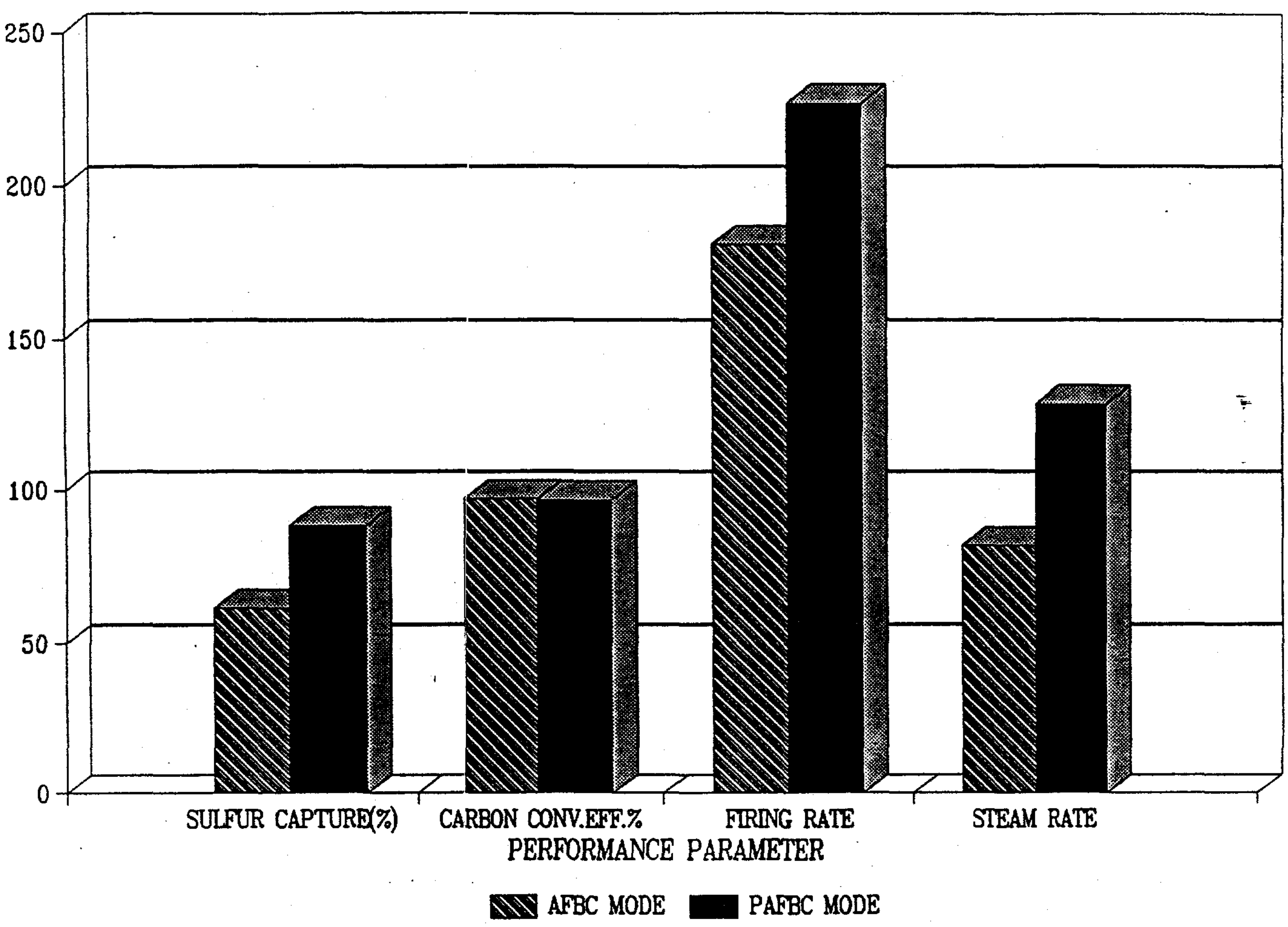


FIGURE 2-8: OTHER PERFORMANCE RESULTS (AFBC & PAFBC MODES)

Figure 2-7 also illustrates that in the PAFBC mode the Ca/S ratio was slightly higher and the total air feed was also higher. All the coal fed to the system was fed to the fluid bed in the AFBC mode without classification of the coal (i.e., as received ¼" x 0 coal), with neither coal nor air fed to the pulse combustor. Therefore, the system was operated as a conventional AFBC.

In the PAFBC mode, the coal was air classified with the fines (28 mesh x 0) being burned in the pulse combustor and the larger coal particles being burned in the fluid bed. The units for the bar chart of Figure 2-6 are as follows:

Bed Temperature	100°F
Bed Depth	Inches
Fluidization Velocity	Ft/sec
Ca/S Ratio	Molar
Bed Feed	10 lbs coal/hr
P/C Feed	10 lbs coal/hr
Total Air	100 SCFM

Thus, the total system firing rate in the AFBC mode was 1.804 MMBtu/hr and in the PAFBC mode, 2.267 MMBtu/hr. Therefore, the nominal heat release rate per unit bed area in the AFBC mode was 451,000 Btu/hr.ft^2 and in the PAFBC mode it was 566,750 Btu/hr.ft^2. Both runs were without heat removal from the bed (no steam raised in the eductor) and therefore sufficient excess air to maintain the bed temperature at the desired operating level was employed. These operating conditions are directly relevant to the proposed demonstration plant for the drying application at the Alpine Island Creek Coal plant. Nevertheless, it should be noted that with heat removal from the bed (steam raising in the eductor), there is enough air flow in the fluidization velocities employed (as evidenced by 11.22% O_2 in the flue [Table 2-2]) for almost double the combustion intensities mentioned above. This is extremely important for steam raising and specially utility applications where lower capital cost (high firing intensity), low cost of fuel, high combustion efficiency, and superior environmental performance are at a premium.

In order to discuss the environmental performance in both the AFBC and PAFBC modes, Figure 2-7 presents the gas analysis test results. The units and scale employed in Figure 2-7 present the data area as follows:

SO_2	(ppm)	100
	(lb/MMBtu)	1
NO_x	(ppm)	100
	(lb/MMBtu)	1
CO	(ppm)	100
	(lb/MMBtu)	1
O_2	(%)	10
CO_2	(%)	10

Even though the fluidization velocity in the PAFBC and total system firing rate were higher than that in the AFBC mode (with similar Ca/S ratio in both cases), the SO_2 levels, both in PPM and lb/MMBtu, were lower with O_2 concentration in the flue being the same for both modes. This demonstrates the unique features of the PAFBC process in dramatically improving sulfur capture in the system while allowing higher system throughput in a compact combustor system design; hence, lower capital cost per Btu/hr fired (i.e., there is no need to go to capital-intensive CFBs). The specific analysis of the emissions results are provided below.

It should be noted that the bed depths employed were deliberately low with the fluidization velocities high so as to have nominally low gas residence time in the sorbent bed. This is to explicitly test the effect of the PAFBC process as compared to a strictly AFBC combustor. Thus, in the AFBC mode the gas residence time in the entire combustor at operating conditions was 1.9 seconds and for the PAFBC mode it was 1.7 seconds. With the above in mind, the ppm sulfur oxide level was reduced in the flue gas to 36 percent of the emission found in the AFBC mode with the O_2 concentration in the flue maintained the same but with an increase in system firing rate in the PAFBC modes to 126 percent of that for the AFBC mode. The fluid bed heights were such that the PAFBC bed height was even slightly lower than in the AFBC. The calcium-to-sulfur molar ratio in the overall system feed was similar but slightly higher in the PAFBC mode.

Detailed engineering and process evaluation of the performance data provided a number of important findings and process considerations. First, the design and performance of the AFBC part of this hybrid combustion system is robust and is of good performance. This is reflected in the healthy amount of sulfur capture (61%) in the AFBC mode in spite of the shallow bed height (16" as opposed to 36" to 45" normal height) with relatively high fluidization velocity (11.5 ft/sec). We should also note that the carbon conversion efficiency was quite respectable (96.8%) for the operating conditions which confirms that the solid separator is performing efficiently in returning solids to the bed at a pressure drop investment of merely 2 inches of water. The cyclone catch reinjection is also believed to make a modest contribution in carbon burnout of char, found in the fine ash which escapes the solids separator but becomes captured by the cyclone. We should also note that the cyclone catch included 6 percent material of less than 1½ micron in size with only 3½ inches of water in cyclone pressure drop. This steam-raising cyclone was designed and built in-house by MTCI. Test conditions in the AFBC were made such that the SO_2 concentration in the flue was the same as that achieved for the PAFBC test. The CO emissions of 200 ppm in the AFBC mode were only viewed as adequate considering that unclassified coal, containing fines, was fed overbed. This result, however, suggests that the freeboard temperature was sufficient together with the amount of excess air used to maintain the bed temperature (without heat removal from the bed), prevented CO levels to become more excessive.

In the PAFBC mode of operation, dramatic reductions in SO_2, NO_x and CO emissions were obtained at the same O_2 partial pressure in the flue gas even though the bed of sorbent was slightly shallower (14" and the effective fluidization velocity was significantly higher (50% higher). The fluidization velocity in the PAFBC mode is calculated taking into account the gas flow emanating from the pulse combustor, the educted flue gas from the splash zone above the bed being recirculated back through the sorbent bed, the amount of fluidization air from the air distributor at the bottom of the bed and the smaller area around the eductor where all such gases must travel upwards in the bed. The flow inside the eductor, in the PAFBC operating mode, is downwards and is comprised of a mixture of the pulse jet flow, the educted flow gas from the splash zone, and the entrained solids (preferentially the more entrainable fines) educted with the splash zone flue.

Clearly, in the PAFBC mode of operation, there is an advantage to having only the larger particle sizes fed overbed into the fluid-bed combustor. All the fines are burned in the pulse combustor *upstream* of the sorbent bed. The fines coal particles, which would have elutriated faster in the fluid bed or even elutriated in the freeboard if fed overbed (before they reach the sorbent bed), are completely combusted in the pulse combustor. Fines that are burned in the pulse combustor that are at the larger end of the particle size distribution of the material (fed into the pulse combustor) would at least be devolatilized with the volatiles burned in the pulse combustor and the remaining char ignited before exiting the resonance tube into the eductor. Therefore, all the SO_2 generated from the combustion of such fines upstream of the fluid bed is passed through the sorbent bed through the down flow in the eductor. Should there be CO in the resonance tube exit flow, this also is mixed with the educted flue gas and passed through the fluid bed. It should also be noted that the hot down flow from the lower exit of the eductor is met by an upward flow of oxygen-rich gas from the distributor below the eductor. The result is that the mixture moves laterally outward towards the annular region around the eductor and then upward through the fluid bed towards the freeboard.

Eductor flow calculations suggest that the amount of flue educted from the splash zone back into the eductor is approximately 50 percent of that leaving the pulse combustor. This can be varied by design through tailpipe and eductor design modifications. In the present pilot plant system, the resonance tube is designed with a slight divergent diffuser at the bottom end of the tailpipe to slow down the jet flow from the pulse combustor and the design of the eductor was accordingly selected.

In order to complete our discussion of the system performance evaluation, we now refer to Figure 2-8 which depicts other performance results. The scale and units employed in Figure 2-8 are as follows:

Performance Parameter	Scale/Units
Sulfur Capture	1%
Carbon Conversion Efficiency	1%

Firing Rate	100,000 Btu/hr
Steam Rate	10 lbs/hr

In this figure, the percent sulfur capture in the AFBC modes was understandably lower (61%) due to the shallow sorbent bed height and the relatively high fluidization velocity. Furthermore, as presented earlier, the overbed feed in the AFBC mode was with the "as-received" unclassified coal, thus some of the fines did burn in the freeboard, downstream of the sorbent bed, giving rise to more sulfur oxides in the flue leaving the system. In the PAFBC shallow bed mode, even with the high fluidization velocity, the sulfur capture was nearly 90 percent and the emissions were at 0.42 lbs/MMBtu. This is 35 percent of the allowable emissions in the EPA regulations (40 CFR Ch. 1 [7-1-90 edition]) for coal-fired steam boilers for electric utility applications. This great performance, particularly in view of the Clean Air Act, is achieved with a Ca/S molar ratio of 3.4:1 which is not excessively high, especially in view of the shallow bed used.

We also note from Figure 2-8 that the dramatic improvement in environmental performance was achieved with significant increase in plant firing rate and steam output. This is also achieved at essentially the same carbon conversion efficiency and O_2 in the flue gas.

This performance is possibly due to the effect of the pulsation on the fluidization of the sorbent bed in improving mixing and reducing the extent of bubble growth even with 50 percent increase in the effective fluidization velocity. Inhibiting bubble growth enhances gas solids contact thus improving both sulfur capture and combustion efficiency with low NO_x production at the bed operating temperatures.

2.2 CLEMSON DEMONSTRATION PLANT

A Pulse Enhanced Atmospheric Fluidized Bed Combustor (PAFBC) designed to produce approximately 50,000 lbs/hr of steam was designed, constructed and tested at Clemson University. The boiler consists of a bubbling fluid-bed combustor with in-bed heat transfer modules, and a water-jacketed pulse combustor. Although the main fuel for both combustors is coal, it is possible to fire both the bed and the pulse combustor on natural gas. One of the main benefits of this design, should a problem with coal feeding arise, is that he boiler can be run on gas until the problem is alleviated. Under normal operations the coarse coal is burned in the bubbling fluid bed, while the coal fines are burned in the pulse combustor. This practice lowers SO_2 emissions by minimizing the amount of coal fines being burned in the freeboard of the bubbling fluid bed. SO_2 emissions are minimized in the fluid bed by the use of limestone to capture the sulfur in the coal.

The plant is controlled by a PLC which is interfaced through a personal computer. As far as operation of the plant is concerned, most of the plant controls are fully controlled through the PLC. However, both pilots, as well as the coal processing required to load the silos are controlled manually without interaction with the control computer.

2.2.1 Design Parameters

Table 2-3 outlines overall general design data and system specifications for the PAFBC System. Also included in Table 2-3 is the plant efficiency calculations based on the design specifications.

Figures 2-9 through *2-11* represent the entire system Process Flow Diagrams. The flow diagrams show all input, intermediate and output resources for the system including fuels (natural gas and coal), water, steam, air and ash.

TABLE 2-3:

PAFBC SYSTEM SPECIFICATIONS - GENERAL DATA

Total coal flow lb/hr	6,000
Coal composition, Wt.%	
N_2	1.09
O_2	2.02
C	66.35
S	3.13
H_2	3.6
H_2O	3.69
Ash	20.12
Excess air, %	15
Limestone flow, lb/hr	1,757.8
Inerts, lb/hr	195.3
Total limestone flow, lb/hr	1,953.1
Calcium/sulfur ratio	3
Total air flow, lb/hr	62,476.2
Cooling medium	water
Water inlet temperature, °F	170
Water inlet pressure, psia	190
Steam condition at outlet	saturated
Steam flow at outlet, lb/hr	52,200
Circulation ratio	10
Circulation rate, GPM	1,025
Steam outlet temperature, °F	377.53
Steam outlet pressure, psia	190
Flue gas flow at outlet, lb/hr	66,624
Flue gas outlet temperature, °F	350
Ash outlet flow, lb/hr	3,005

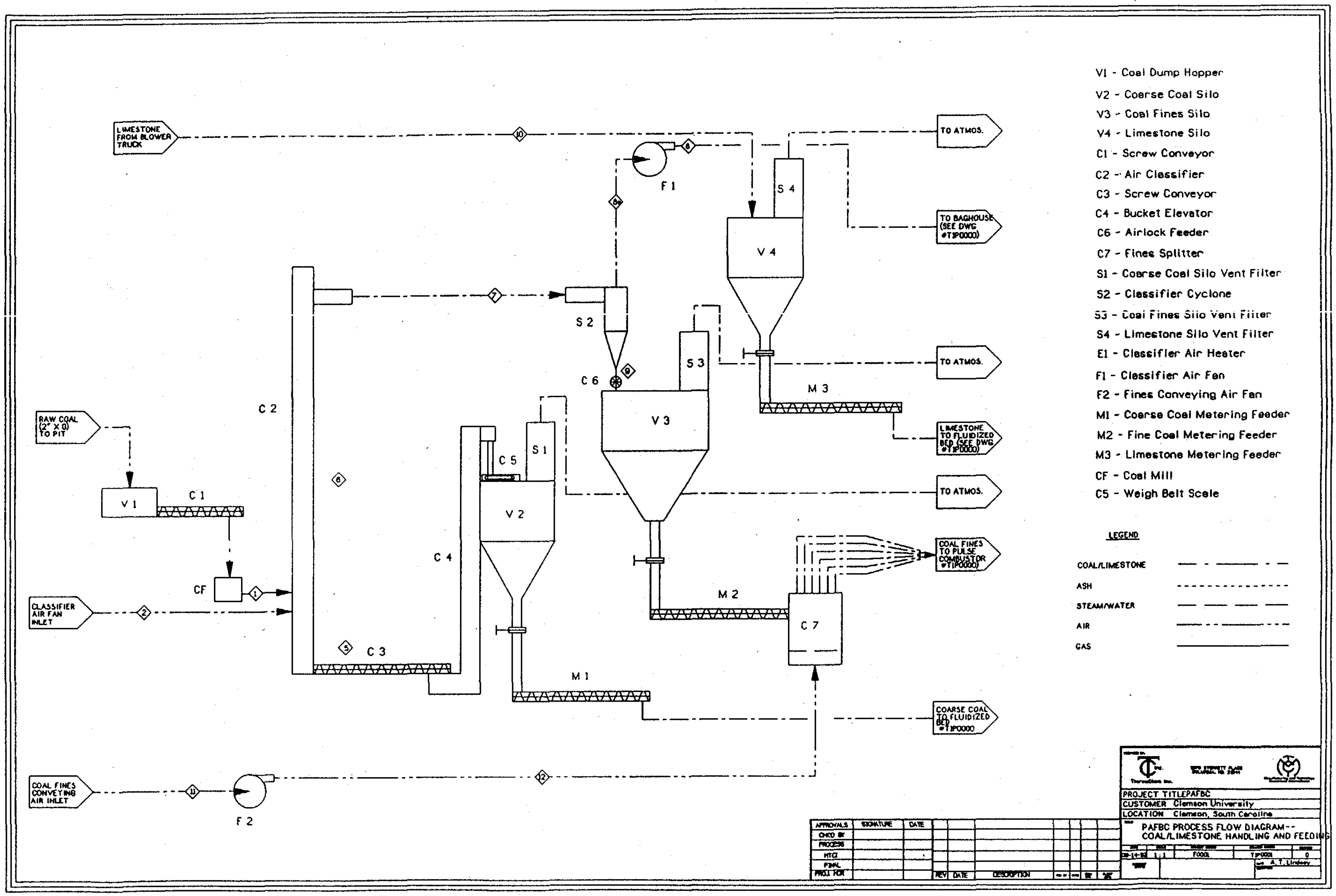


FIGURE 2-9: PROCESS FLOW DIAGRAM

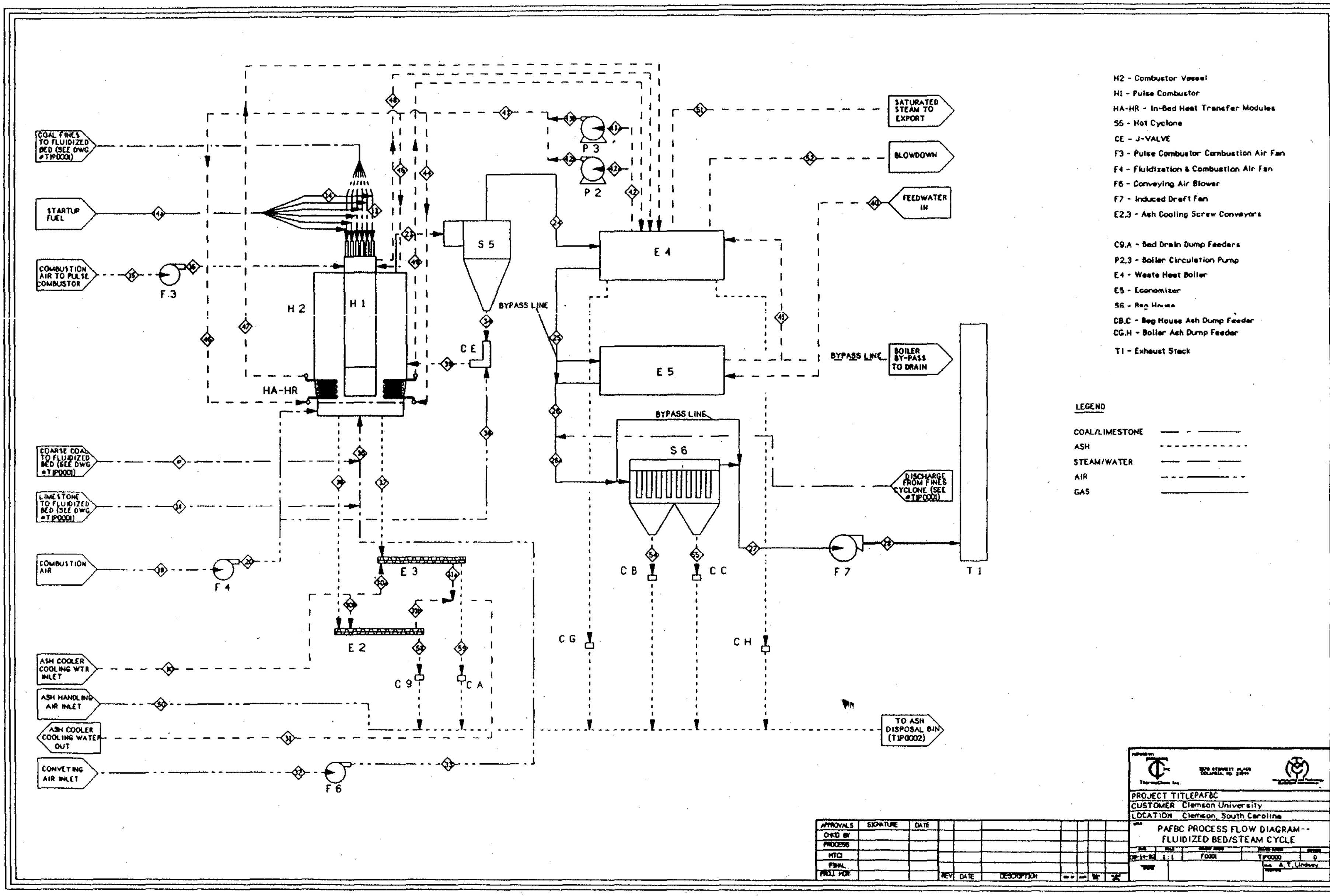


FIGURE 2-10: PROCESS FLOW DIAGRAM

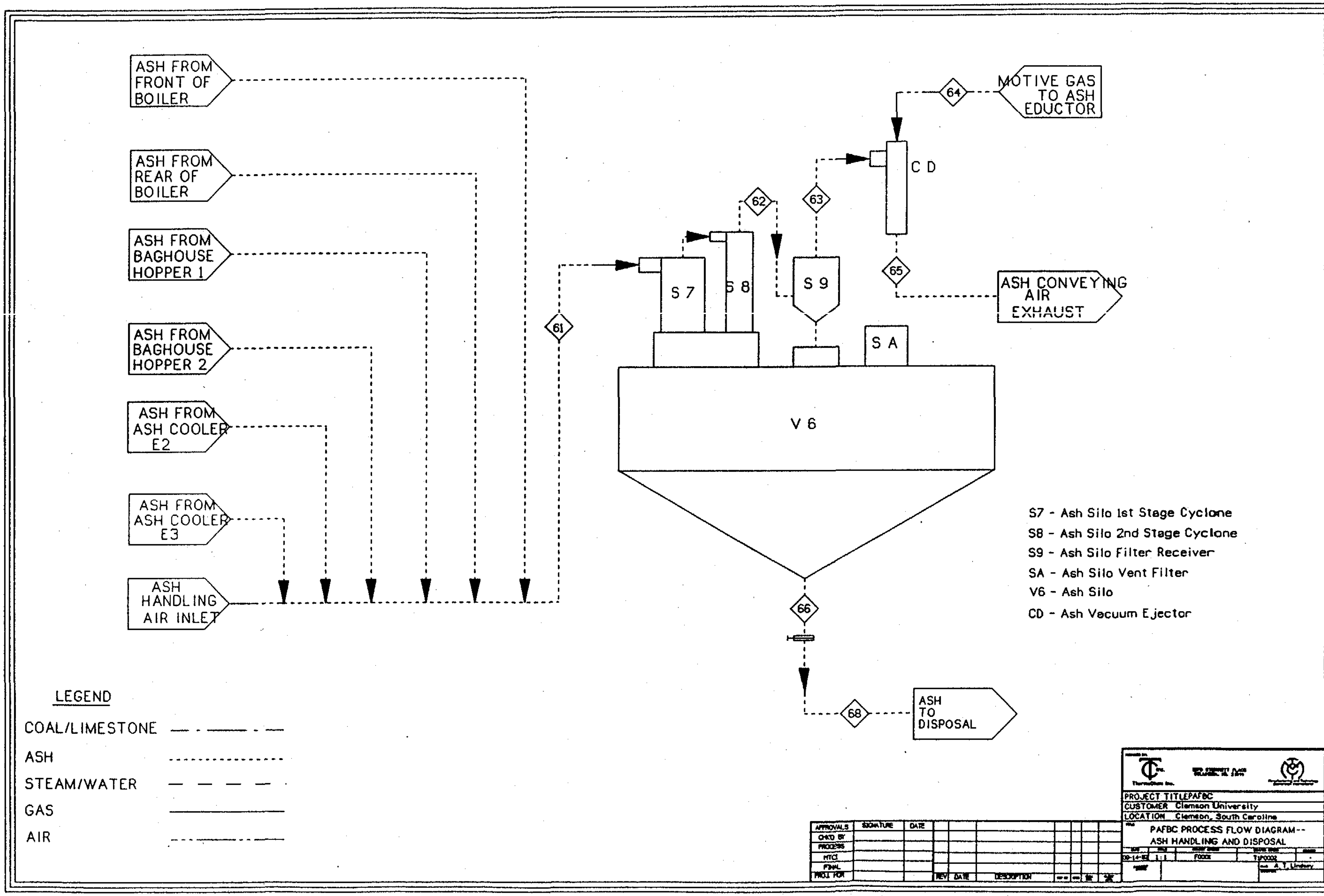


FIGURE 2-11: PROCESS FLOW DIAGRAM

System installation began on May, 1993. *Figure 2-12* is the overall system installation plan and *Figure 2-13* shows the site plan. *Figure 2-14* shows the general assembly for the actual PAFBC vessel. *Figures 2-15* through *2-21* depict six different elevations and views of the unit and structure. *Figure 2-22* is a photograph of the facility after 90 percent completion. *Figure 2-23* shows the structural steel and *Figure 2-24* is the steam delivery lime going to the existing boiler house.

2.2.2 Coal/Limestone Preparation and Feed System

The PAFBC has been designed to burn coal in two locations. The overall size of the coal is ½" by zero with 25 percent by weight in the 30 mesh by zero category. The fines (30 mesh by zero) will be burned in the pulse combustor while the coarse particles (½" by 30 mesh) will be burned in the fluidized bed. The limestone has to be prepared to a single size range which is 11 mesh by zero. The appropriate design specifications for the coal and limestone mills screening equipment and storage bins are presented in *Appendix 1*. For preliminary tests conducted on the PAFBC, coal and limestone was procured already prepared to the correct sizes and the coal was air classified to separate the fines (30 mesh by zero) from the coarse coal.

The coal and limestone will be mixed in the correct proportion using an appropriate arrangement consisting of a weigh feeder and metering auger and the mixture will be air conveyed to bed and fed underbed through a four feed ports located directly below the pulse combustor exit in the bed. The fines used as fuel for the pulse combustor will be air conveyed to the distributor that will split the feed into three separate streams each feeding one of the ports in the pulse combustor combustion chamber. The fine coal will be fed coaxially into the throat of each of the three aerovalves of the pulse combustor.

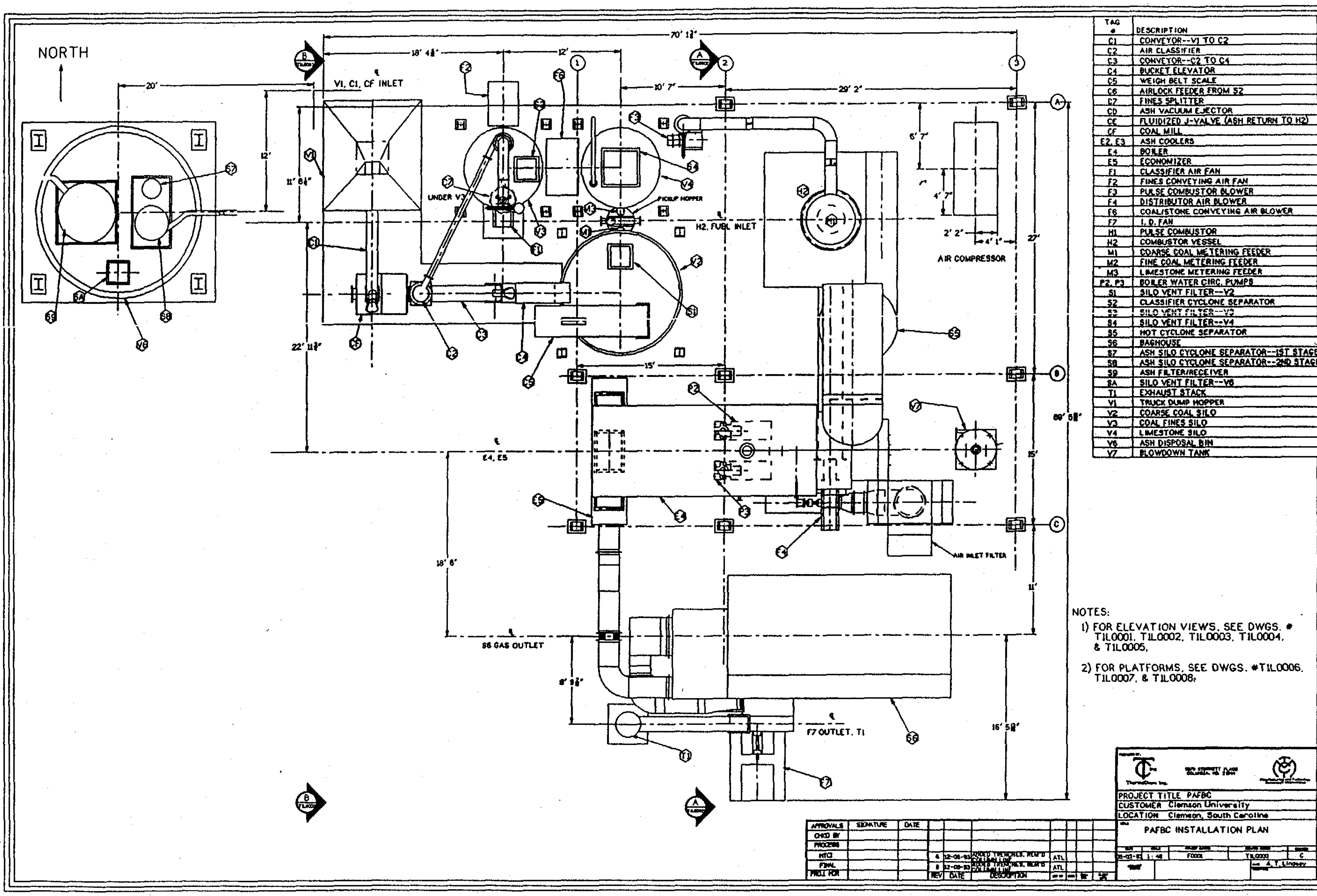


FIGURE 2-12: PAFBC INSTALLATION PLAN

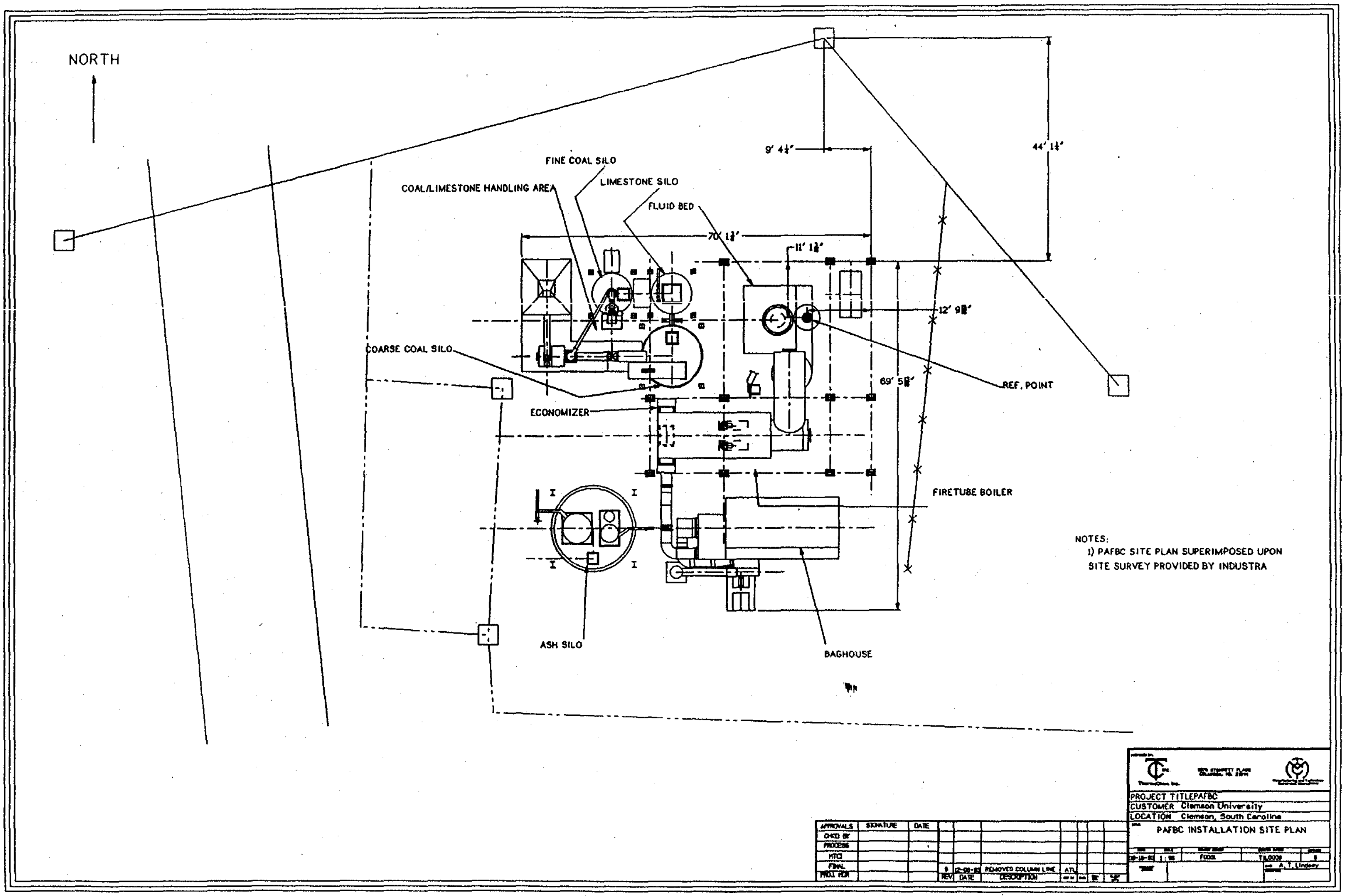
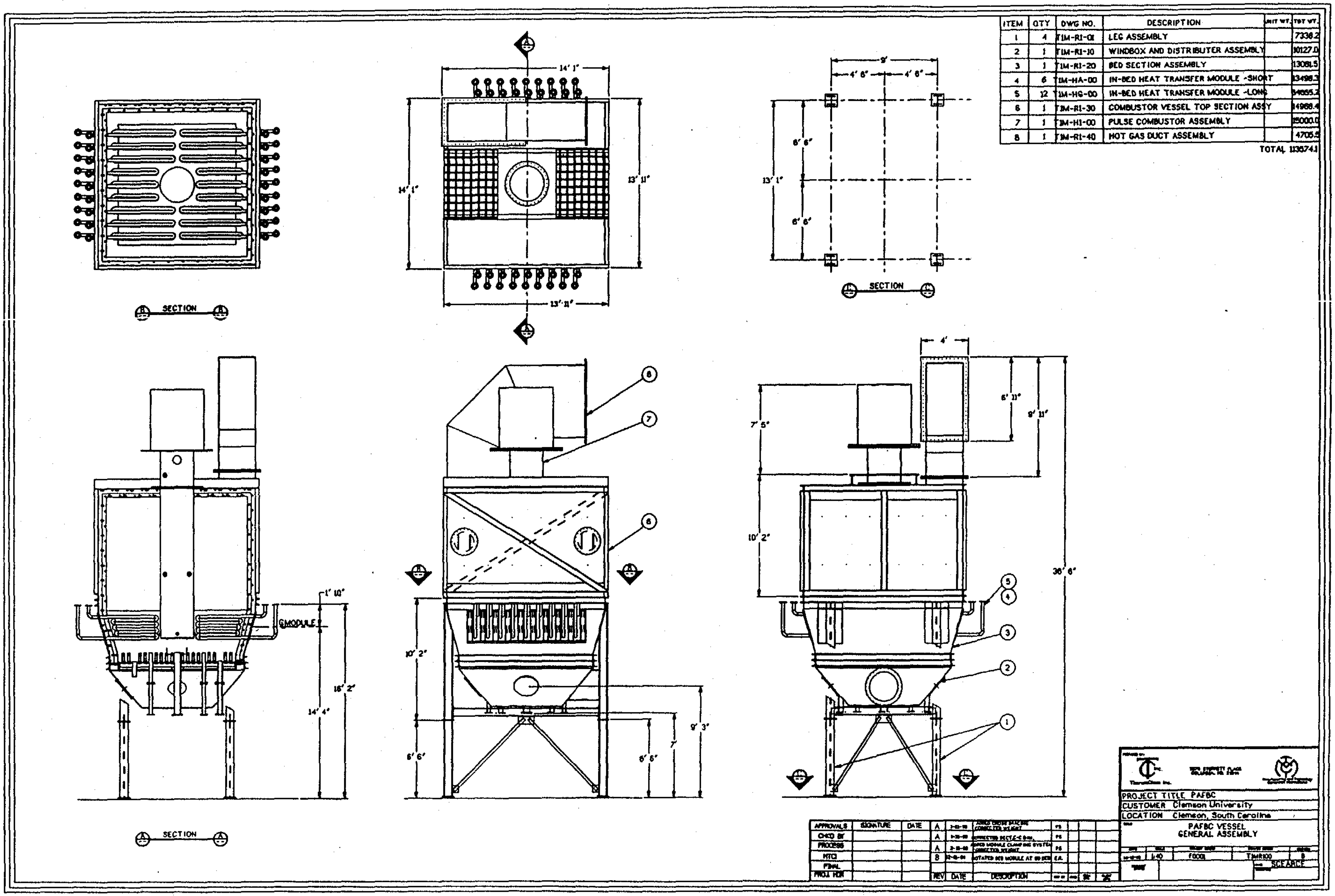


FIGURE 2-13: PAFBC INSTALLATION SITE PLAN



ITEM	QTY	DWG NO.	DESCRIPTION	UNIT WT.	TOT WT.
1	4	TIM-RI-01	LEG ASSEMBLY		7338.2
2	1	TIM-RI-10	WINDBOX AND DISTRIBUTER ASSEMBLY		10127.0
3	1	TIM-RI-20	BED SECTION ASSEMBLY		13081.5
4	6	TIM-HA-00	IN-BED HEAT TRANSFER MODULE -SHORT		13498.3
5	12	TIM-HG-00	IN-BED HEAT TRANSFER MODULE -LONG		34655.2
6	1	TIM-RI-30	COMBUSTOR VESSEL TOP SECTION ASSY		14968.4
7	1	TIM-H1-00	PULSE COMBUSTOR ASSEMBLY		15000.0
8	1	TIM-RI-40	HOT GAS DUCT ASSEMBLY		4705.5
				TOTAL	113574.1

FIGURE 2-14: PAFBC VESSEL - GENERAL ASSEMBLY

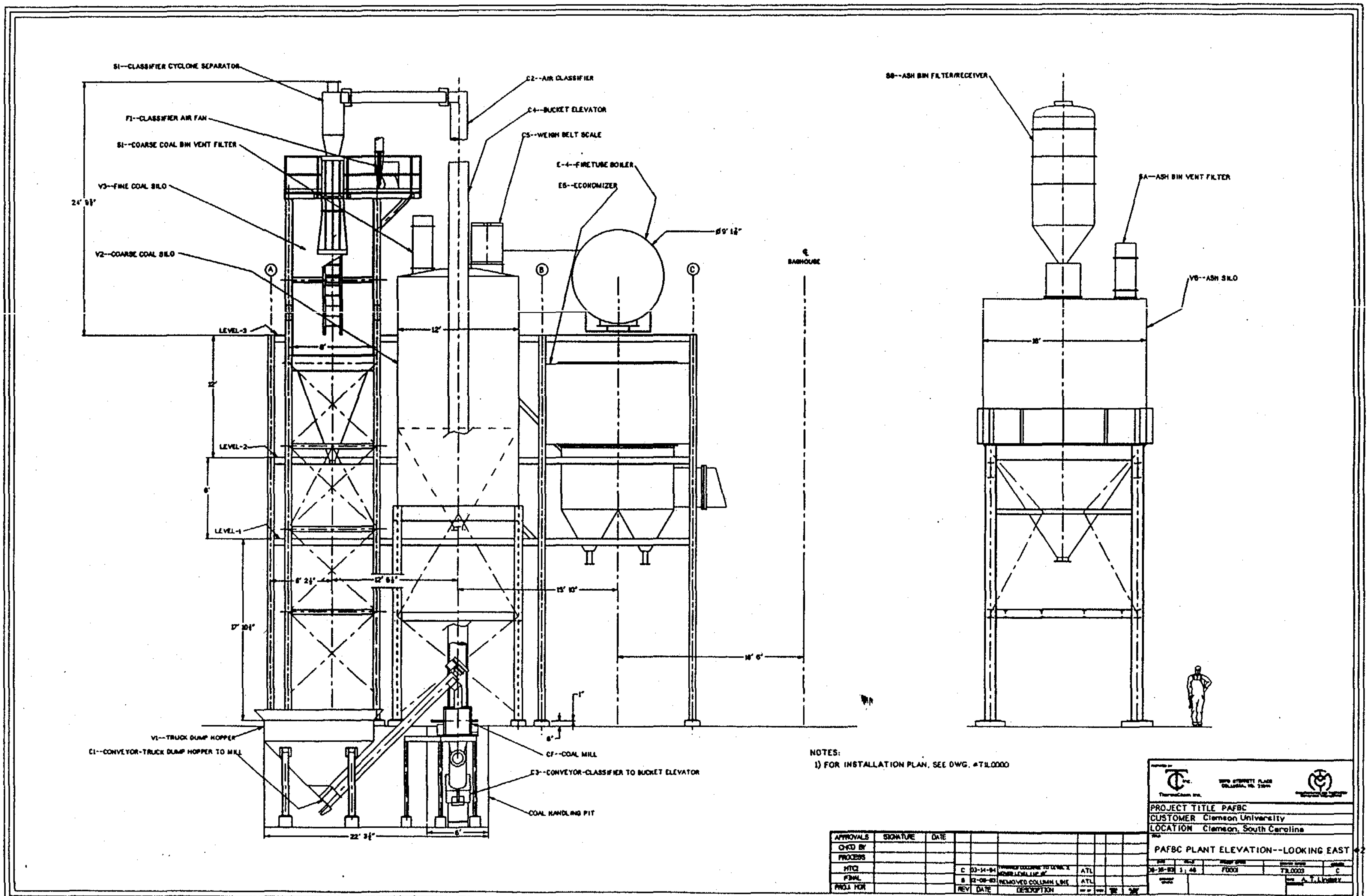


FIGURE 2-15: PAFBC PLANT ELEVATION - LOOKING EAST #2

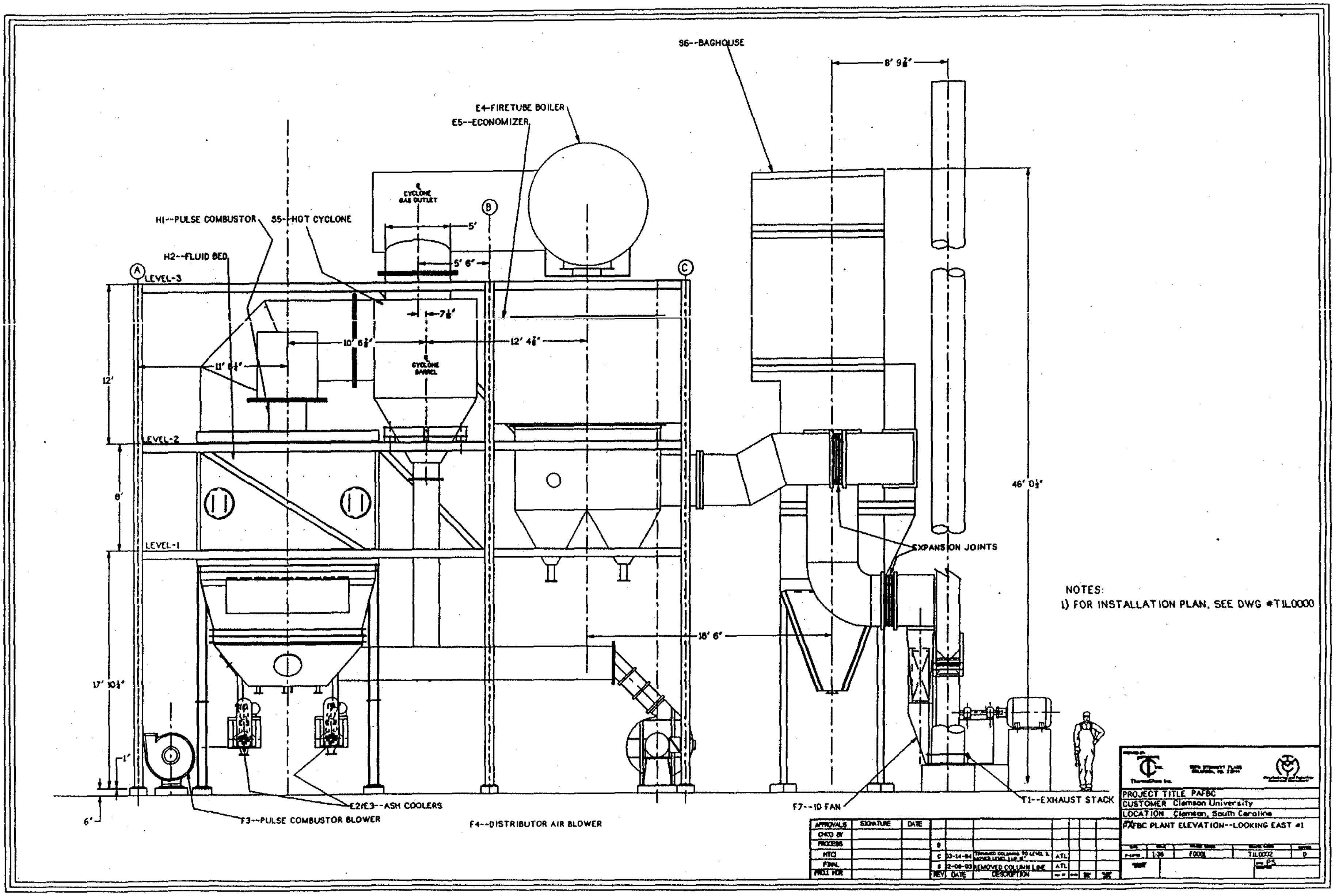


FIGURE 2-16: PAFBC PLANT ELEVATION - LOOKING EAST #1

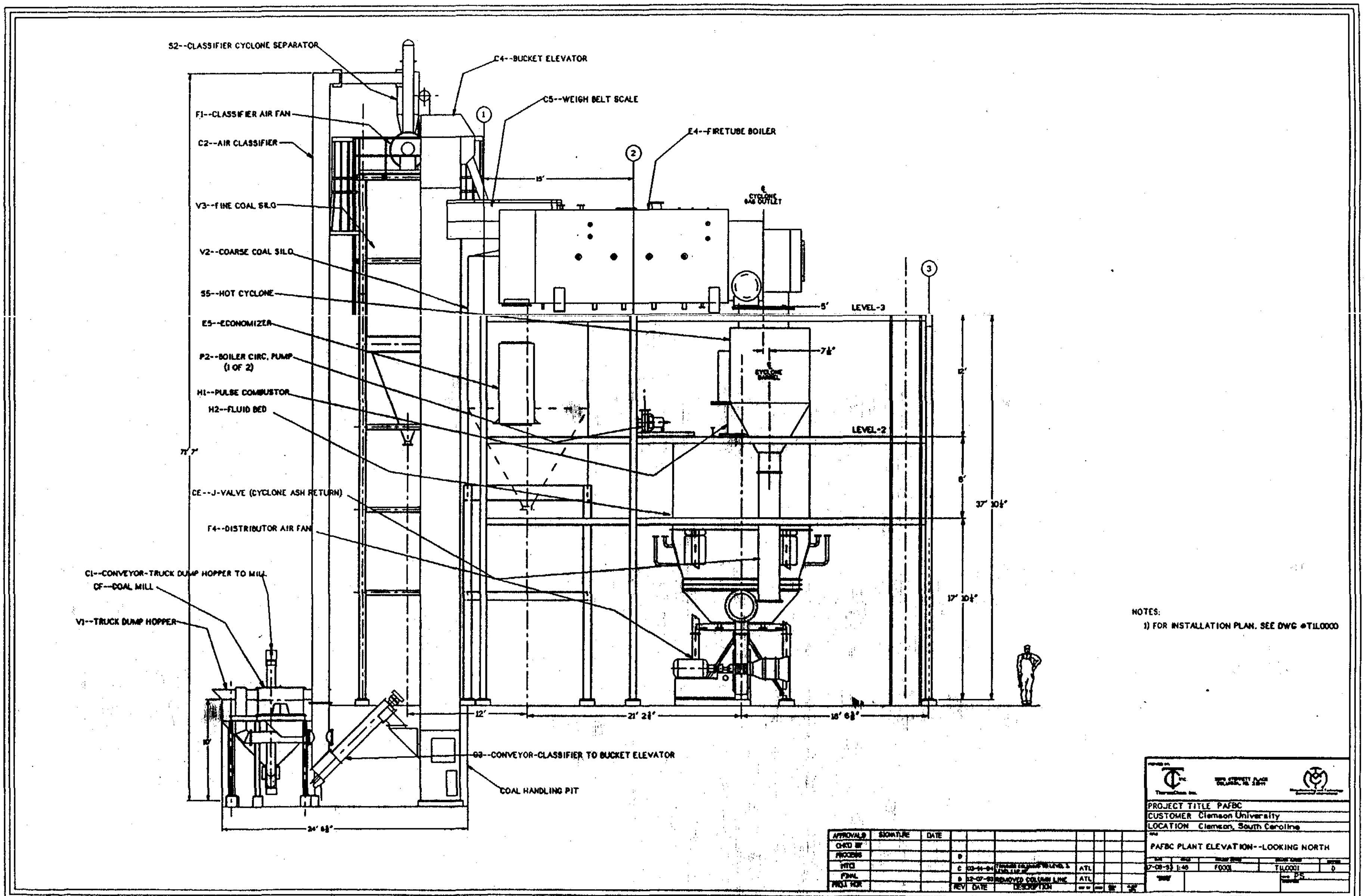


FIGURE 2-17: PAFBC PLANT ELEVATION - LOOKING NORTH

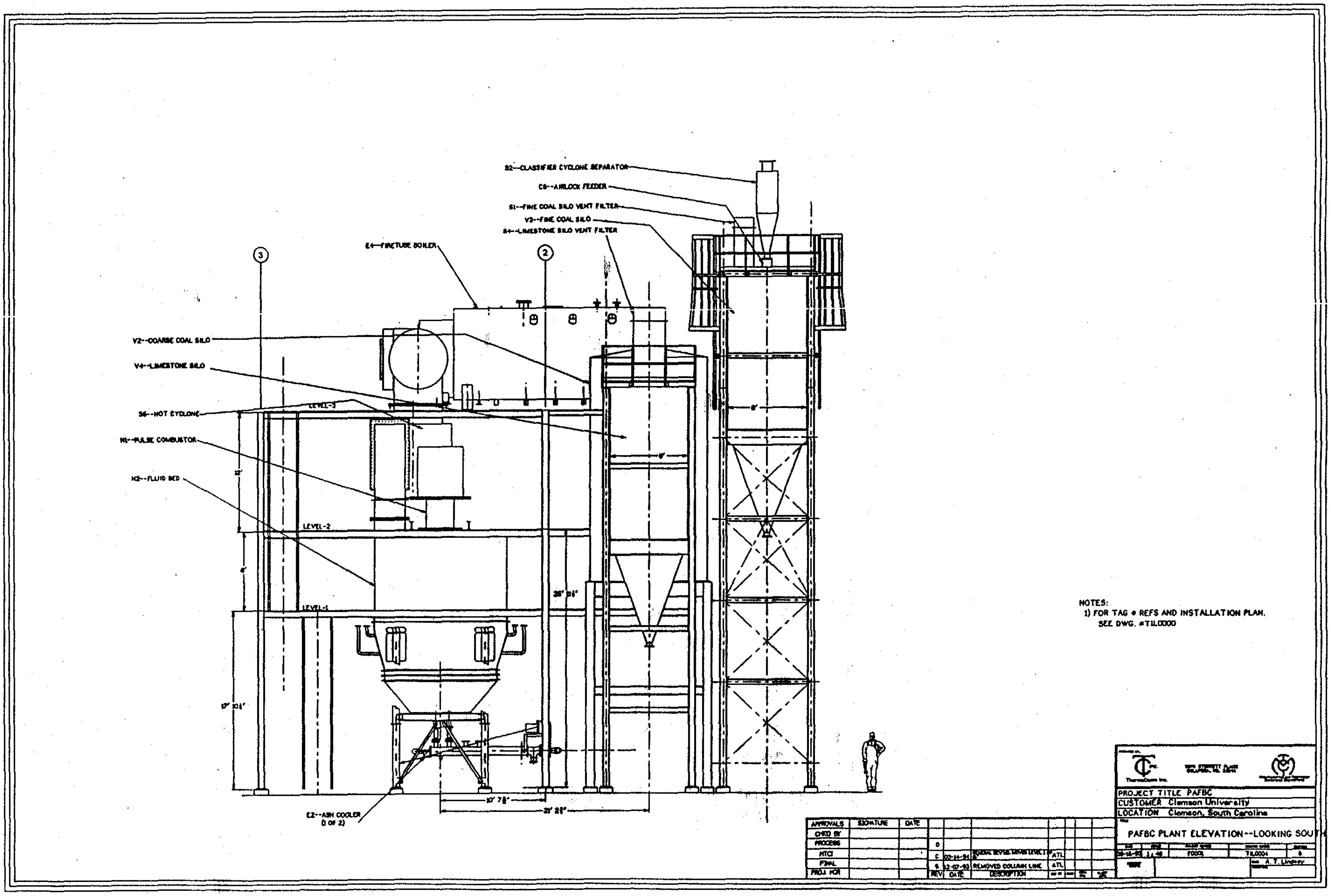


FIGURE 2-18: PAFBC PLANT ELEVATION - LOOKING SOUTH

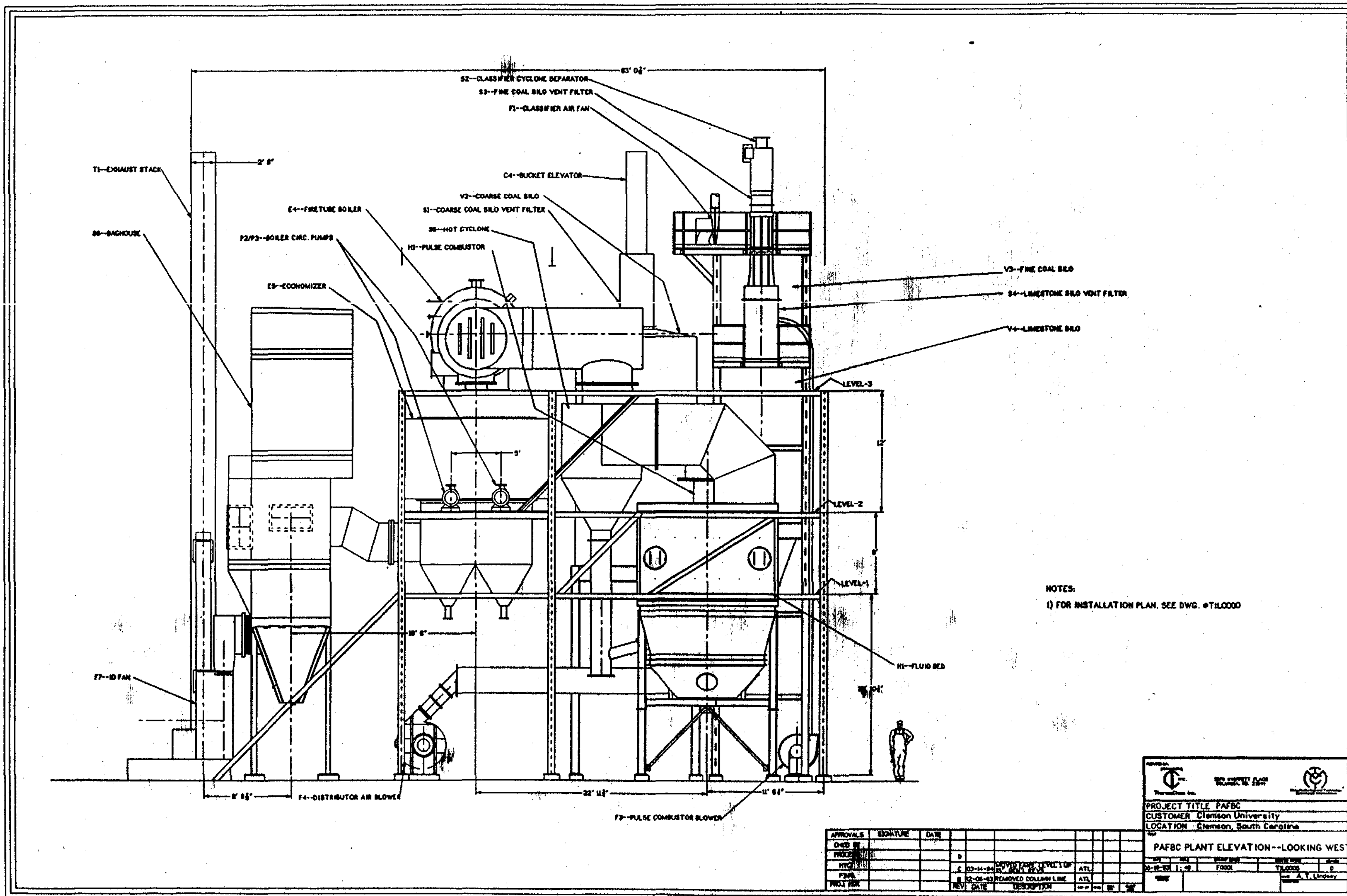


FIGURE 2-19: PAFBC PLANT ELEVATION - LOOKING WEST

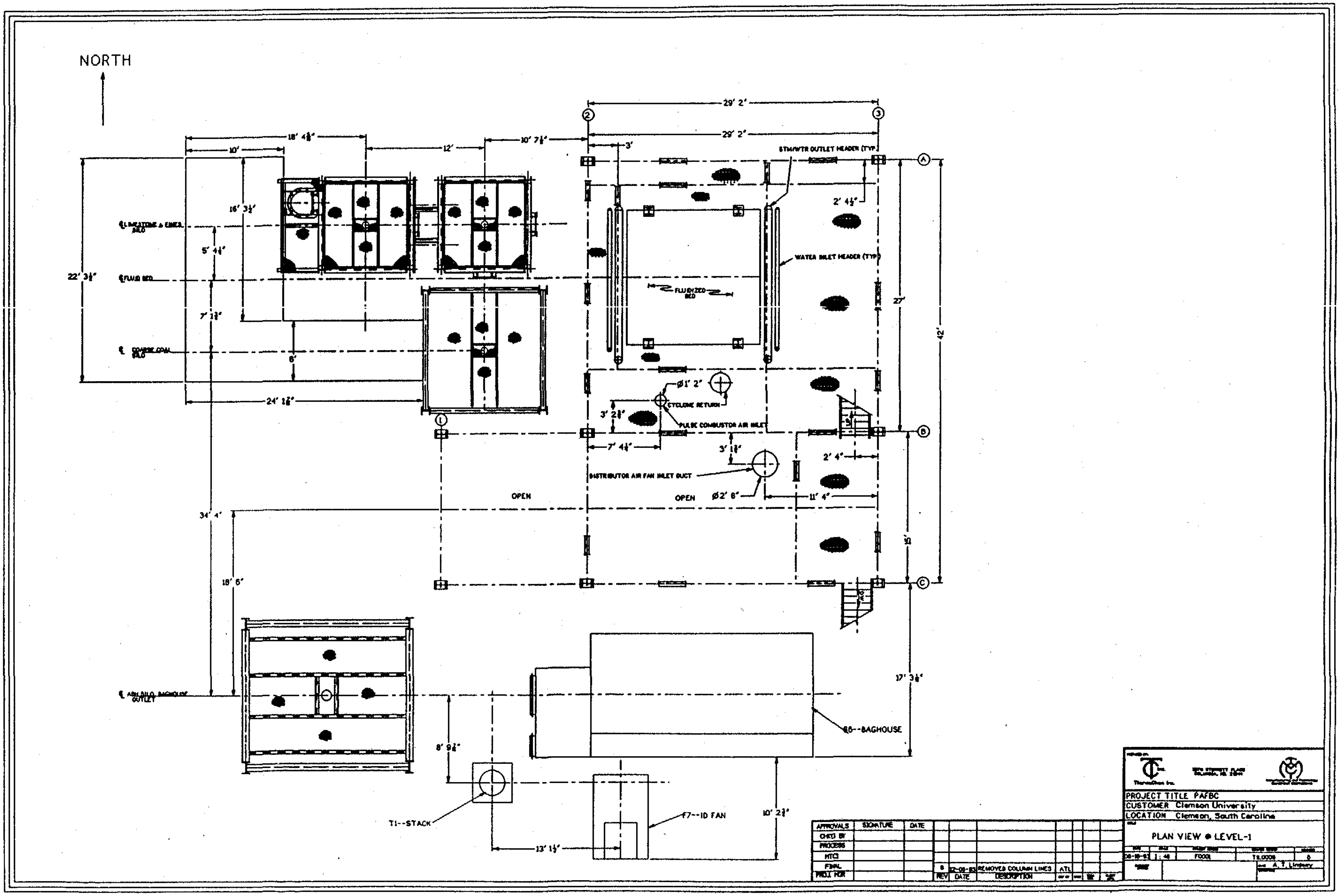


FIGURE 2-20: PLAN VIEW - LEVEL 1

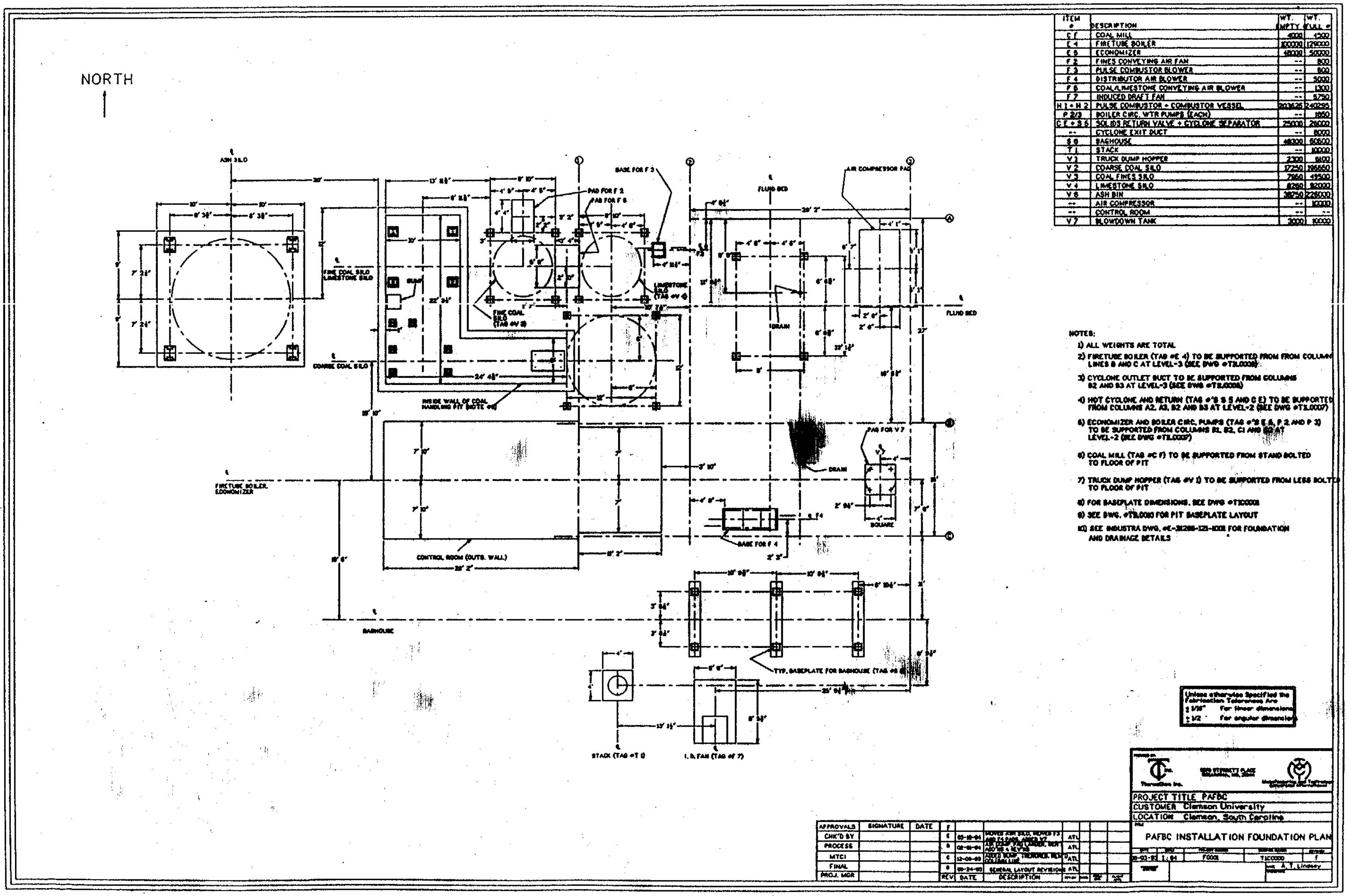


FIGURE 2-21: PAFBC INSTALLATION FOUNDATION PLAN

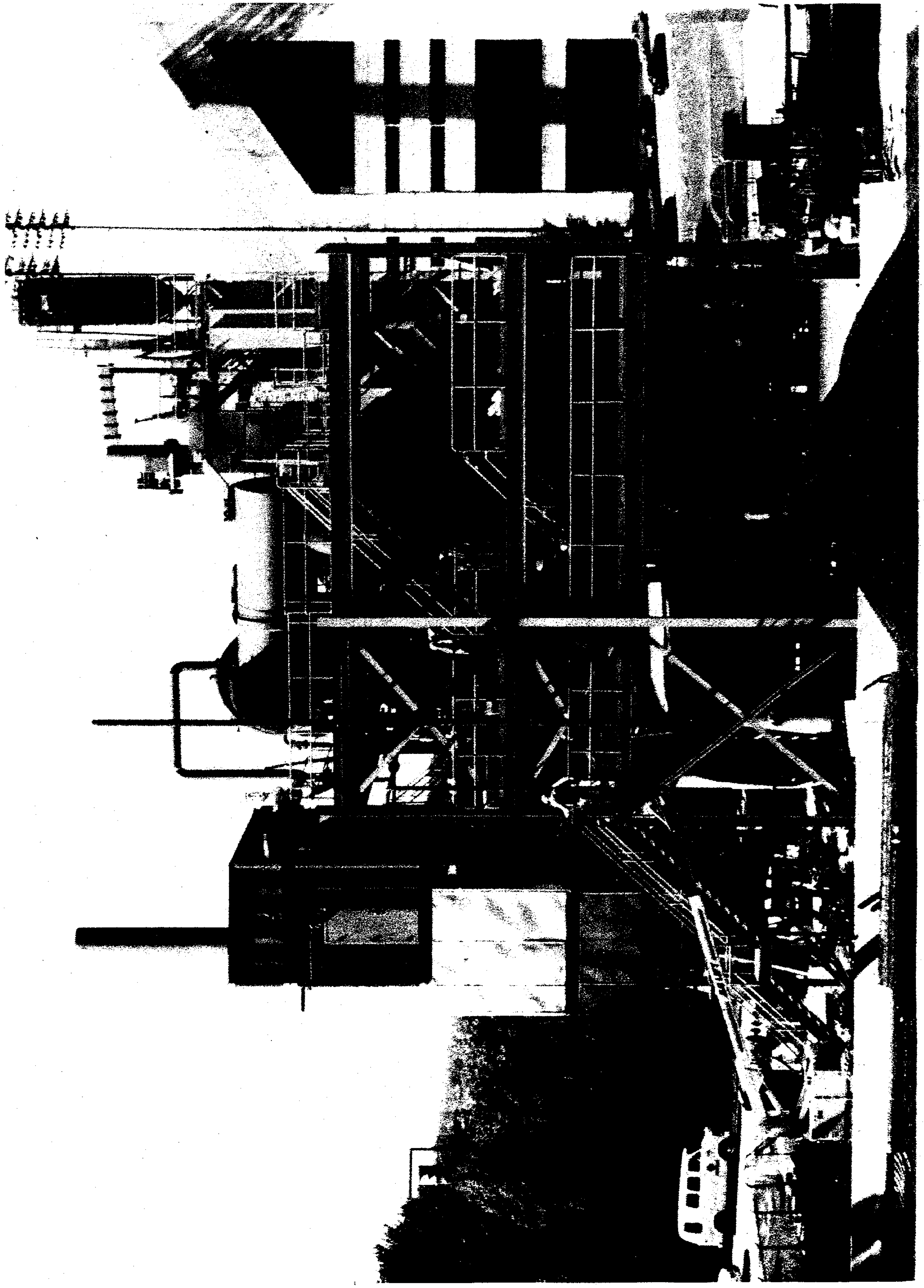


FIGURE 2-22: PAFBC DEMONSTRATION UNIT AT CLEMSON UNIVERSITY, SOUTH CAROLINA

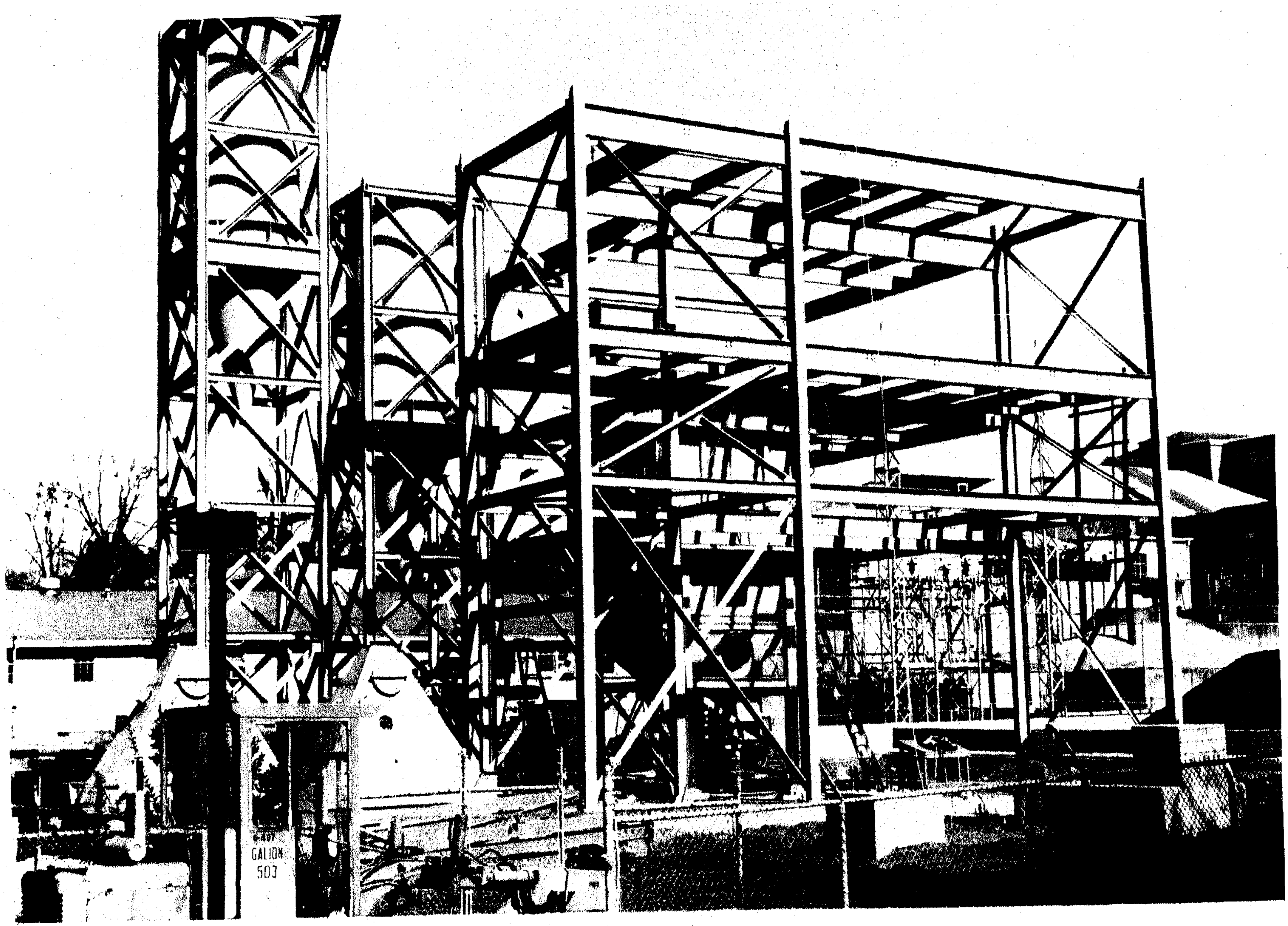


FIGURE 2-23: CLEMSON PAFBC UNIT STRUCTURAL STEEL

FIGURE 2-24: CLEMSON PAFBC UNIT - STEAM LINE CONNECTING TO UNIVERSITY STEAM SYSTEM

Coal Handling

The dump hopper is enclosed in a shack to help keep the coal as dry as possible when processing. When coal is begin processed, scoops of raw coal are loaded in the Coal Dump Hopper (V-1) (*Figure 2-25*). As the coal is dumped into the hopper, it passes through a grating, designed to minimize the number of foreign objects from entering the hopper, to the Coal Mill Screw Conveyer (C-1). To facilitate the draining of the hopper into the Hopper to Coal Mill Screw Conveyor (C-1), the Coal Dump Hopper (V-1) has several pneumatic vibrators and air jet ports in order to facilitate the draining of the hopper into the Hopper to Coal Mill Screw Conveyor (C-1). When the coal exits the Hopper to Coal Mill Screw Conveyor (C-1), it flows through a chute, over a powerful magnet, and into the Coal Crusher (C-F). This magnet is powerful enough to capture large pieces of metal that pass by the magnet up to eight inches away. As the coal leaves the coal crusher, it drops down into a small chute and is conveyed by a screw conveyor to the Air Classifier (C-2). The Air Classifier (C-2) separates the coal into fine and coarse. The amount of fine coal taken up to the fines-handling system is controlled by a manual damper on the Classifier I.D. Fan.

Coarse Coal Handling

Coarse coal exiting the bottom of the Air Classifier (C-2) falls onto a screw conveyor, and is then conveyed to the Bucket elevator (C-4). The coarse coal then travels to the top of the silo, and is deposited on a Weigh Belt Feeder (C-5). The Weigh Belt Feeder (C-5) displays the current tons/hr flow rate of coarse coal into the coal silo as well as how much coal has been fed so far. The Weigh Belt Feeder (C-5) totalizer has been mounted on the wall in the control room. After the flow rate is measured by the Weigh Belt Feeder (C-5), the coal falls into the Coarse Coal Silo (V-2). In order to prevent coal dust from leaking into the atmosphere, the Coarse Coal Silo (V-2) has a Bin Vent Filter (S-1).

During the bed feeding process, coarse coal is conveyed from the Coarse Coal Silo (V-2) through the Coarse Coal Metering Feeder (M-1), and into the Coal/Limestone Mixing Hopper. The Coal/Limestone mixture is then pneumatically conveyed from the Coal/Limestone Mixing Hopper into the Fluid Bed (R-1).

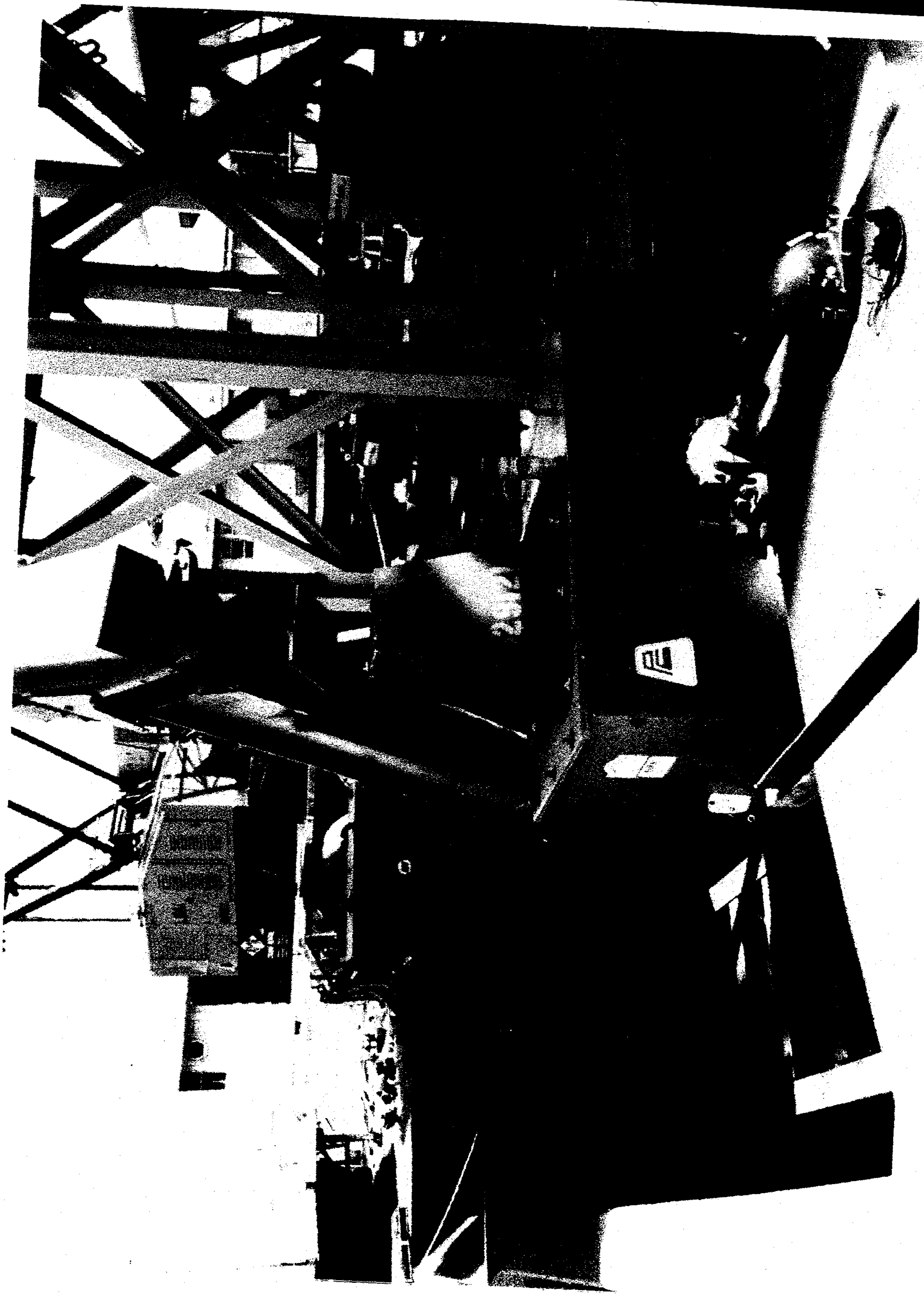


FIGURE 2-25: CLEMSON PAFBC UNIT - COAL-HANDLING HOPPE

Fine Coal Handling

The fine coal exits the top of the Air Classifier (C-2) and enters the Fine Coal Cyclone (S-2). The solids discharge of the cyclone exits directly into the Fine Coal Silo (V-3). The air stream with the remaining fine coal is sucked through the Fine Coal Baghouse (S-3). The fine coal dust then drops off the filter bags and into the silo. The air flow continues on its path through the Classifier I.D. Fan and is then vented to the atmosphere.

When running the pulse combustor on coal, the fine coal exits the bottom of the Fine Coal Silo (V-3) and is conveyed through the Fine Coal Metering Feeder (M-2) and onto the fine coal eductors. The fine coal is then conveyed pneumatically into the Pulse Combustor Coal through the inner annulus of the fuel injectors.

Limestone Handling

Limestone is injected into the Limestone Hopper (V-4) via a pneumatic limestone truck. The limestone enters the Limestone Hopper (V-4) and is kept inside the Silo by the Limestone Bin Vent Filter (S-4).

During operation, the limestone exits the bottom of the Limestone Silo (V-4) and is conveyed through the Limestone Metering Feeder (M-3) into the Coal/Limestone mixing hopper. After the limestone is mixed with the coal, the coal/limestone mixture is conveyed pneumatically into the Fluidized Bed Combustor (R-1).

Figures 2-26 through 2-32 depict the coal and limestone receiving and handling systems.

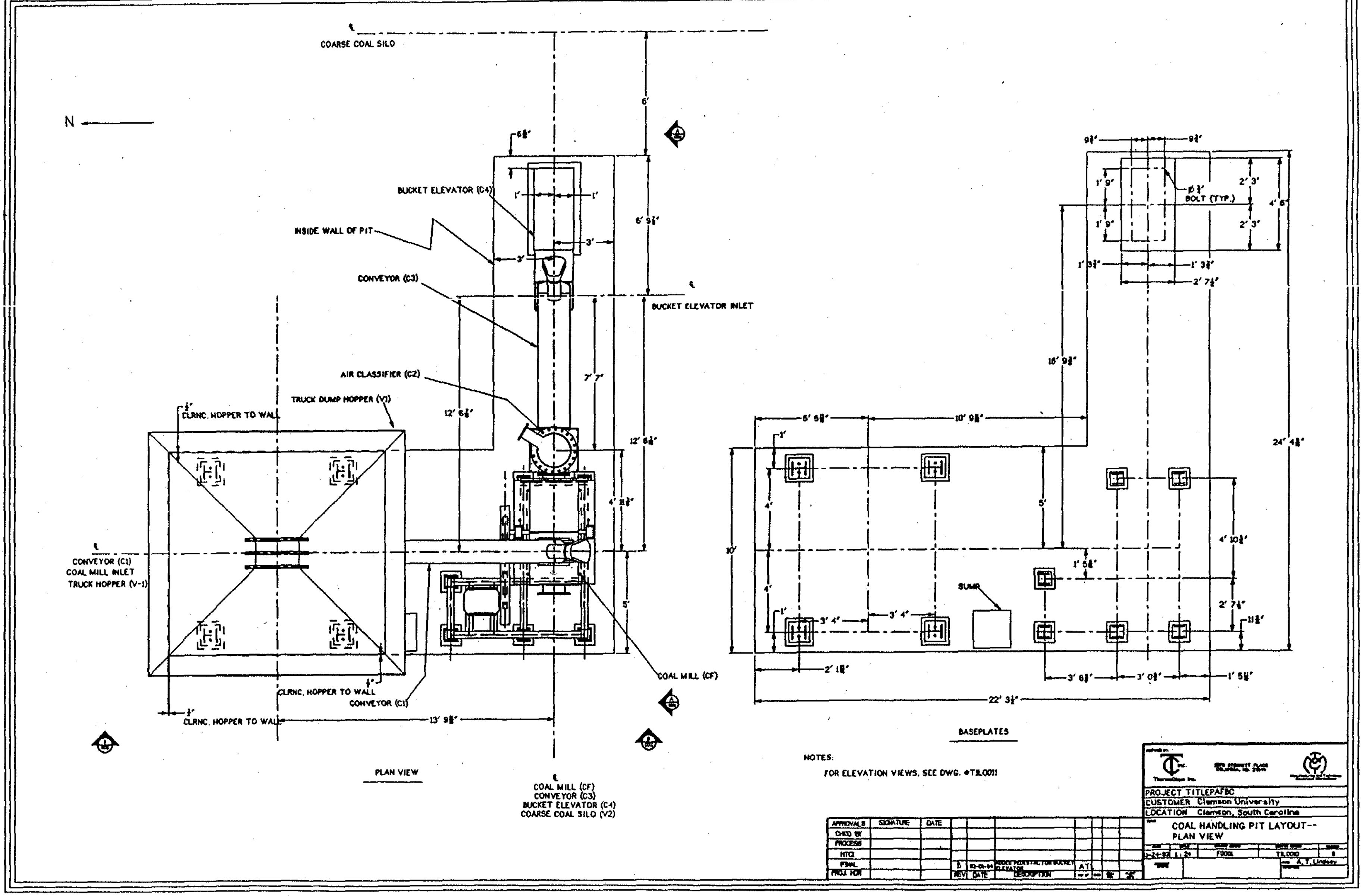


FIGURE 2-26: COAL HANDLING PIT LAYOUT - PLAN VIEW

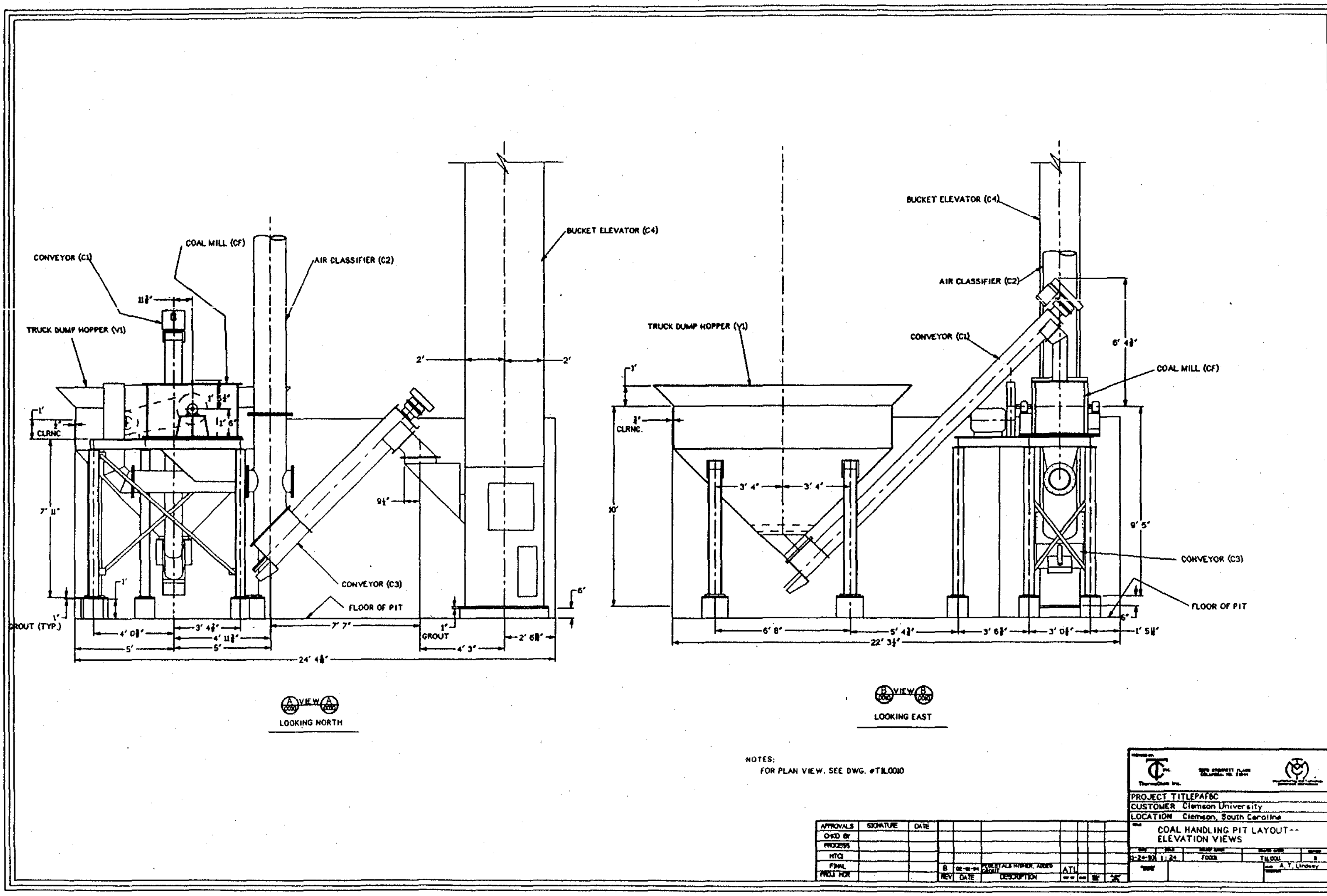
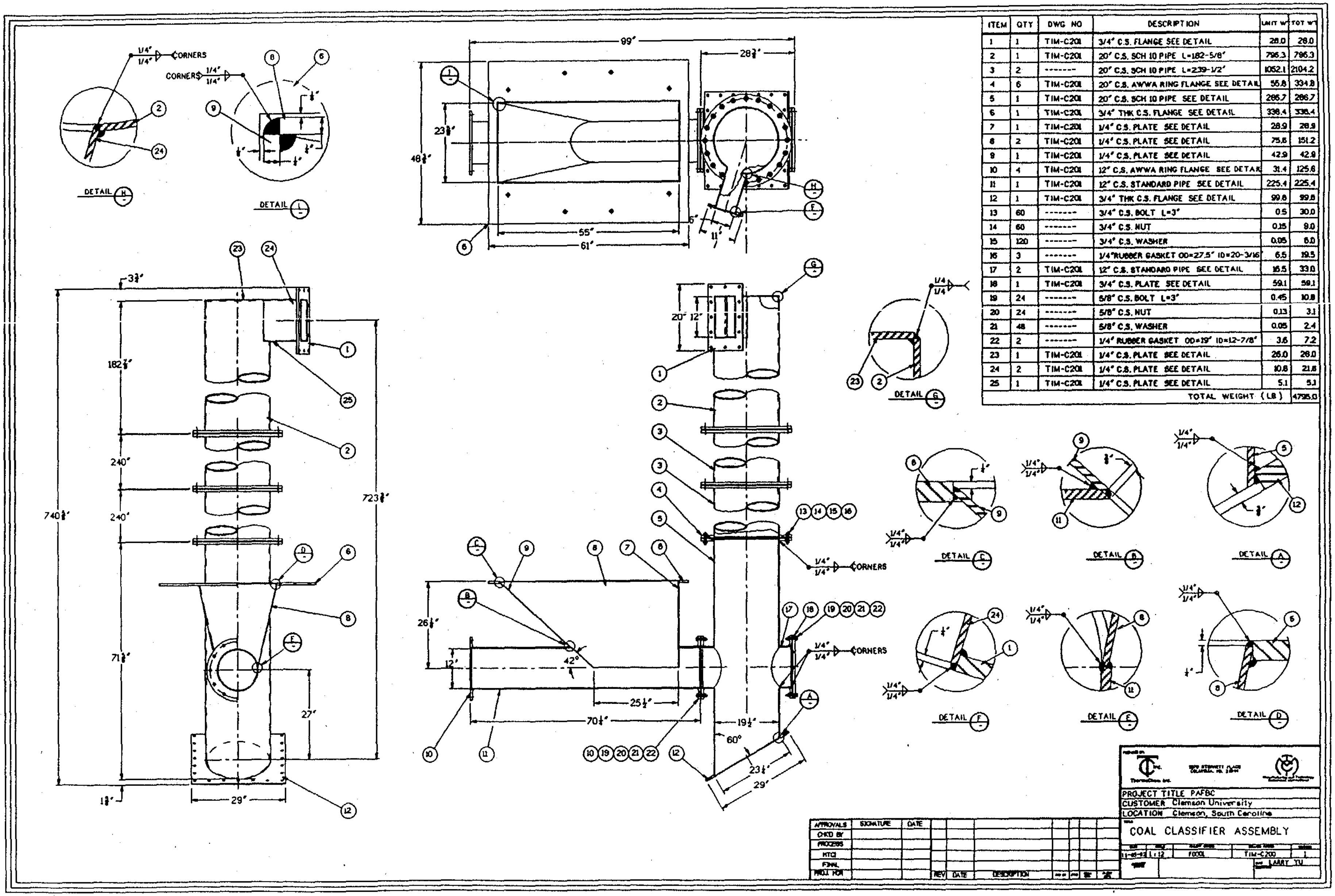
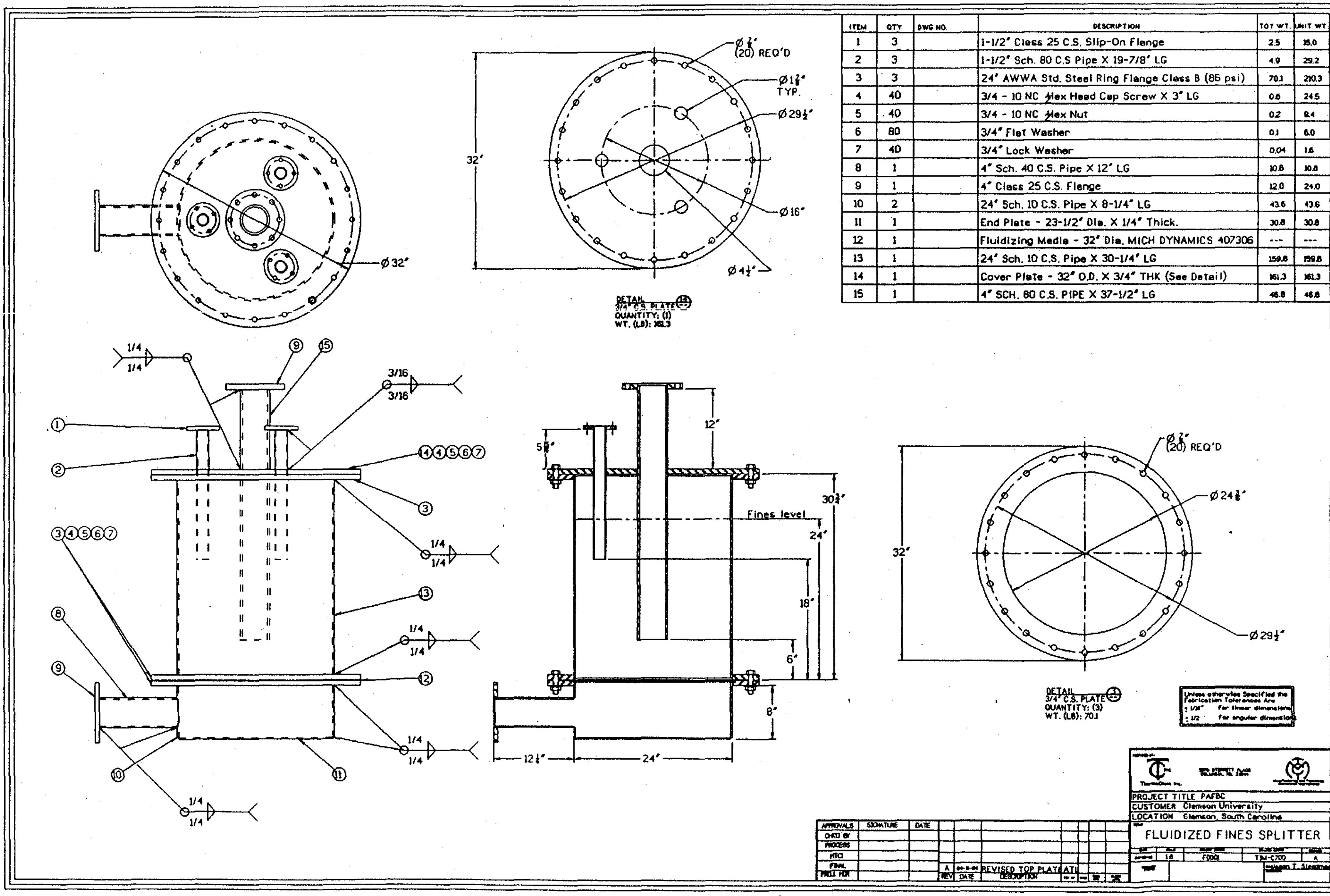


FIGURE 2-27: COAL HANDLING PIT LAYOUT - ELEVATION VIEWS



ITEM	QTY	DWG NO	DESCRIPTION	UNIT WT	TOT WT
1	1	TIM-C201	3/4" C.S. FLANGE SEE DETAIL	28.0	28.0
2	1	TIM-C201	20" C.S. SCH 10 PIPE L=182-5/8"	796.3	796.3
3	2	-------	20" C.S. SCH 10 PIPE L=239-1/2"	1052.1	2104.2
4	6	TIM-C201	20" C.S. AWWA RING FLANGE SEE DETAIL	55.8	334.8
5	1	TIM-C201	20" C.S. SCH 10 PIPE SEE DETAIL	286.7	286.7
6	1	TIM-C201	3/4" THK C.S. FLANGE SEE DETAIL	338.4	338.4
7	1	TIM-C201	1/4" C.S. PLATE SEE DETAIL	28.9	28.9
8	2	TIM-C201	1/4" C.S. PLATE SEE DETAIL	75.6	151.2
9	1	TIM-C201	1/4" C.S. PLATE SEE DETAIL	42.9	42.9
10	4	TIM-C201	12" C.S. AWWA RING FLANGE SEE DETAIL	31.4	125.6
11	1	TIM-C201	12" C.S. STANDARD PIPE SEE DETAIL	225.4	225.4
12	1	TIM-C201	3/4" THK C.S. FLANGE SEE DETAIL	99.8	99.8
13	60	-------	3/4" C.S. BOLT L=3"	0.5	30.0
14	60	-------	3/4" C.S. NUT	0.15	9.0
15	120	-------	3/4" C.S. WASHER	0.05	6.0
16	3	-------	1/4" RUBBER GASKET OD=27.5" ID=20-3/16"	6.5	19.5
17	2	TIM-C201	12" C.S. STANDARD PIPE SEE DETAIL	16.5	33.0
18	1	TIM-C201	3/4" C.S. PLATE SEE DETAIL	59.1	59.1
19	24	-------	5/8" C.S. BOLT L=3"	0.45	10.8
20	24	-------	5/8" C.S. NUT	0.13	3.1
21	48	-------	5/8" C.S. WASHER	0.05	2.4
22	2	-------	1/4" RUBBER GASKET OD=19" ID=12-7/8"	3.6	7.2
23	1	TIM-C201	1/4" C.S. PLATE SEE DETAIL	26.0	26.0
24	2	TIM-C201	1/4" C.S. PLATE SEE DETAIL	10.8	21.6
25	1	TIM-C201	1/4" C.S. PLATE SEE DETAIL	5.1	5.1
			TOTAL WEIGHT (LB)		4795.0

FIGURE 2-28: COAL CLASSIFIER ASSEMBLY



ITEM	QTY	DWG NO.	DESCRIPTION	TOT WT.	UNIT WT
1	3		1-1/2" Class 25 C.S. Slip-On Flange	2.5	15.0
2	3		1-1/2" Sch. 80 C.S Pipe X 19-7/8" LG	4.9	29.2
3	3		24" AWWA Std. Steel Ring Flange Class B (86 psi)	70.1	210.3
4	40		3/4 - 10 NC Hex Head Cap Screw X 3" LG	0.6	24.5
5	40		3/4 - 10 NC Hex Nut	0.2	9.4
6	80		3/4" Flat Washer	0.1	6.0
7	40		3/4" Lock Washer	0.04	1.6
8	1		4" Sch. 40 C.S. Pipe X 12" LG	10.8	10.8
9	1		4" Class 25 C.S. Flange	12.0	24.0
10	2		24" Sch. 10 C.S. Pipe X 8-1/4" LG	43.6	43.6
11	1		End Plate - 23-1/2" Dia. X 1/4" Thick.	30.8	30.8
12	1		Fluidizing Media - 32" Dia. MICH DYNAMICS 407306	---	---
13	1		24" Sch. 10 C.S. Pipe X 30-1/4" LG	159.8	159.8
14	1		Cover Plate - 32" O.D. X 3/4" THK (See Detail)	161.3	161.3
15	1		4" SCH. 80 C.S. PIPE X 37-1/2" LG	46.8	46.8

FIGURE 2-29: FLUIDIZED FINES SPLITTER

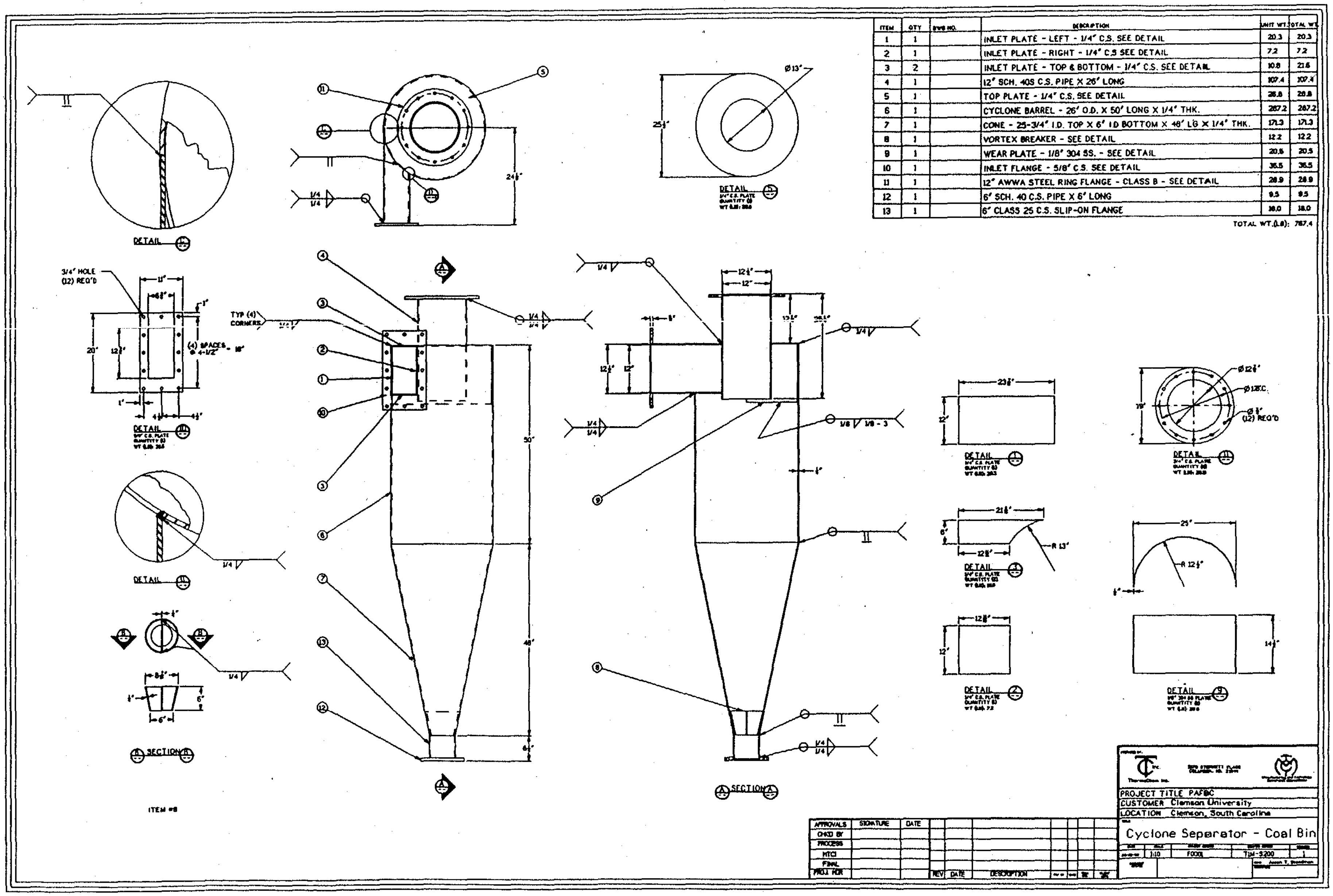


FIGURE 2-30: CYCLONE SEPARATOR - COAL BIN

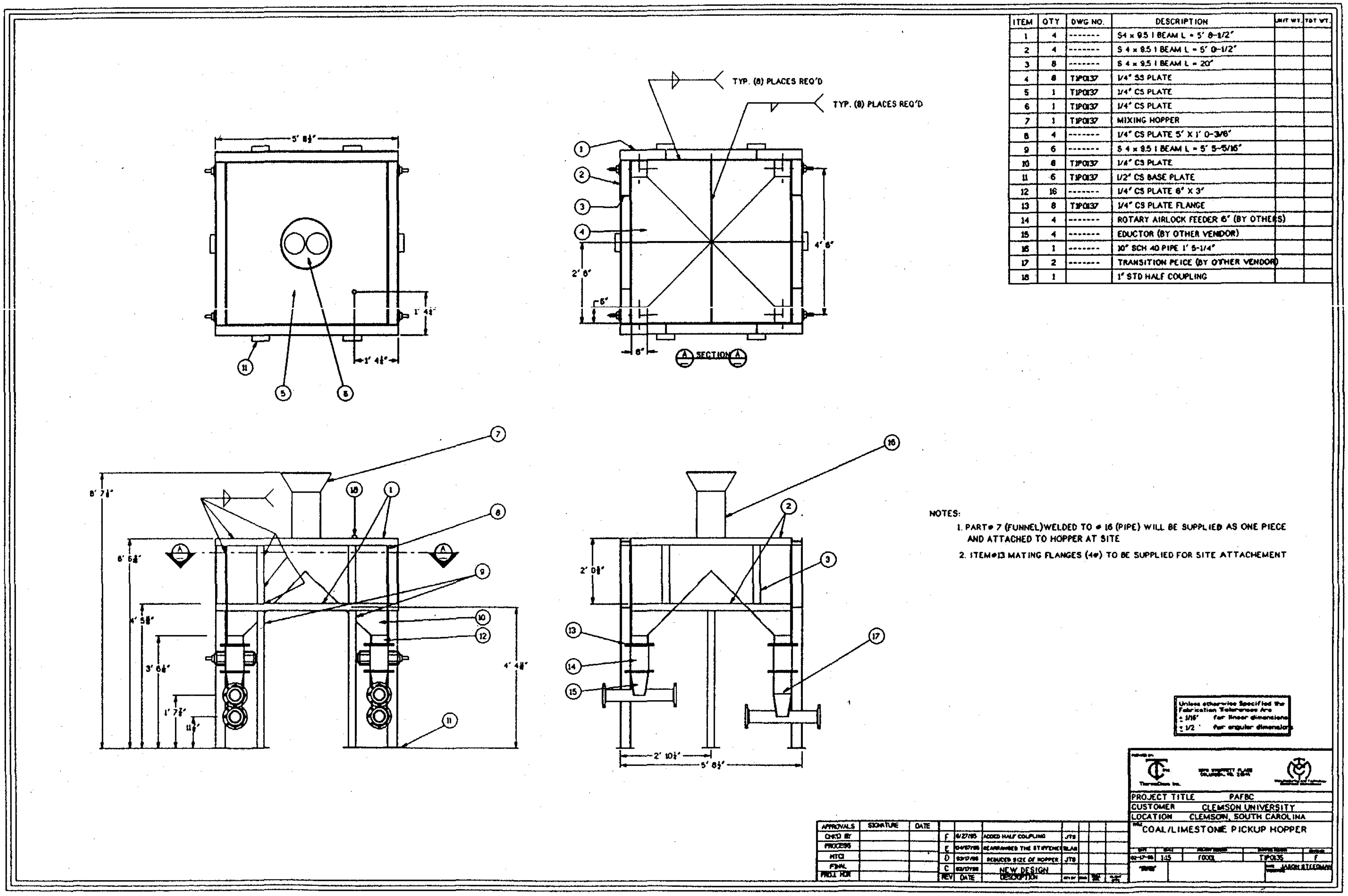


FIGURE 2-31: COAL/LIMESTONE PICKUP HOPPER

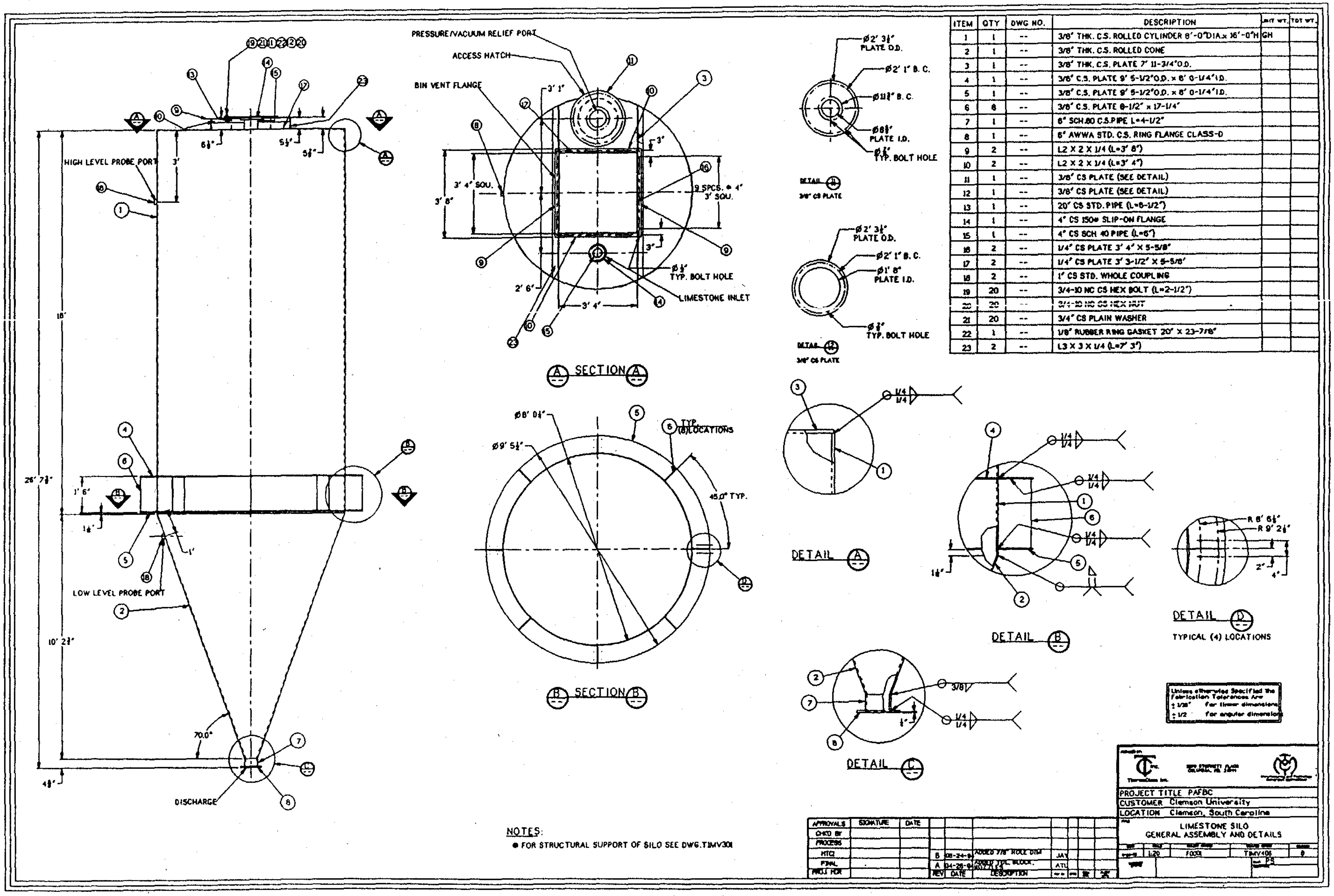


FIGURE 2-32: LIMESTONE SILO - GENERAL ASSEMBLY AND DETAILS

2.2.3 Pulse Combustor

The pulse combustor was located centrally on the roof of the bed vessel and will have only the combustion chamber outside the freeboard. The remainder of the pulse combustor will pass through the freeboard and end in the bed at a sufficient height above the distributor so as to place the stagnation point between the coal feed port and the pulse combustor exhaust. The location of the stagnation point is a key factor in the proper distribution of the coal through the bed by the pulse combustor gases. The correct location of the stagnation point will also ensure that the static bed material (layer insulating distributor plate) below the nozzles is not disturbed (see *Figure 2-33*). The design details of the pulse combustor are presented in *Table 2-4*.

The pulse combustor has six tailpipes and the is approximately 10 feet long. The combustor pulses at 62 Hz and the combustion gases issuing from the ends of the tailpipes pass through a decoupling passage before entering the fluid bed (see Figure 2-33.) The combustion chamber, tailpipes and decoupling chamber are all water-jacketed and constitute part of the entire PAFBC boiler surface.

Approximately 15 percent of the entire steam production is raised in the pulse combustor water jacket. The pulse combustor jacket has been designed for a maximum allowable working pressure of 175 psig which is the system working pressure. The pulse combustor produces saturated steam and is to be constructed entirely of carbon steel. The fuel ports and aerovalves are located in a header plate that is refractory-lined on the combustion chamber side. The air plenum flanges directly to the header plate on the opposite side from the combustion chamber and coal and gas feed lines run through the air plenum.

Figures 2-34 through *2-37* depict the coal injector aerovalves and plenum assembly for the pulse combustor. *Figure 2-38* and *2-39* are pictures of the exhaust of the combustor with and without fluid-bed steam tubes installed.

TABLE 2-4:

PULSE COMBUSTOR DATA

Coal flow, lb/hr	1,500
Coal conveying air flow, lb/hr	750
Coal conveying air pressure, "WG	60
Pulse combustor air flow, lb/hr	14,869.05
Pulse combustor air pressure, "WG	40
Flue gas flow, lb/hr	16,218
Combustion chamber surface area, sq.ft.	15.37
Temperature difference, °F	1,523
High temperature coefficient in chamber, Btu/hr-ft^2-°F	26
Steam raising, lb/hr	735.44
Chamber average wall temperature, °F	460.4
Tailpipe surface area, sq.ft.	84.6
Temperature difference, °F	1,523
High temperature coefficient in chamber, Btu/hr-ft^2-°F	32.8
Steam raising in tailpipes, lb/hr	5,140.3
Tailpipe average wall temperature, °F	441.5
Tailpipe jacket surface area in-bed, sq.ft.	20.02
Temperature difference, °F	1,173
Overall high transfer coefficient, Btu/hr-ft^2-°F	45
Steam raising, lb/hr	1,248.4
Tailpipe jacket surface area above bed, sq.ft.	83.3
Temperature difference, °F	1,173
Overall heat transfer coefficient, Btu/hr-ft^2-°F	5.6

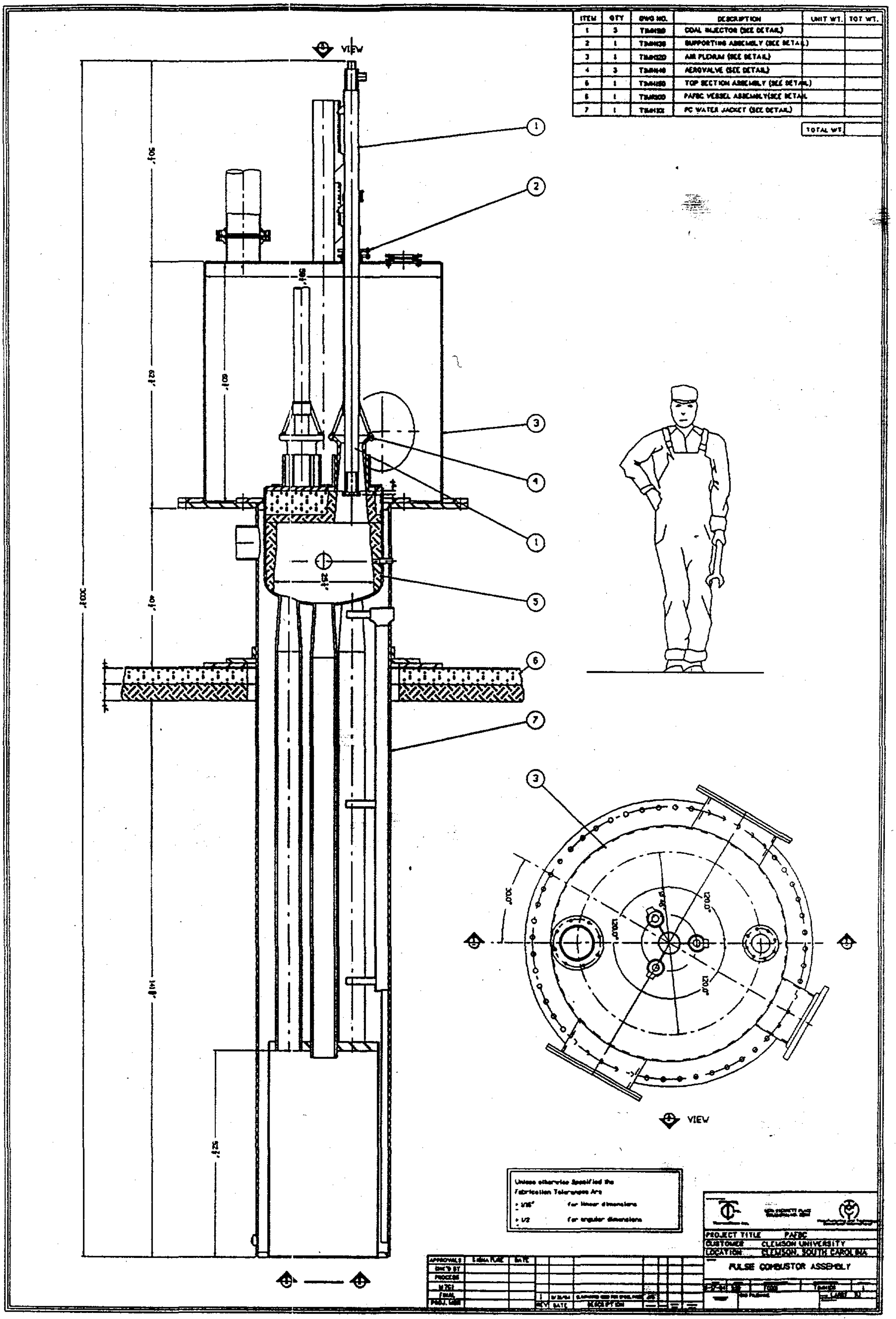


FIGURE 2-33: PULSE COMBUSTOR ASSEMBLY

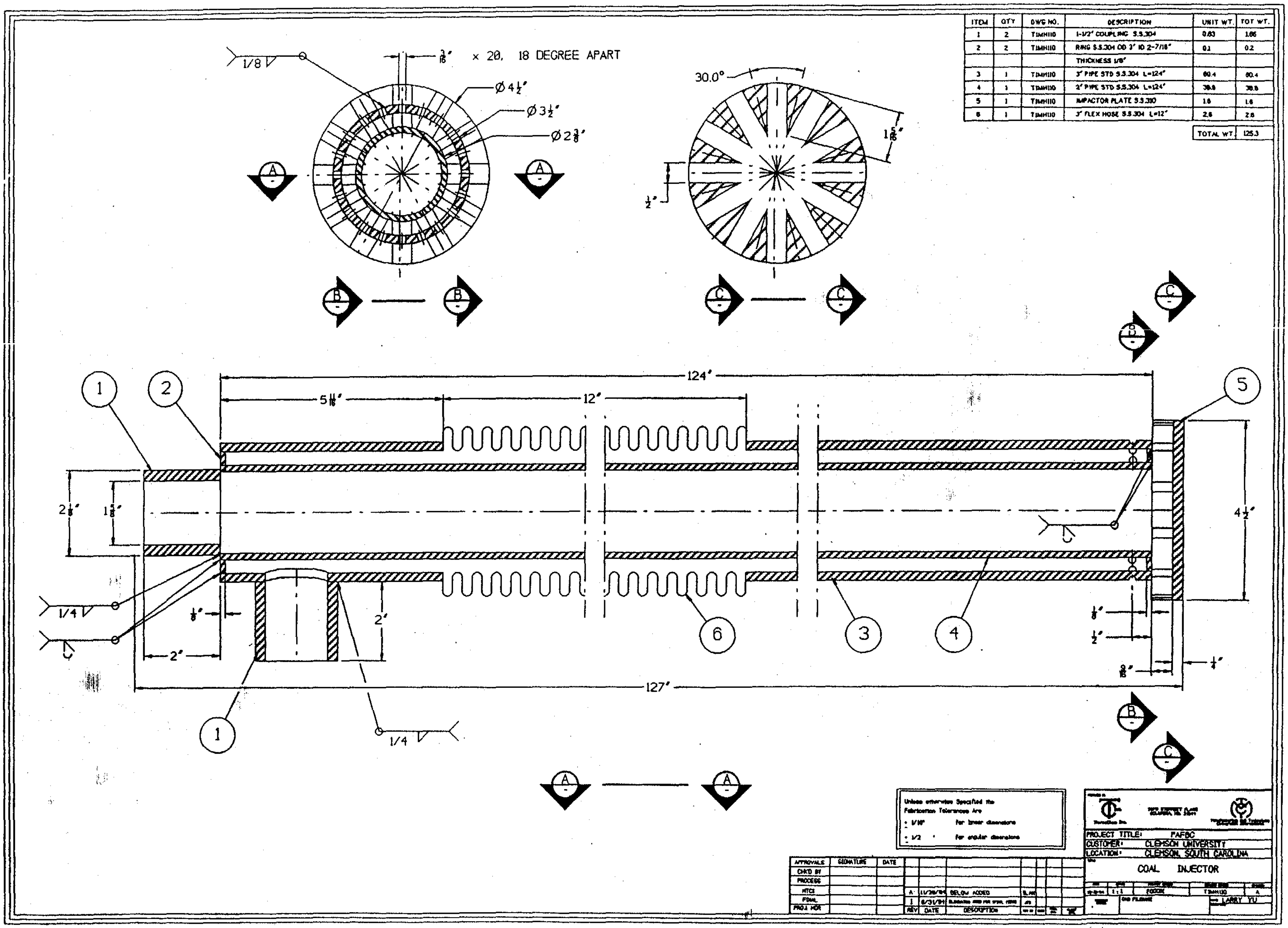
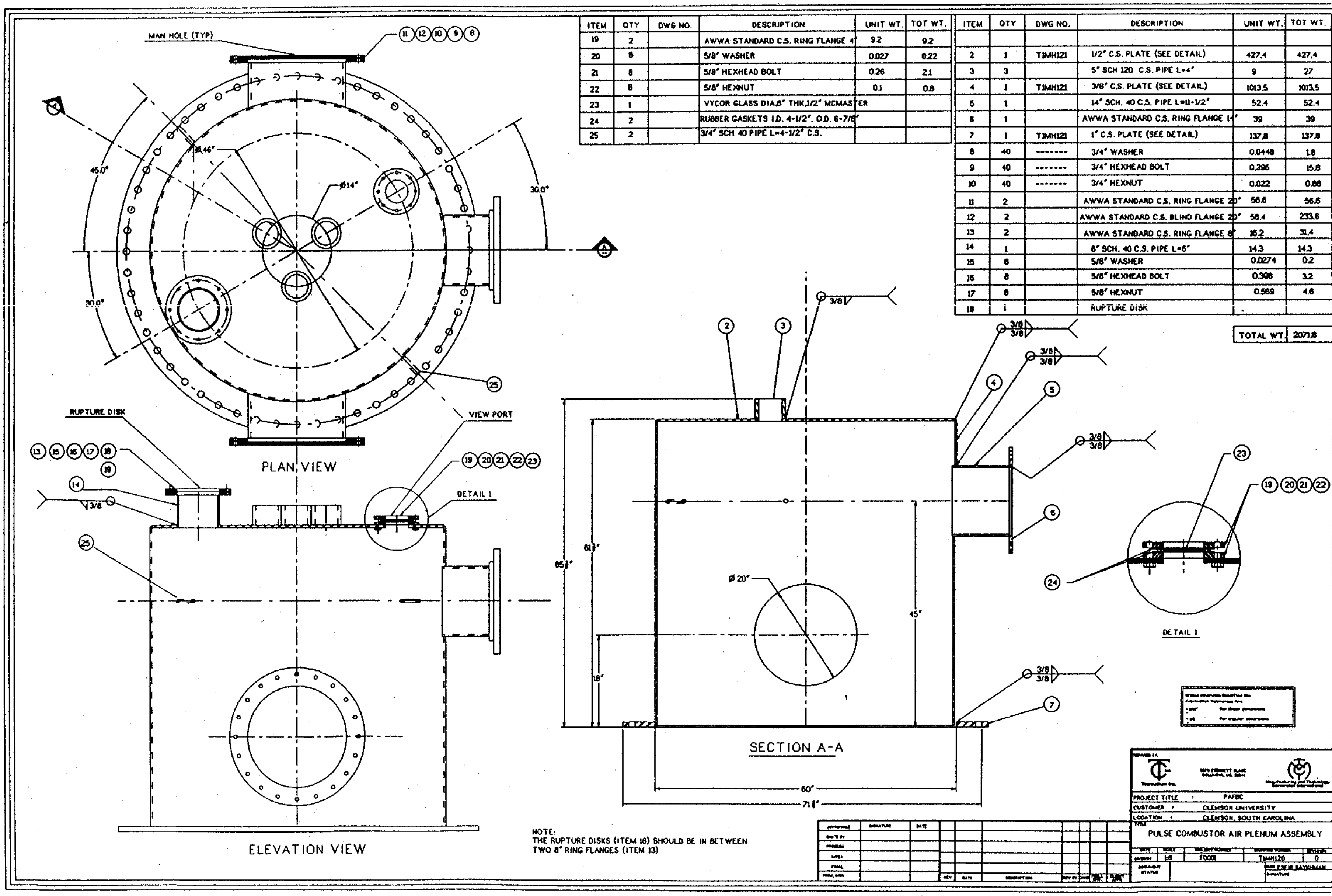


FIGURE 2-34: COAL INJECTOR



ITEM	QTY	DWG NO.	DESCRIPTION	UNIT WT.	TOT WT.
19	2		AWWA STANDARD C.S. RING FLANGE 4"	9.2	9.2
20	8		5/8" WASHER	0.027	0.22
21	8		5/8" HEXHEAD BOLT	0.26	2.1
22	8		5/8" HEXNUT	0.1	0.8
23	1		VYCOR GLASS DIA.6" THK.1/2" MCMASTER		
24	2		RUBBER GASKETS I.D. 4-1/2", O.D. 6-7/8"		
25	2		3/4" SCH 40 PIPE L=4-1/2" C.S.		

ITEM	QTY	DWG NO.	DESCRIPTION	UNIT WT.	TOT WT.
2	1	T1MH121	1/2" C.S. PLATE (SEE DETAIL)	427.4	427.4
3	3		5" SCH 120 C.S. PIPE L=4"	9	27
4	1	T1MH121	3/8" C.S. PLATE (SEE DETAIL)	1013.5	1013.5
5	1		14" SCH. 40 C.S. PIPE L=11-1/2"	52.4	52.4
6	1		AWWA STANDARD C.S. RING FLANGE 14"	39	39
7	1	T1MH121	1" C.S. PLATE (SEE DETAIL)	137.8	137.8
8	40	-------	3/4" WASHER	0.0448	1.8
9	40	-------	3/4" HEXHEAD BOLT	0.396	15.8
10	40	-------	3/4" HEXNUT	0.022	0.88
11	2		AWWA STANDARD C.S. RING FLANGE 20"	56.6	56.6
12	2		AWWA STANDARD C.S. BLIND FLANGE 20"	58.4	233.6
13	2		AWWA STANDARD C.S. RING FLANGE 8"	16.2	31.4
14	1		8" SCH. 40 C.S. PIPE L=6"	14.3	14.3
15	8		5/8" WASHER	0.0274	0.2
16	8		5/8" HEXHEAD BOLT	0.398	3.2
17	8		5/8" HEXNUT	0.569	4.6
18	1		RUPTURE DISK	-	

FIGURE 2-35: PULSE COMBUSTOR AIR PLENUM ASSEMBLY

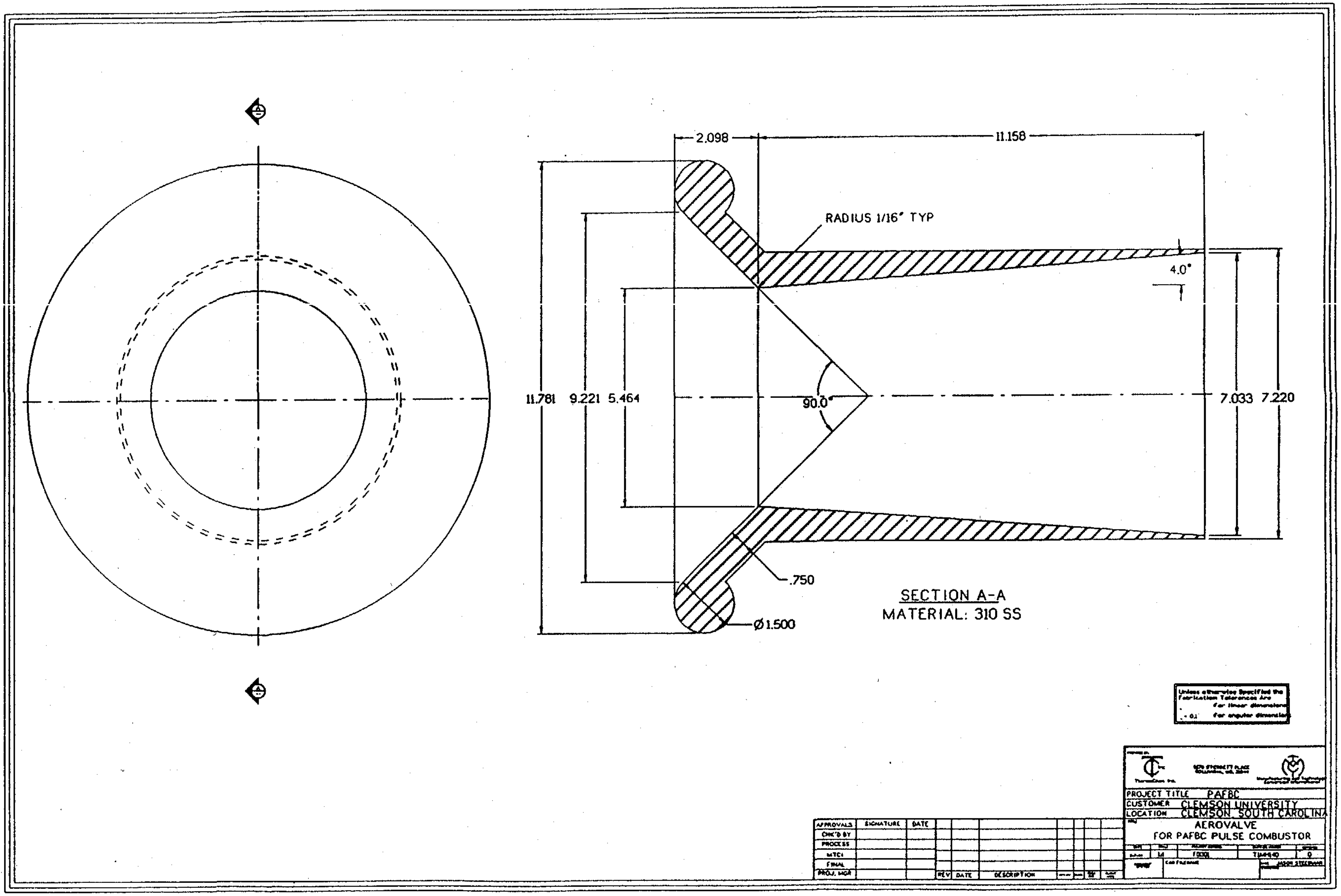
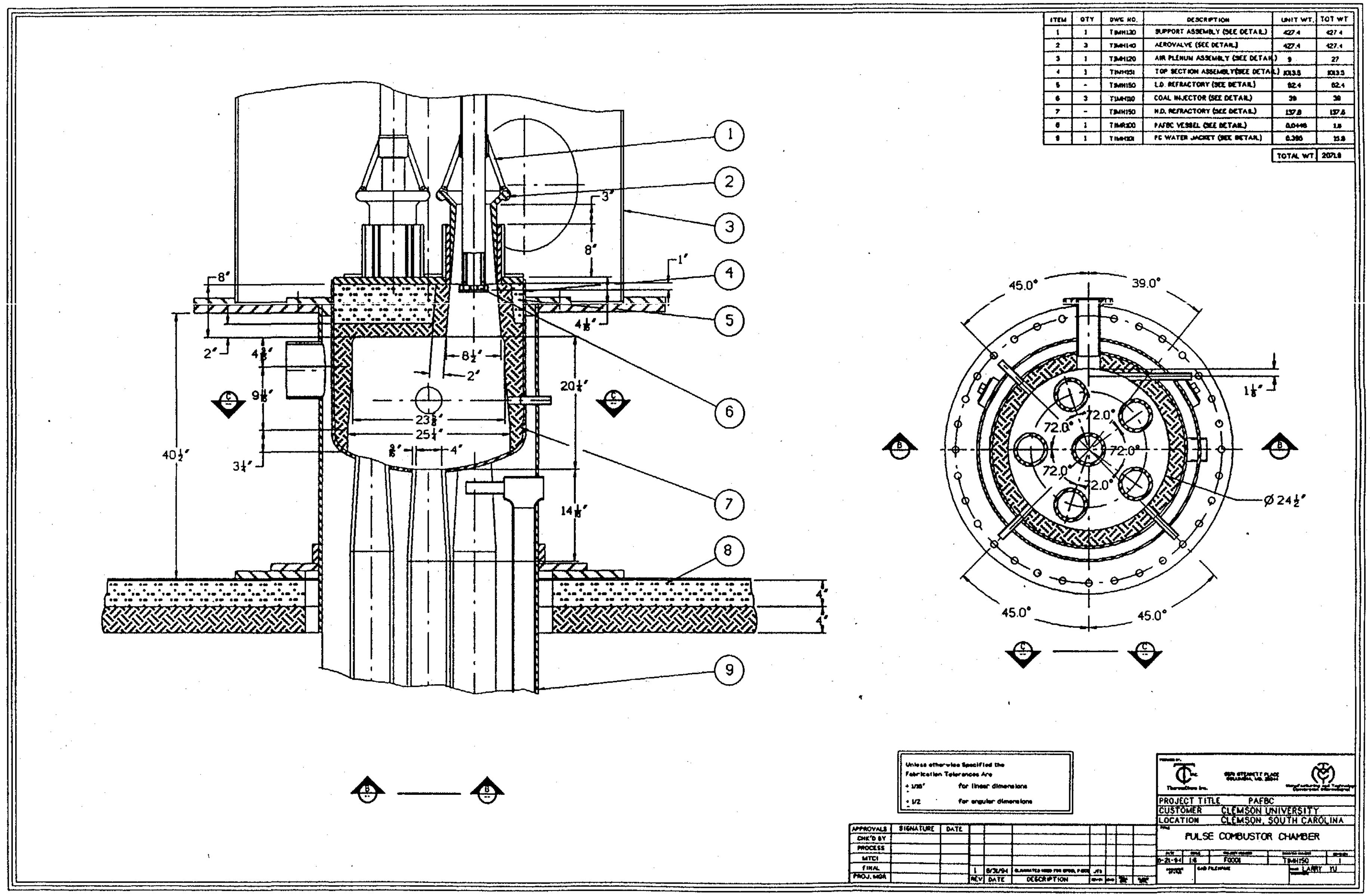


FIGURE 2-36: AEROVALVE FOR PAFBC PULSE COMBUSTOR



ITEM	QTY	DWG NO.	DESCRIPTION	UNIT WT.	TOT WT
1	1	TIMH130	SUPPORT ASSEMBLY (SEE DETAIL)	427.4	427.4
2	3	TIMH140	AEROVALVE (SEE DETAIL)	427.4	427.4
3	1	TIMH120	AIR PLENUM ASSEMBLY (SEE DETAIL)	9	27
4	1	TIMH151	TOP SECTION ASSEMBLY (SEE DETAIL)	1013.5	1013.5
5	-	TIMH150	L.D. REFRACTORY (SEE DETAIL)	82.4	82.4
6	3	TIMH130	COAL INJECTOR (SEE DETAIL)	39	39
7	-	TIMH150	H.D. REFRACTORY (SEE DETAIL)	137.8	137.8
8	1	TIMR100	PAFBC VESSEL (SEE DETAIL)	0.0446	1.8
9	1	TIMH101	FC WATER JACKET (SEE DETAIL)	0.386	15.8
				TOTAL WT	2071.8

FIGURE 2-37: PULSE COMBUSTOR CHAMBER

FIGURE 2-38: CLEMSON PAFBC UNIT - PULSE COMBUSTOR EXHAUST WITH IN-BED HEAT TRANSFER TUBES

FIGURE 2-39: CLEMSON PAFBC UNIT - PULSE COMBUSTOR EXHAUST WITH COAL FEED NOZZLES

The pulse combustor has a maximum firing rate of 18 MMBtu/hr. The 18 MMBtu/hr can be generated by and combination of gas and fine coal. The pulse combustor produces steam by transferring the energy of the exhaust gases into the water circulating in the water jacket surrounding the pulse combustor. A pulse combustor is used in place of a conventional burner because of the added benefits of lower emissions as well as enhanced heat transfer performance over the conventional burner. The fuel injectors for the pulse combustor supply air, natural gas, and/or fine coal to the combustion chamber. Natural gas flows through the outer annulus of the injector and the air and fine coal is delivered to the diffuser through the inner annulus of the injector. After the flue gases exit the pulse combustor, they deflect off of a target cone and become dispersed in the bed.

2.2.4 Bubbling Fluid Bed

The fluid bed has a maximum firing rate of 60 MMBtu/hr. The main fuel for the bed is coal. However, during start-up it is necessary to heat the bed up using natural gas. It is also possible, should something happen to the coal-feeding system, that natural gas can be used in the bed to generate steam. During normal operation, coal is injected into the bed using an underbed feeding mechanism. The natural gas, however, is injected into the bed through a system of branching pipes with holes in them. The pilot for the bed gas is located slightly above the heat transfer modules. Fluidization of the bed occurs by blowing air through 368 bubble caps. A typical bed depth for the Fluid Bed Combustor (R-1) is approximately 40 inches. As the coal burns, it transfer its heat into the in-bed heat transfer modules. Whereas heat transfer from gases to tubes is quite low, a fluid bed typically has a heat transfer coefficient roughly equal to that of boiling water. The bed, consisting of coal, limestone, and ash, loses the majority of its ash through entrainment. The fly ash is too fine to be captured by the Hot Cyclone (S-5) and therefore is captured in hoppers downstream of the Bubbling Bed Combustor.

Fluidized Bed

The fluid-bed vessel was constructed in three sections: 1) air plenum/distributor, 2) bed with cooling coils, and 3) freeboard with gas exit on the roof. The three sections are flanged together and the total height of the vessel is approximately 22 feet. In addition, the vessel legs provide a 5-foot clearance under the vessel for placement of coal/limestone and ash-handling equipment bring the total height of the vessel to 27 feet. See <ins>*Table 2-5*</ins> for the fluid bed detailed specifications.

The distributor plate is a filter media nozzle distributor design with the nozzles manufactured with a cylindrical top section made of a filter media constructed by laminating a perforated plate with two different sizes of mesh. The media-type distributor is intended to reduce, if not eliminate, backsifting. The distributor plate is made up of four, ¾-inch carbon steel quadrants that rest on a support grid of box sections. Coal and limestone are to be fed underbed through four separate ports and spent bed material is to be drained from four ports, one in each of the quadrants. The bed itself is 10' x 10' at the distributor plate and diverges to 12' x 12' at the top of its 44-inch expanded height.

The bed has 500 sq. ft. of heat transfer surface arranged in 18 modular coals. The heat transfer coils are designed so that surface is applied evenly throughout the height of the bed so that a constant rate of turndown can be achieved as the bed material is drained to expose the coils and reduce heat transfer. The coils are to be made of 2-1/2" Sch. 80 pipe with standard short radius 180-degree elbows. The modules are easily removable for repair or replacement and since there are only two kinds of modules (short and long), modules can be interchanged to any position in the bed, within the two groups.

The gases from the pulse combustor are directed through the 28-inch diameter decoupling chamber into the center of the bed directly over the coal feed port to ensure distribution of coal radially through the bed.

<ins>*Figures 2-40*</ins> through <ins>*2-45*</ins> depict the fluidized bed major components.

TABLE 2-5:
FLUID-BED DATA

DISTRIBUTOR SECTION	
Distributor air flow, lb/hr	43,875
Distributor air temperature, °F	200
Dp across distributor, "WG	12.6
Bed pressure drop, "WG	42
Dp distributor/Dp bed ratio	0.30
Number of nozzles per sq. ft.	3.92
Nozzle centerline spacing, inches	5.5
Number of nozzles in bed	392
FLUIDIZED BED SECTION	
Flue gas flow, lb/hr	66,624
Flue gas temperature, °F	1,550
Flue gas density, lb/cu.ft.	0.0197
Flue gas flow, CFM	56,365.48
Bed bottom cross-section area, sq.ft.	100
Bed top cross-section area, sq.ft.	144
Mid-bed cross-section area, sq.ft.	121
Inside area of pulse combustor gas duct, sq.ft.	4.28
"True" mid-bed cross-section area, sq.ft.	114.7
Min. fluidization velocity, FPS	1.5

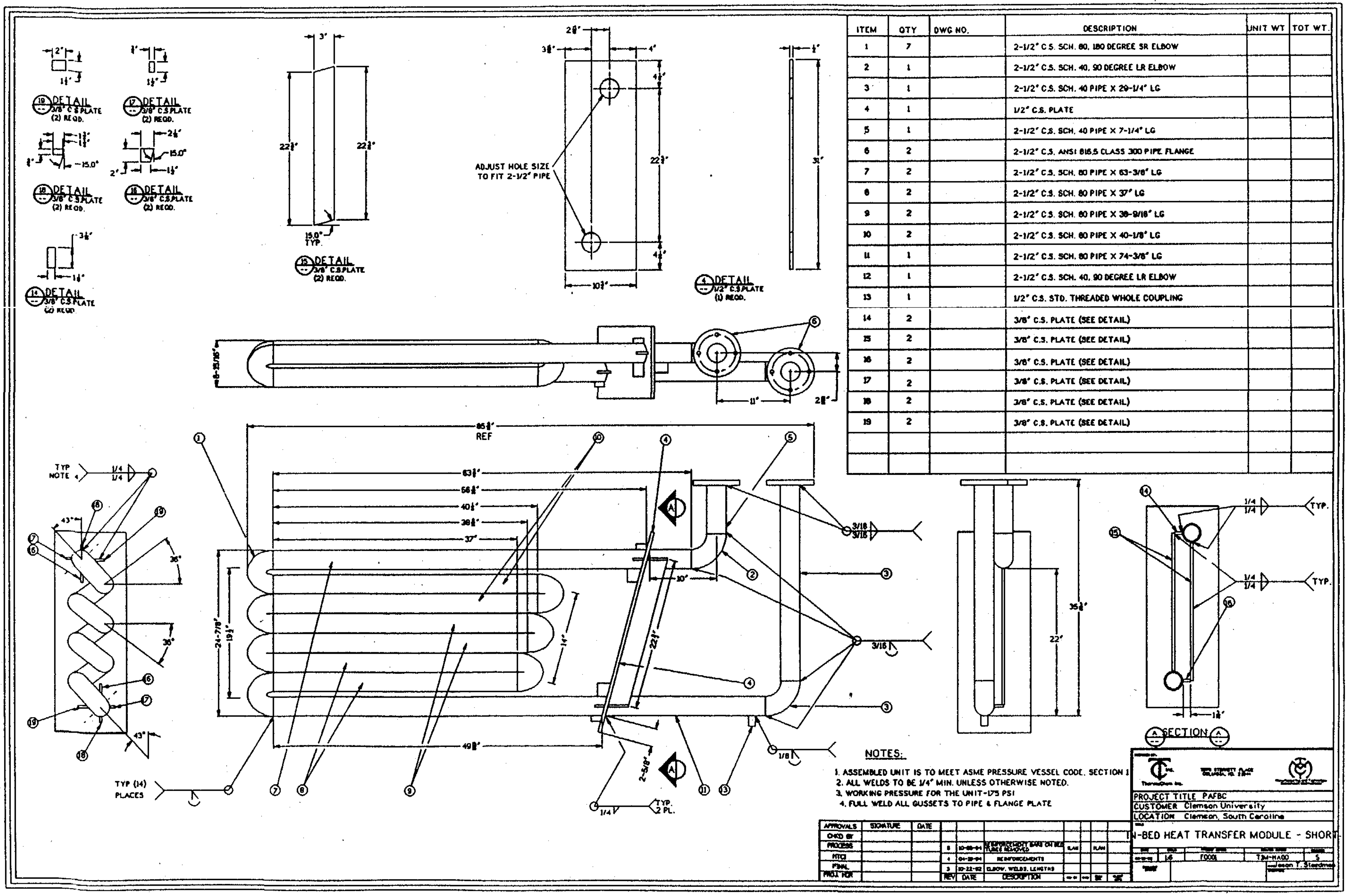


FIGURE 2-40: IN-BED HEAT TRANSFER MODULE - SHORT

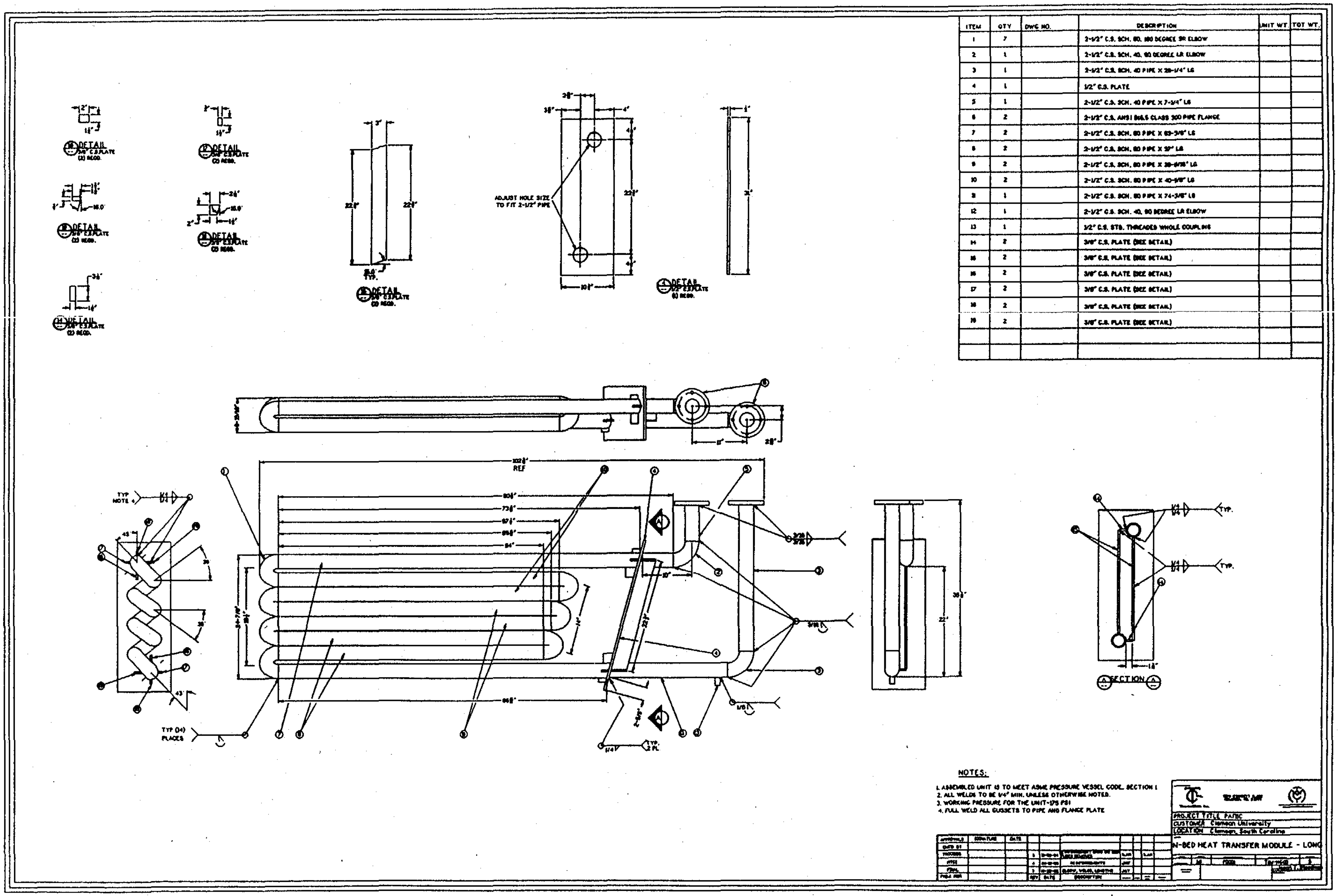
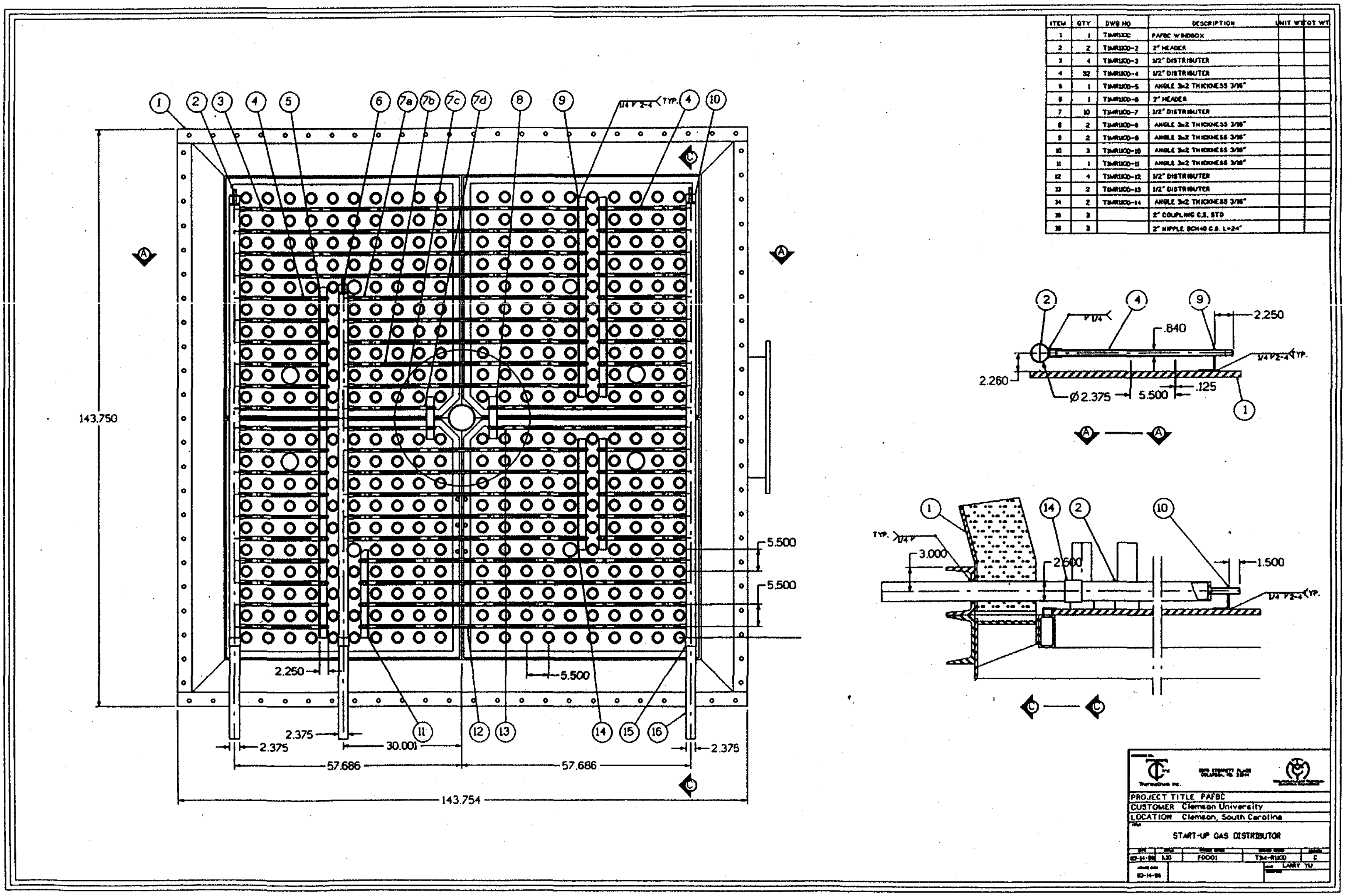
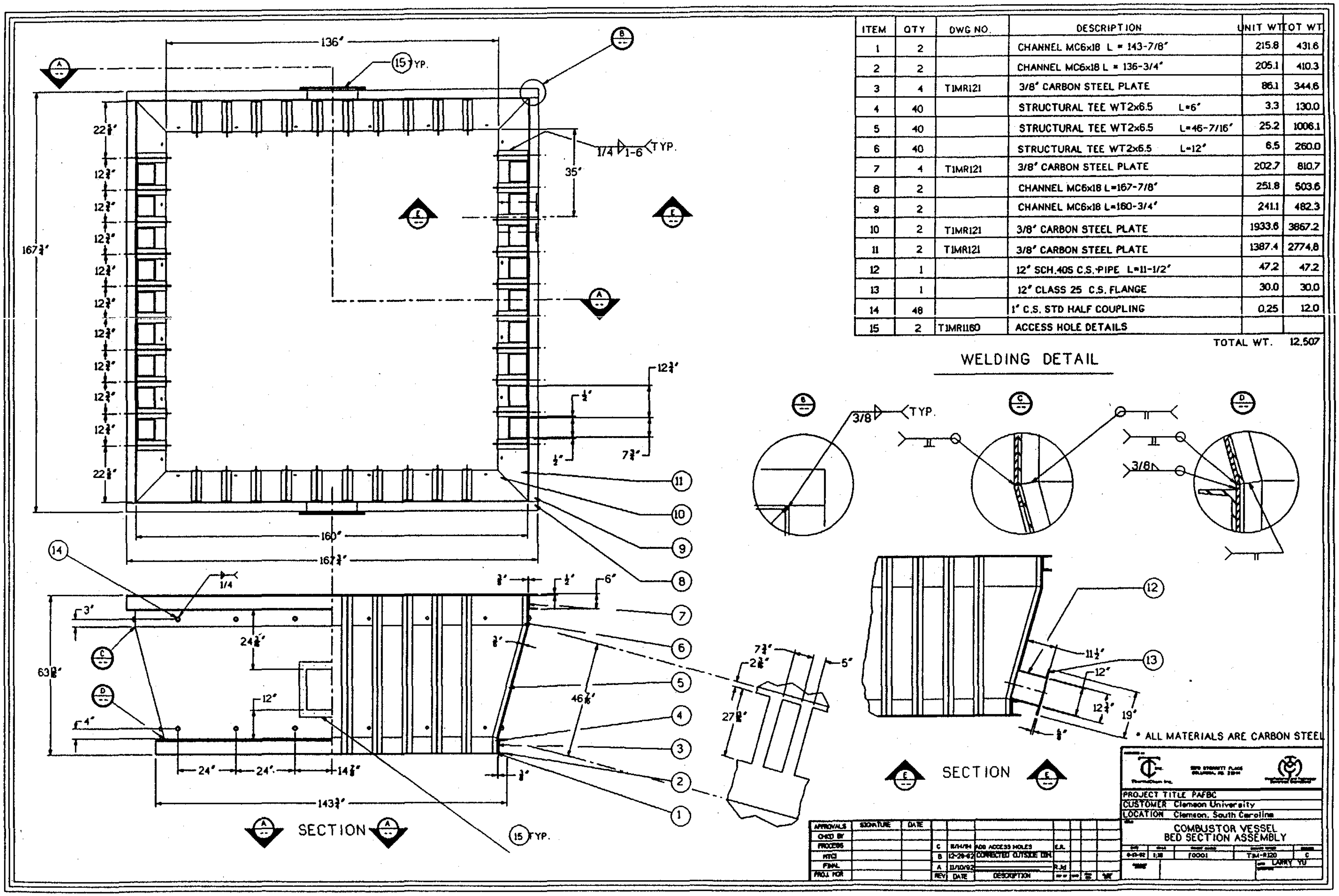


FIGURE 2-41: IN-BED HEAT TRANSFER MODULE - LONG



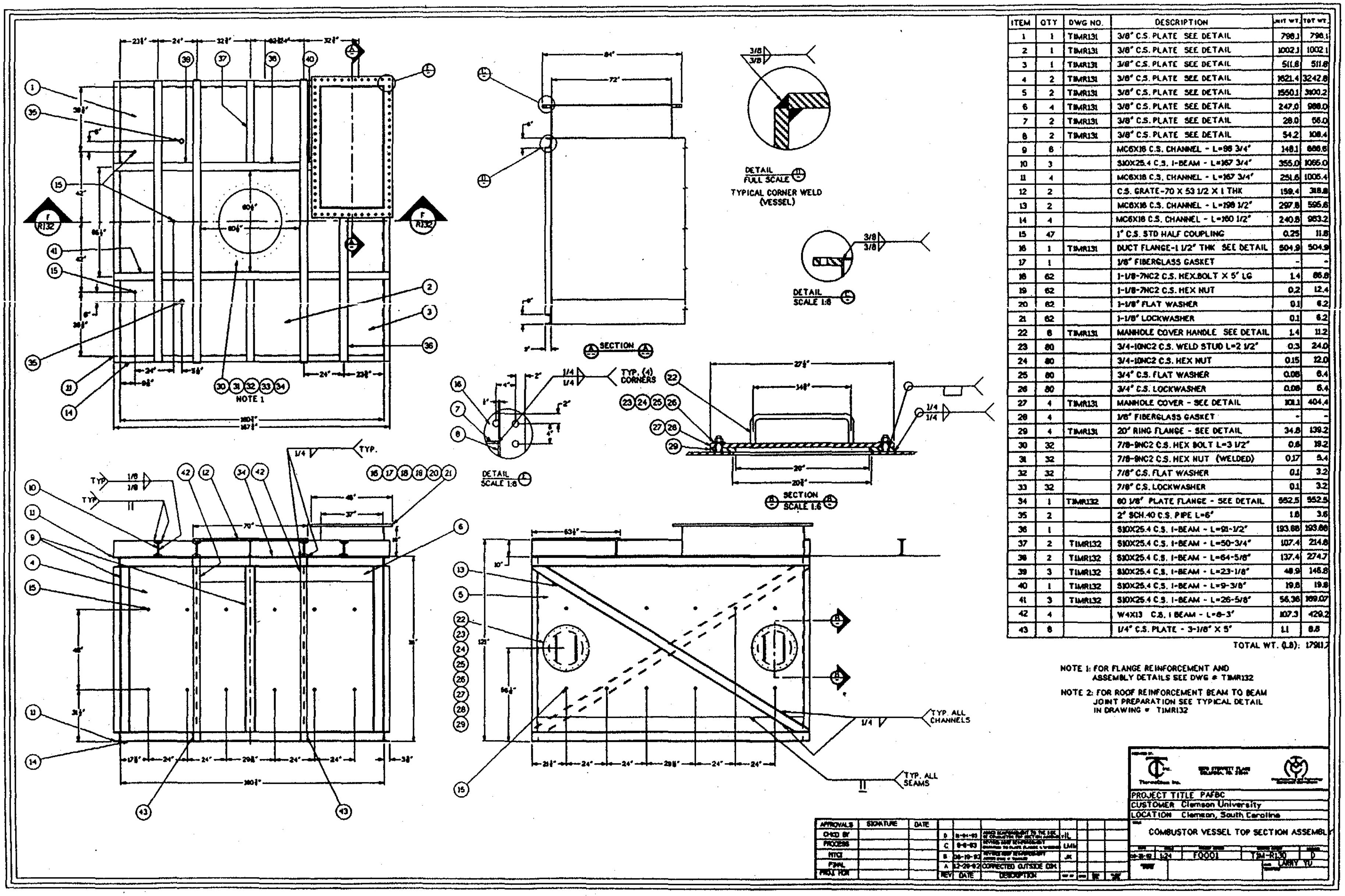
ITEM	QTY	DWG NO	DESCRIPTION	UNIT WT	TOT WT
1	1	T3MRU00	PAFBC WINDBOX		
2	2	T3MRU00-2	2" HEADER		
3	4	T3MRU00-3	1/2" DISTRIBUTER		
4	32	T3MRU00-4	1/2" DISTRIBUTER		
5	1	T3MRU00-5	ANGLE 3x2 THICKNESS 3/16"		
6	1	T3MRU00-6	2" HEADER		
7	10	T3MRU00-7	1/2" DISTRIBUTER		
8	2	T3MRU00-8	ANGLE 3x2 THICKNESS 3/16"		
9	2	T3MRU00-9	ANGLE 3x2 THICKNESS 3/16"		
10	3	T3MRU00-10	ANGLE 3x2 THICKNESS 3/16"		
11	1	T3MRU00-11	ANGLE 3x2 THICKNESS 3/16"		
12	4	T3MRU00-12	1/2" DISTRIBUTER		
13	2	T3MRU00-13	1/2" DISTRIBUTER		
14	2	T3MRU00-14	ANGLE 3x2 THICKNESS 3/16"		
15	3		2" COUPLING C.S. STD		
16	3		2" NIPPLE SCH40 C.S. L=24"		

FIGURE 2-42: START-UP GAS DISTRIBUTOR



ITEM	QTY	DWG NO.	DESCRIPTION	UNIT WT	TOT WT
1	2		CHANNEL MC6x18 L = 143-7/8"	215.8	431.6
2	2		CHANNEL MC6x18 L = 136-3/4"	205.1	410.3
3	4	T1MR121	3/8" CARBON STEEL PLATE	86.1	344.6
4	40		STRUCTURAL TEE WT2x6.5 L=6"	3.3	130.0
5	40		STRUCTURAL TEE WT2x6.5 L=46-7/16"	25.2	1006.1
6	40		STRUCTURAL TEE WT2x6.5 L=12"	6.5	260.0
7	4	T1MR121	3/8" CARBON STEEL PLATE	202.7	810.7
8	2		CHANNEL MC6x18 L=167-7/8"	251.8	503.6
9	2		CHANNEL MC6x18 L=160-3/4"	241.1	482.3
10	2	T1MR121	3/8" CARBON STEEL PLATE	1933.6	3867.2
11	2	T1MR121	3/8" CARBON STEEL PLATE	1387.4	2774.8
12	1		12" SCH.40S C.S. PIPE L=11-1/2"	47.2	47.2
13	1		12" CLASS 25 C.S. FLANGE	30.0	30.0
14	48		1" C.S. STD HALF COUPLING	0.25	12.0
15	2	T1MR1160	ACCESS HOLE DETAILS		
			TOTAL WT.		12,507

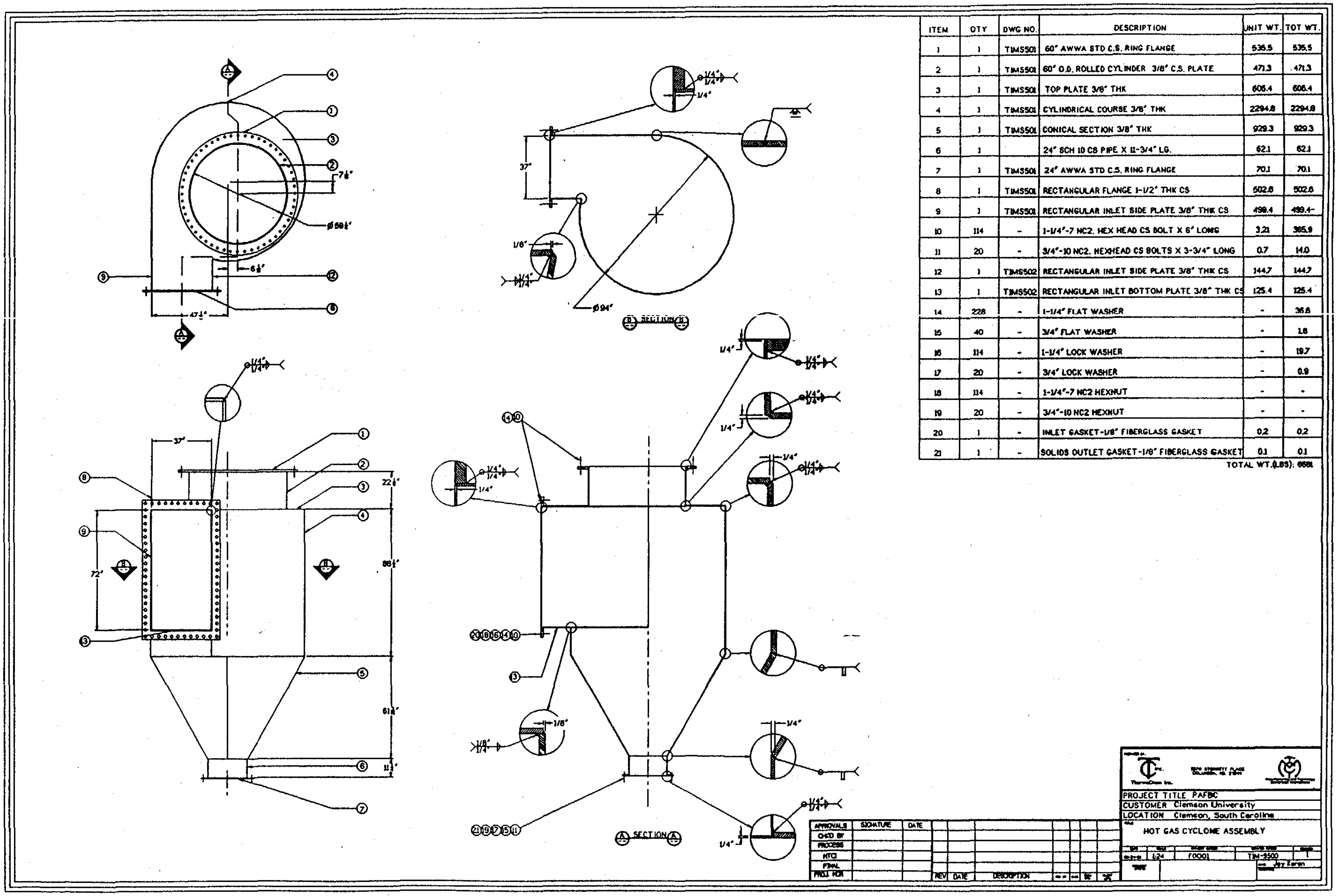
FIGURE 2-43: COMBUSTOR VESSEL BED SECTION ASSEMBLY



ITEM	QTY	DWG NO.	DESCRIPTION	UNIT WT.	TOT WT.
1	1	TIMR131	3/8" C.S. PLATE SEE DETAIL	798.1	798.1
2	1	TIMR131	3/8" C.S. PLATE SEE DETAIL	1002.1	1002.1
3	1	TIMR131	3/8" C.S. PLATE SEE DETAIL	511.8	511.8
4	2	TIMR131	3/8" C.S. PLATE SEE DETAIL	1621.4	3242.8
5	2	TIMR131	3/8" C.S. PLATE SEE DETAIL	1550.1	3100.2
6	4	TIMR131	3/8" C.S. PLATE SEE DETAIL	247.0	988.0
7	2	TIMR131	3/8" C.S. PLATE SEE DETAIL	28.0	56.0
8	2	TIMR131	3/8" C.S. PLATE SEE DETAIL	54.2	108.4
9	6		MC6X18 C.S. CHANNEL - L=98 3/4"	148.1	888.6
10	3		S10X25.4 C.S. I-BEAM - L=167 3/4"	355.0	1065.0
11	4		MC6X18 C.S. CHANNEL - L=167 3/4"	251.6	1006.4
12	2		C.S. GRATE-70 X 53 1/2 X 1 THK	159.4	318.8
13	2		MC6X18 C.S. CHANNEL - L=198 1/2"	297.8	595.6
14	4		MC6X18 C.S. CHANNEL - L=160 1/2"	240.8	963.2
15	47		1" C.S. STD HALF COUPLING	0.25	11.8
16	1	TIMR131	DUCT FLANGE-1 1/2" THK SEE DETAIL	504.9	504.9
17	1		1/8" FIBERGLASS GASKET	-	-
18	62		1-1/8-7NC2 C.S. HEX.BOLT X 5" LG	1.4	86.8
19	62		1-1/8-7NC2 C.S. HEX NUT	0.2	12.4
20	62		1-1/8" FLAT WASHER	0.1	6.2
21	62		1-1/8" LOCKWASHER	0.1	6.2
22	8	TIMR131	MANHOLE COVER HANDLE SEE DETAIL	1.4	11.2
23	80		3/4-10NC2 C.S. WELD STUD L=2 1/2"	0.3	24.0
24	80		3/4-10NC2 C.S. HEX NUT	0.15	12.0
25	80		3/4" C.S. FLAT WASHER	0.08	6.4
26	80		3/4" C.S. LOCKWASHER	0.08	6.4
27	4	TIMR131	MANHOLE COVER - SEE DETAIL	101.1	404.4
28	4		1/8" FIBERGLASS GASKET	-	-
29	4	TIMR131	20" RING FLANGE - SEE DETAIL	34.8	139.2
30	32		7/8-9NC2 C.S. HEX BOLT L=3 1/2"	0.6	19.2
31	32		7/8-9NC2 C.S. HEX NUT (WELDED)	0.17	5.4
32	32		7/8" C.S. FLAT WASHER	0.1	3.2
33	32		7/8" C.S. LOCKWASHER	0.1	3.2
34	1	TIMR132	80 1/8" PLATE FLANGE - SEE DETAIL	552.5	552.5
35	2		2" SCH.40 C.S. PIPE L=6"	1.8	3.6
36	1		S10X25.4 C.S. I-BEAM - L=91-1/2"	193.88	193.88
37	2	TIMR132	S10X25.4 C.S. I-BEAM - L=50-3/4"	107.4	214.8
38	2	TIMR132	S10X25.4 C.S. I-BEAM - L=64-5/8"	137.4	274.7
39	3	TIMR132	S10X25.4 C.S. I-BEAM - L=23-1/8"	48.9	146.8
40	1	TIMR132	S10X25.4 C.S. I-BEAM - L=9-3/8"	19.8	19.8
41	3	TIMR132	S10X25.4 C.S. I-BEAM - L=26-5/8"	56.36	169.07
42	4		W4X13 C.S. I BEAM - L=8-3"	107.3	429.2
43	8		1/4" C.S. PLATE - 3-1/8" X 5"	1.1	8.8

TOTAL WT. (LB): 17911.7

FIGURE 2-44: COMBUSTOR VESSEL TOP SECTION ASSEMBLY



ITEM	QTY	DWG NO.	DESCRIPTION	UNIT WT.	TOT WT.
1	1	TIMS501	60" AWWA STD C.S. RING FLANGE	535.5	535.5
2	1	TIMS501	60" O.D. ROLLED CYLINDER 3/8" C.S. PLATE	471.3	471.3
3	1	TIMS501	TOP PLATE 3/8" THK	606.4	606.4
4	1	TIMS501	CYLINDRICAL COURSE 3/8" THK	2294.8	2294.8
5	1	TIMS501	CONICAL SECTION 3/8" THK	929.3	929.3
6	1		24" SCH 10 CS PIPE X 11-3/4" LG.	62.1	62.1
7	1	TIMS501	24" AWWA STD C.S. RING FLANGE	70.1	70.1
8	1	TIMS501	RECTANGULAR FLANGE 1-1/2" THK CS	502.6	502.6
9	1	TIMS501	RECTANGULAR INLET SIDE PLATE 3/8" THK CS	499.4	499.4-
10	114	-	1-1/4"-7 NC2, HEX HEAD CS BOLT X 6" LONG	3.21	365.9
11	20	-	3/4"-10 NC2, HEXHEAD CS BOLTS X 3-3/4" LONG	0.7	14.0
12	1	TIMS502	RECTANGULAR INLET SIDE PLATE 3/8" THK CS	144.7	144.7
13	1	TIMS502	RECTANGULAR INLET BOTTOM PLATE 3/8" THK CS	125.4	125.4
14	228	-	1-1/4" FLAT WASHER	-	36.6
15	40	-	3/4" FLAT WASHER	-	1.6
16	114	-	1-1/4" LOCK WASHER	-	19.7
17	20	-	3/4" LOCK WASHER	-	0.9
18	114	-	1-1/4"-7 NC2 HEXNUT	-	-
19	20	-	3/4"-10 NC2 HEXNUT	-	-
20	1	-	INLET GASKET-1/8" FIBERGLASS GASKET	0.2	0.2
21	1	-	SOLIDS OUTLET GASKET-1/8" FIBERGLASS GASKET	0.1	0.1

TOTAL WT.(LBS): 6681

FIGURE 2-45: HOT GAS CYCLONE ASSEMBLY

2.2.5 Heat Recovery and Particulate Removal Section

This section consists of the waste heat recovery boiler, an economizer, a baghouse, an induced-draft fan, and the stack. The boiler is a package boiler consisting of a chamber through which the flue gases pass, a mud drum located under the boiler chamber, and a steam drum above. Tubes connecting the mud drum and steam drum line the boiler chamber and provide the heat transfer surface. Gas flows horizontally through the boiler. Entrained solids in the flue gas will drop through spaces between tubes along the floor of the boiler chamber and be discharged through ash drops into the ash-handling system. Water is fed into the boiler as required to maintain levels from the economizer. Boiler design specifications are shown in *Table 2-6*.

The economizer provides non-steaming heat transfer surface to further cool the gases exiting the boiler while preheating the boiler feed water. This unit consists of widely spaced tubes with the gas outside the tubes flowing downward and the water inside the tubes flowing upward. This counterflow design minimizes the surface area required and reduces draft loss. The upflow of water eliminates unstable water flow, providing a more effective, uniform distribution. An ash collection and discharge hopper is provided. The baghouse is of conventional design. Flue gas enters the baghouse and passes through fabric bags that remove essentially all of the entrained dust. Solids are discharged into the ash-handling system. Clean flue gases from the baghouse enter the induced draft fan and then are stacked. *Figures 2-46* and *2-47* show the boiler inlet duct and inlet ash hopper. *Figures 2-48* and *2-49* show the economizer. The PAFBC in-bed heat transfer surface is shown in *Figure 2-50*. The ash screw catch hopper flow diagram is shown in *Figure 2-51* and the ash control hopper is shown in *Figure 2-52*. The stack is shown in *Figure 2-53*. *Figure 2-54* shows the water-cooled ash removal system.

2.2.6 Ash

When the bed gets too high, the bed ash is removed through two water-cooled ash conveying screws (E-2 and E-3). These screws are located at the bottom of the Fluid Bed Combustor (R-1). Each screw empties the ash into its own holding hopper until the ash removal system empties the hoppers. The ash-handling system design specifications are shown in *Table 2-7*.

TABLE 2-6:
BOILER DESIGN DATA

CONVECTIVE PASS BOILER BANK (Fire Tube)	
Inlet gas flow, lb/hr	66,624
Inlet gas temperature, °F	1,550
Outlet temperature, °F	650
Boiler bank surface area, sq.ft.	2,590
Temperature difference, °F	600
Overall heat transfer coefficient, Btu/hr-ft^2-°F	10.2
Steam raising in boiler bank, lb/hr	18,721.6
Average metal temperature, °F	381.75
ECONOMIZER	
Inlet gas flow, lb/hr	66,624
Inlet gas temperature, °F	650
Outlet gas temperature, °F	350
Economizer surface area, sq.ft.	2,800
Temperature difference, °F	272
Overall heat transfer coefficient, Btu/hr-ft^2-°F	8.1
Inlet water temperature, °F	170
Inlet enthalpy, 80°F basis, Btu/lb	199.93
Water temperature out, °F	276

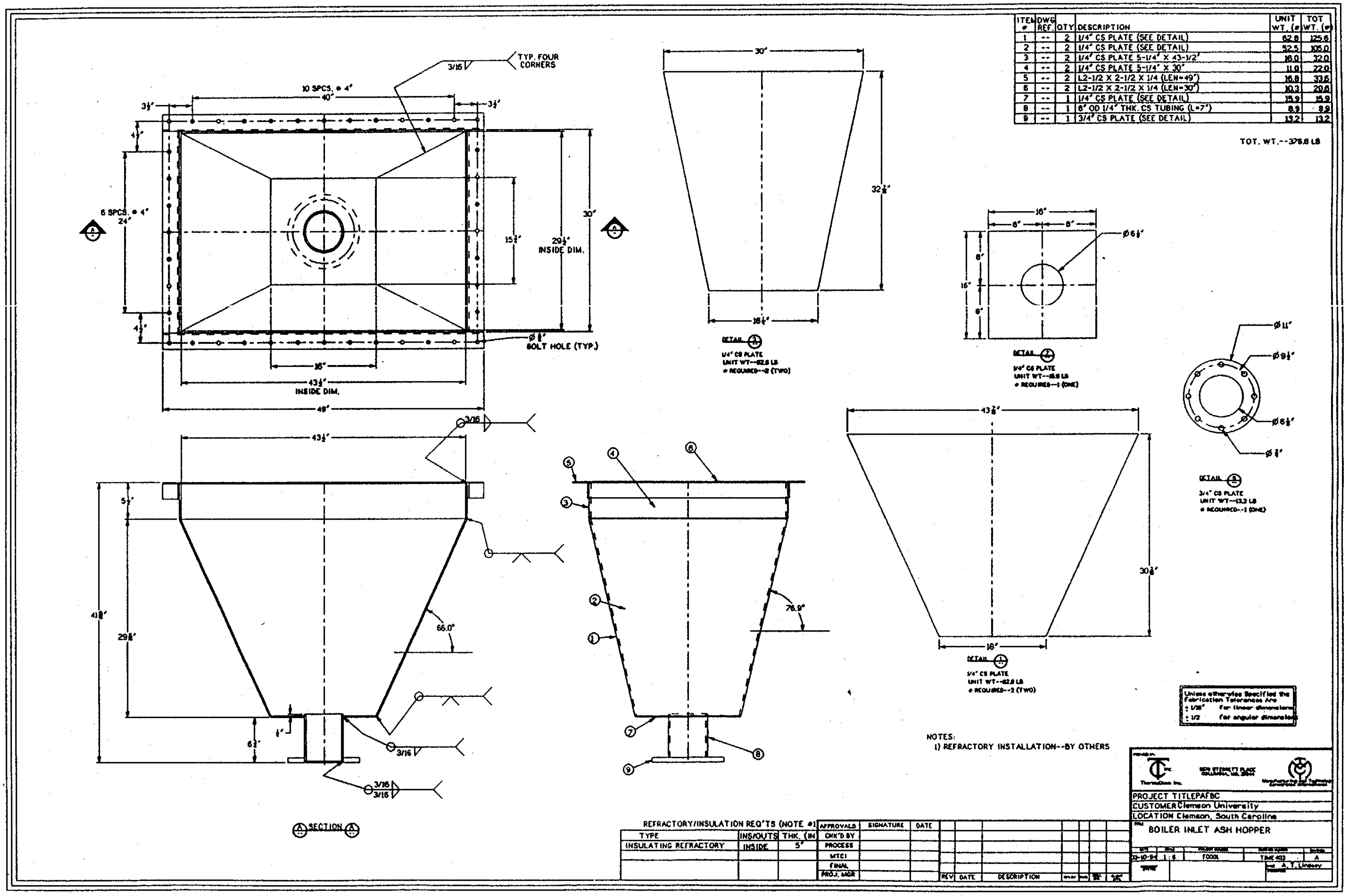


FIGURE 2-46: BOILER INLET ASH HOPPER

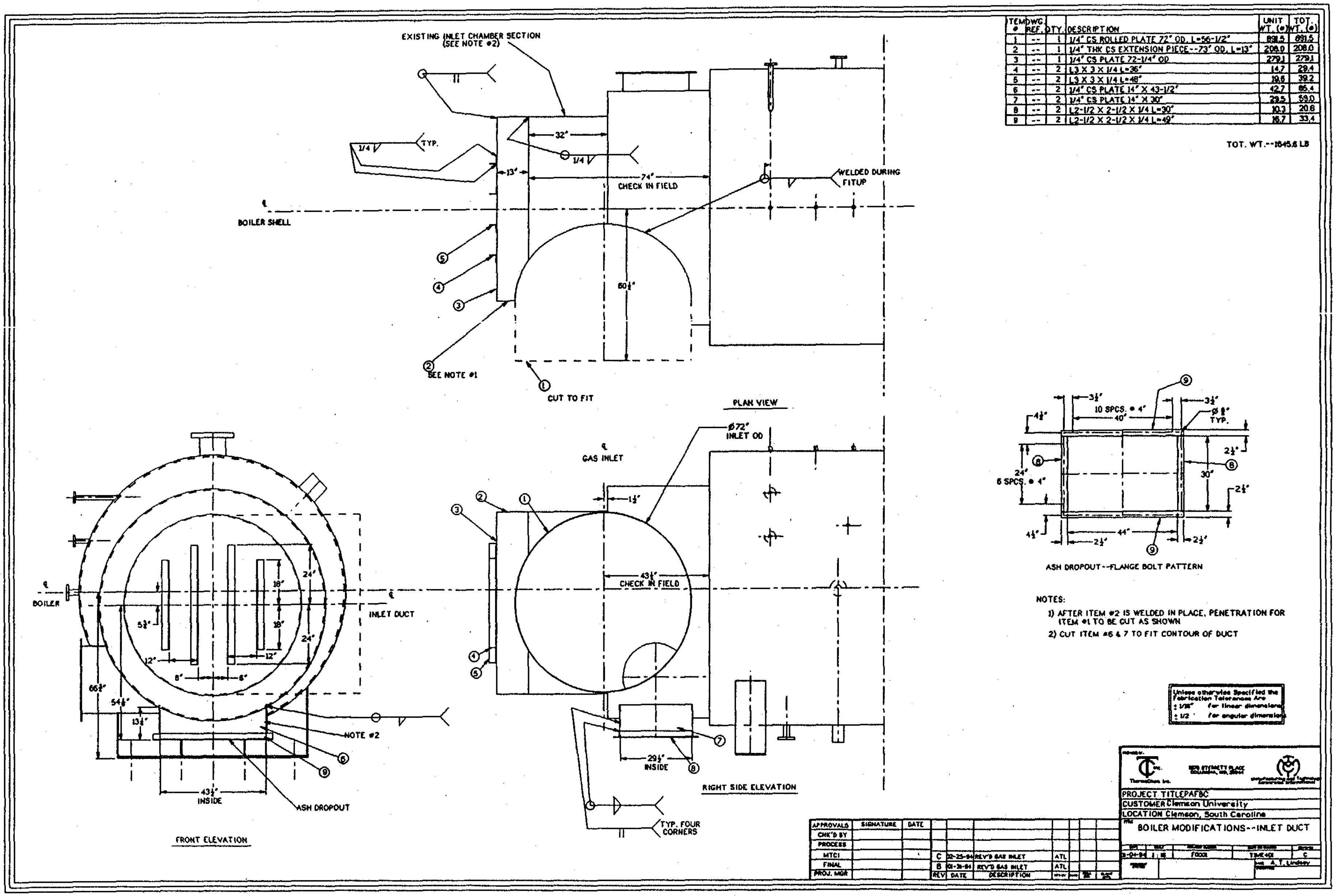


FIGURE 2-47: BOILER MODIFICATIONS - INLET DUCT

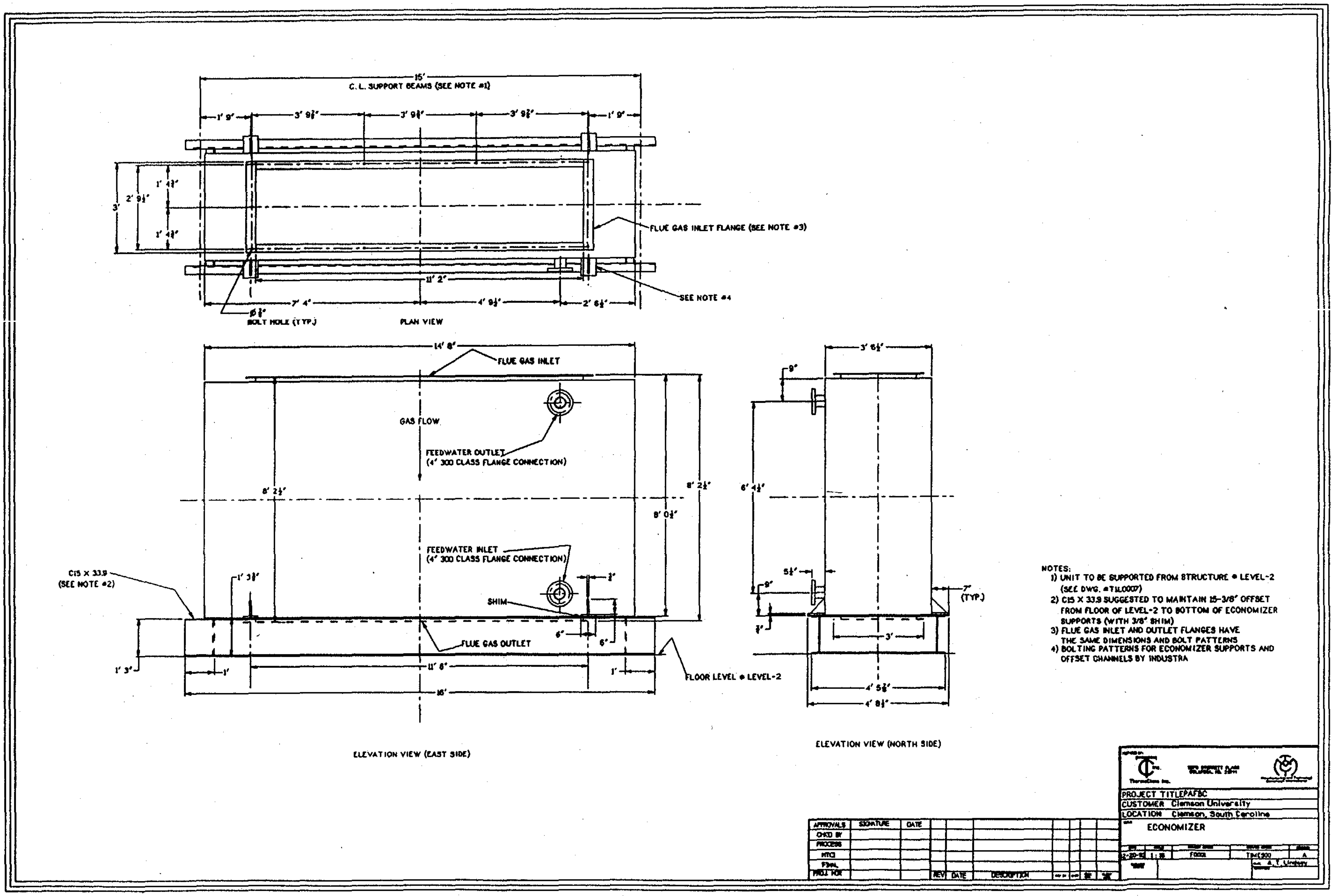
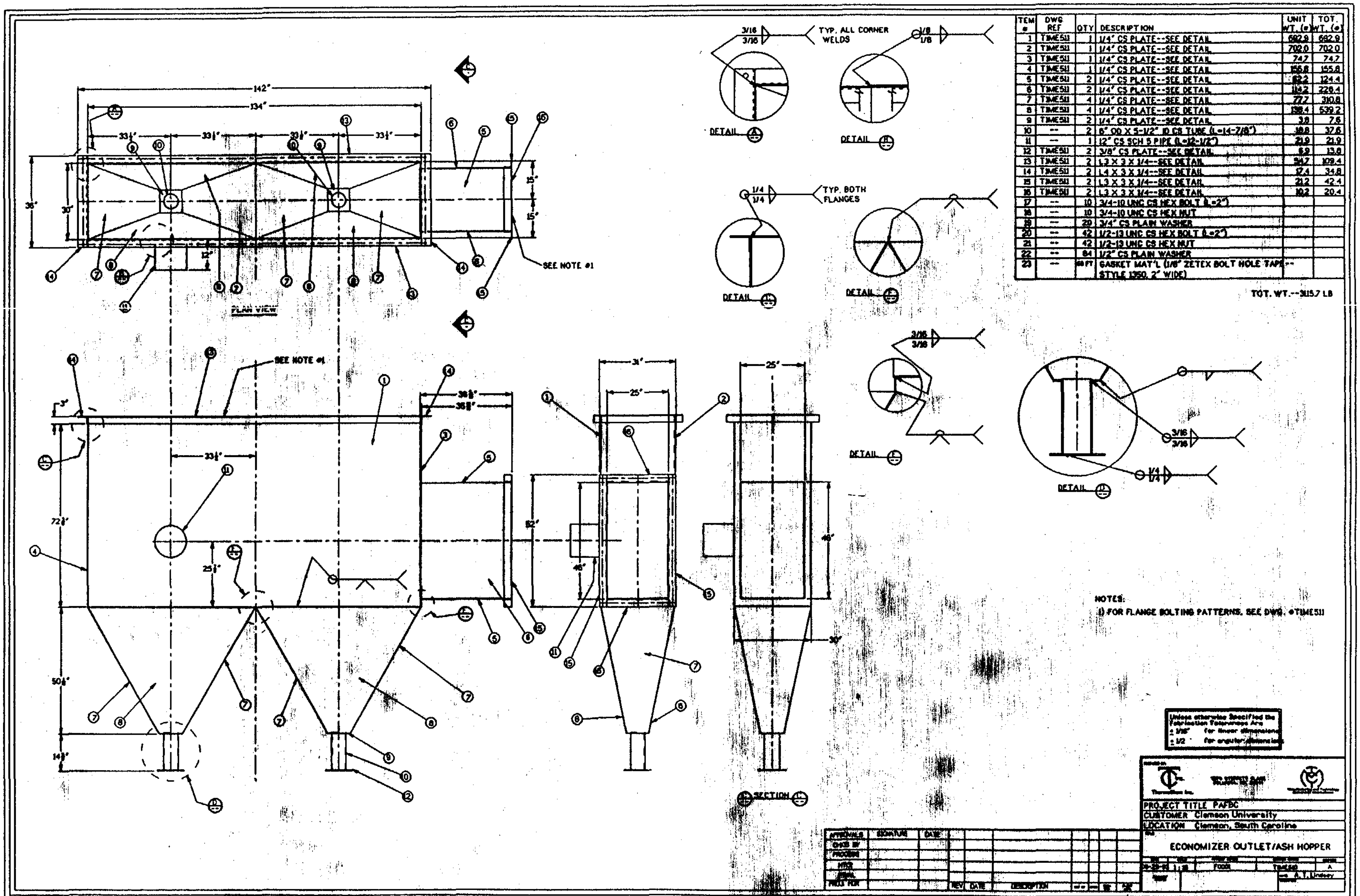


FIGURE 2-48: ECONOMIZER



ITEM #	DWG REF	QTY	DESCRIPTION	UNIT WT. (#)	TOT. WT. (#)
1	TIME511	1	1/4" CS PLATE--SEE DETAIL	682.9	682.9
2	TIME511	1	1/4" CS PLATE--SEE DETAIL	702.0	702.0
3	TIME511	1	1/4" CS PLATE--SEE DETAIL	74.7	74.7
4	TIME511	1	1/4" CS PLATE--SEE DETAIL	155.8	155.8
5	TIME511	2	1/4" CS PLATE--SEE DETAIL	62.2	124.4
6	TIME511	2	1/4" CS PLATE--SEE DETAIL	110.2	220.4
7	TIME511	4	1/4" CS PLATE--SEE DETAIL	77.7	310.8
8	TIME511	4	1/4" CS PLATE--SEE DETAIL	134.8	539.2
9	TIME511	2	1/4" CS PLATE--SEE DETAIL	3.8	7.6
10	--	2	6" OD X 5-1/2" ID CS TUBE (L=14-7/8")	18.8	37.6
11	--	1	12" CS SCH 5 PIPE (L=12-1/2")	21.9	21.9
12	TIME511	2	3/8" CS PLATE--SEE DETAIL	6.9	13.8
13	TIME511	2	L3 X 3 X 1/4--SEE DETAIL	54.7	109.4
14	TIME511	2	L4 X 3 X 1/4--SEE DETAIL	17.4	34.8
15	TIME511	2	L3 X 3 X 1/4--SEE DETAIL	21.2	42.4
16	TIME511	2	L3 X 3 X 1/4--SEE DETAIL	10.2	20.4
17	--	10	3/4-10 UNC CS HEX BOLT (L=2")		
18	--	10	3/4-10 UNC CS HEX NUT		
19	--	20	3/4" CS PLAIN WASHER		
20	--	42	1/2-13 UNC CS HEX BOLT (L=2")		
21	--	42	1/2-13 UNC CS HEX NUT		
22	--	84	1/2" CS PLAIN WASHER		
23	--	60 FT	GASKET MAT'L (1/8" ZETEX BOLT HOLE TAPE--STYLE 1950, 2" WIDE)		

FIGURE 2-49: ECONOMIZER OUTLET/ASH HOPPER

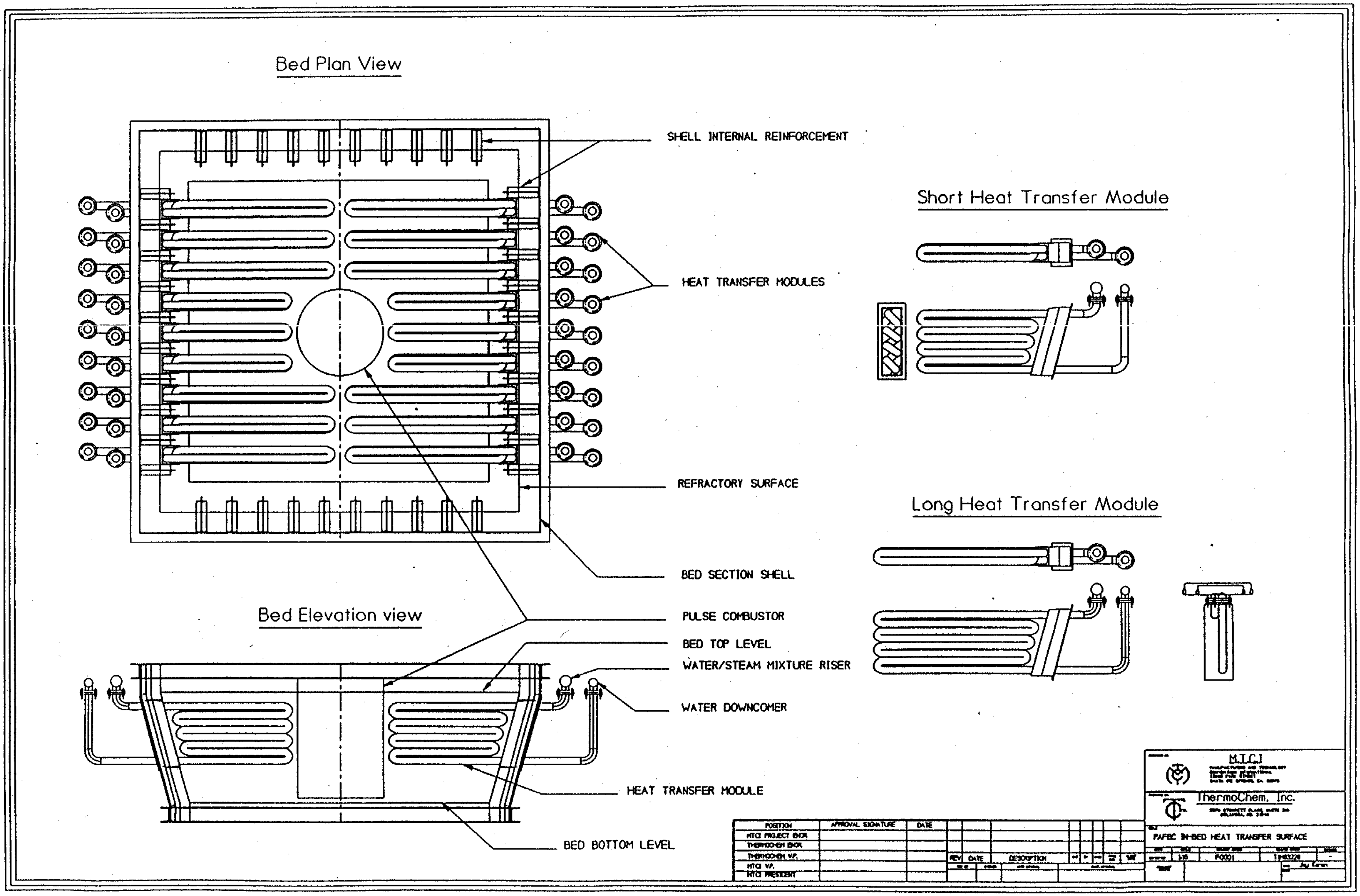


FIGURE 2-50: PAFBC IN-BED HEAT TRANSFER SURFACE

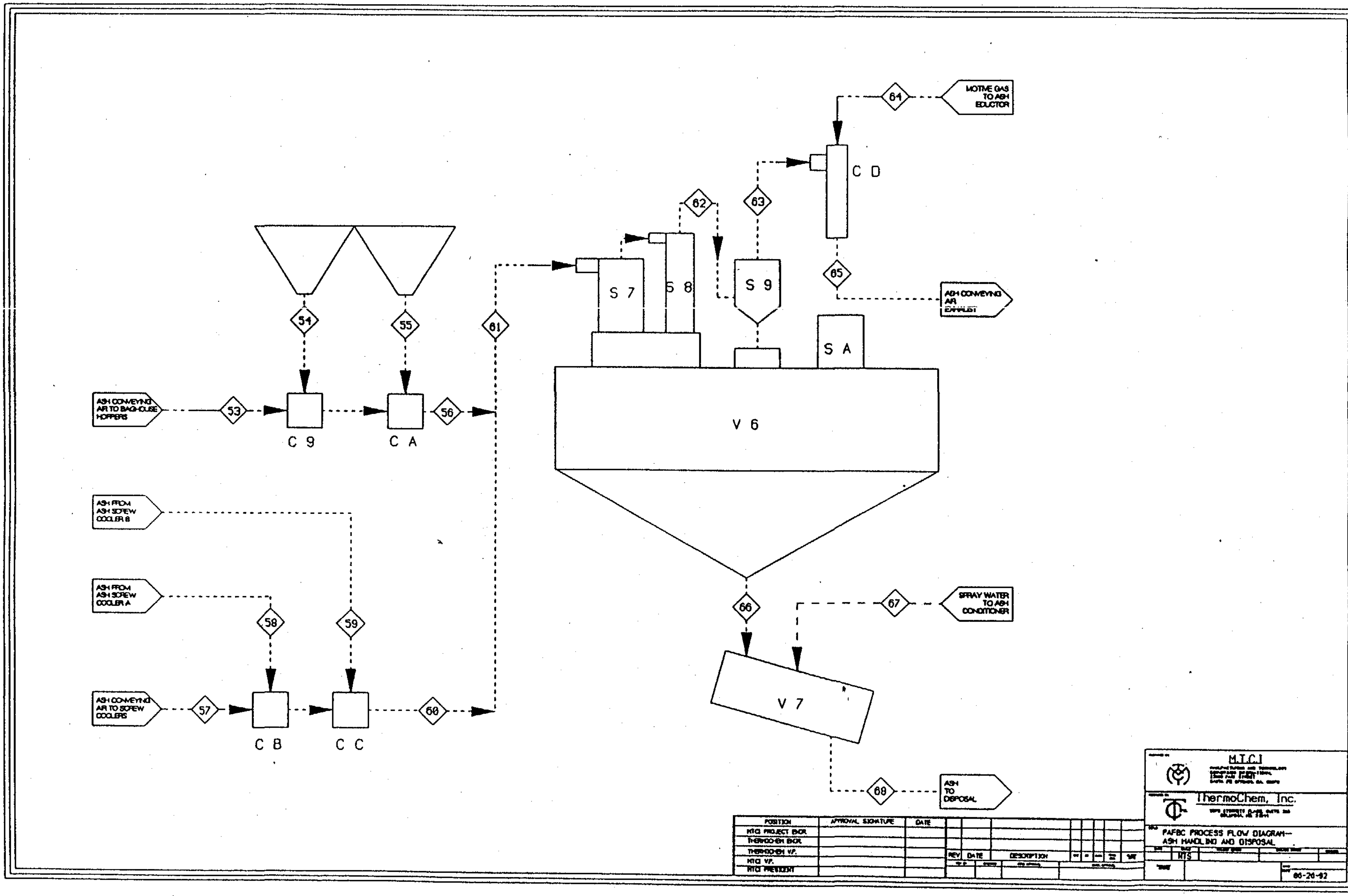


FIGURE 2-51: PAFBC PROCESS FLOW DIAGRAM - ASH HANDLING AND DISPOSAL

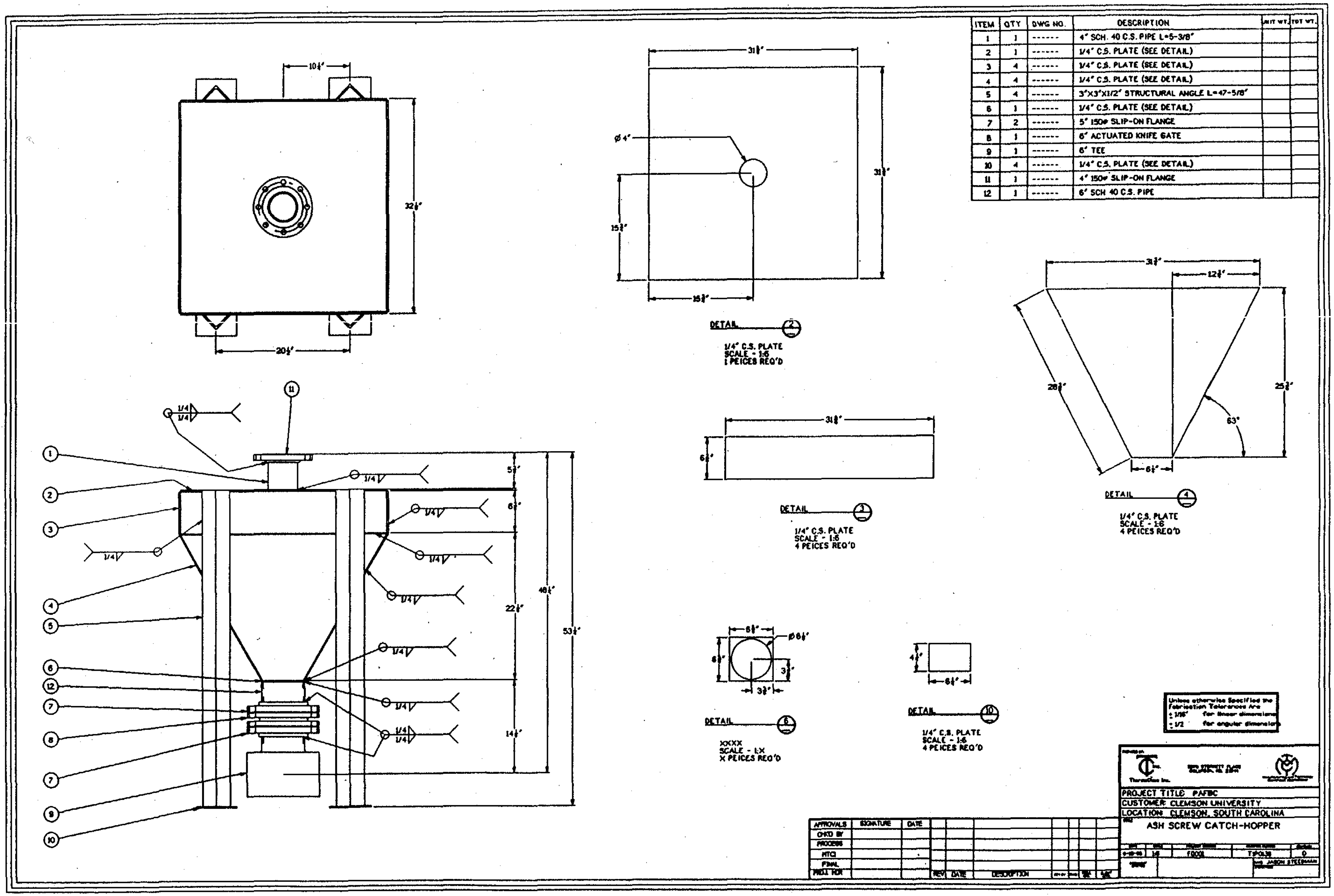


FIGURE 2-52: ASH SCREW CATCH HOPPER

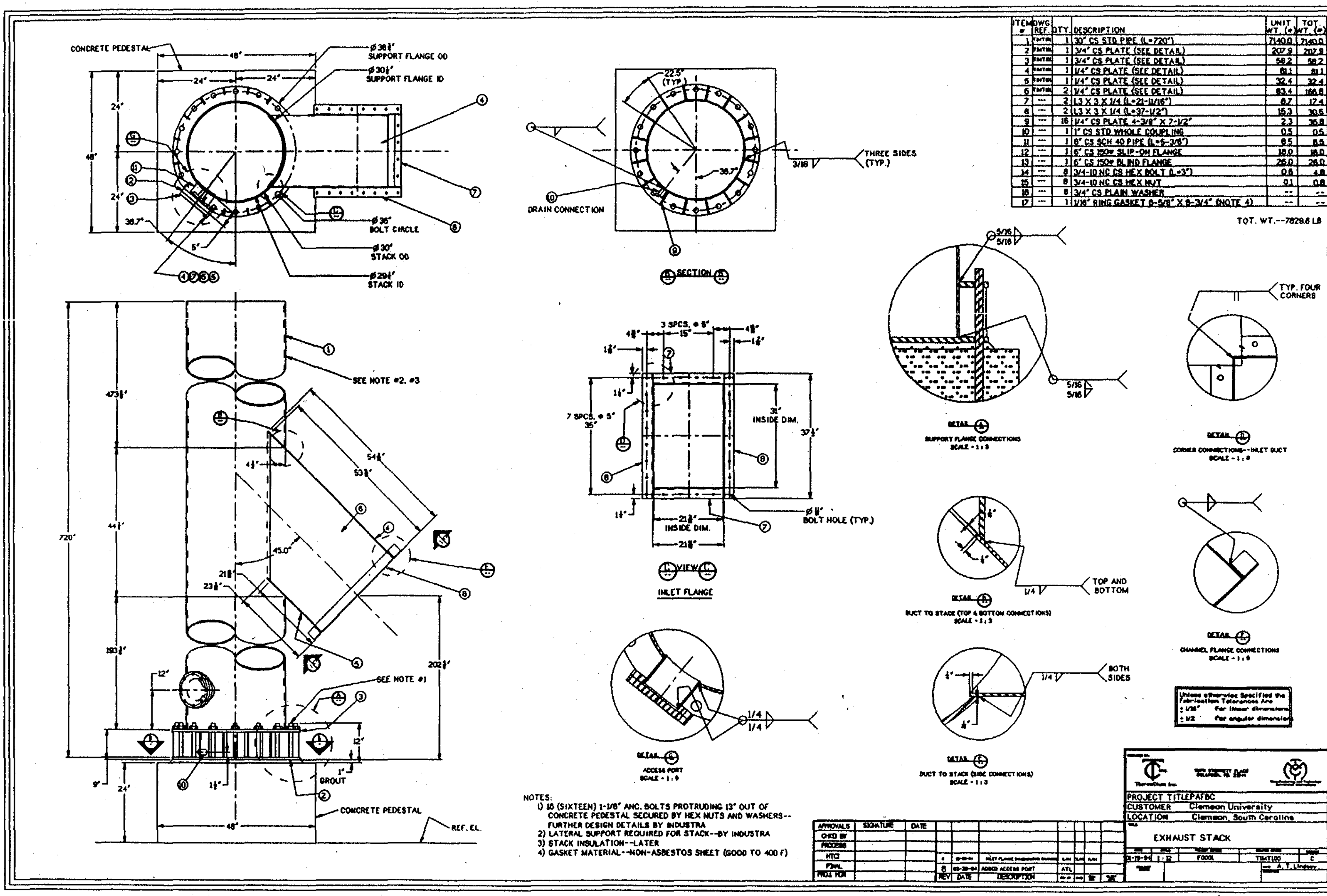


FIGURE 2-53: EXHAUST STACK

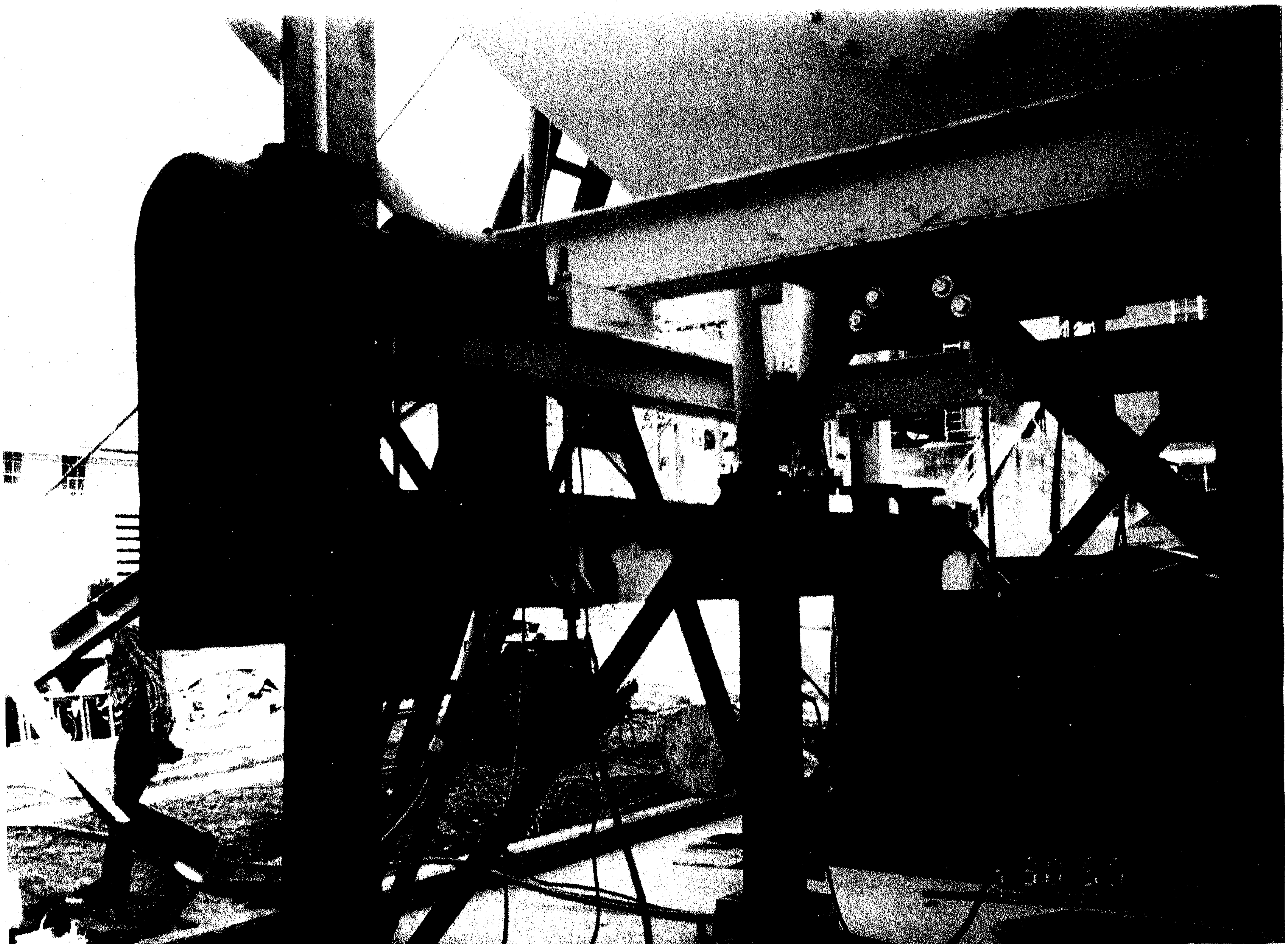


FIGURE 2-54: CLEMSON PAFBC UNIT - ASH-REMOVAL SYSTEM

TABLE 2-7:

ASH-HANDLING/RECYCLE SYSTEM

BAGHOUSE	
Gas inlet temperature, °F	350
Gas inlet density, lb/cu.ft.	0.05
Gas inlet flow, CFM	24,675.56
Inlet ash flow, lb/hr	1,600
Inlet ash flow, lb/cu.ft. of gas	1.08E-03
Inlet ash flow grains/cu.ft. of gas	7.6
Outlet ash flow grains/cu.ft. of gas	0.02
Outlet ash flow, lb/hr	4.23
Heat rejection, Btu/lb	67,500
BED ASH	
Ash flow, lb/hr	1,405
Inlet temperature to cooler, °F	1,550
Outlet temperature from cooler, °F	400
Heat rejection, Btu/hr	403,937.5
ASH-HANDLING SYSTEM	
Total ash flow, lb/hr	3,005
Ash handling air flow, lb/hr	4,000
Ash conveying method	vacuum
Air inlet pressure, psia	14.7
Pressure ratio across system	1.2
Pressure at vacuum ejector inlet, psia	12.25
Motive fluid	steam
Steam flow, lb/hr	2,000
Pressure at motive fluid inlet, psia	54.7

The fly ash that makes it through the Cyclone Separator (S-5) is collected at the boiler entrance, economizer outlet at two locations, and from the two baghouse hoppers. The fly ash is then removed from the hoppers by the ash-removal system.

The ash-removal system uses a Steam Eductor (C-D) to create the vacuum which transfer the solid ash to the Ash Silo (V-6). When the ash-removal system is in operation, each outlet port is drained one at a time for a fixed amount of time or until the hopper becomes empty. When the ash reaches the Ash Silo (V-6), the ash is sucked through two cyclones (S-7 and S-8) which are fitted with pneumatically assisted dump gates. The remaining entrained ash is then finally removed by the Ash Silo Filter Receiver (S-9). A steam silencer is fitted on the outlet of the Steam Eductor (C-D) in order to lower the noise generated while the ash system is in operation.

2.2.7 Plant Electrical and Control Systems

The PAFBC plant operated on a 480V, 3-phase, 60 Hz 1200A electrical supply. Clemson University supplied the electrical interface and transformer for the system. This supply was fed into an Allen Bradley Motor Control Center (see *Figure 2-55*). The MCC fed the power for all motors, plant lighting, PLC power, transformers and distribution panels. The single-line diagram (see *Figure 2-56*) and load schedule (see *Figure 2-57*) show all the corresponding motor loads for the plant and the power utilized by each system. The plant electrical wiring and layout was designed by ThermoChem, reviewed by Duke Fluor Daniels, and installed by Davis Electrical Contractors. The piping and instrumentation diagrams (see *Figures 2-58* through *2-62*), PLC layout (see *Figure 2-63*), and burner management panel (see *Figure 2-64*) were also designed by ThermoChem.

All control logic and control loops for the PLC (see *Figures 2-65* through *2-67*) and the burner management systems (see *Figure 2-68*).were developed by ThermoChem in accordance with safety standards established by various organizations. These organizations include Factory Mutual (FM), Underwriters' Laboratories (UL), The National Fire Protection Association (NFPA), and Industrial Risk Insurers (IRI).

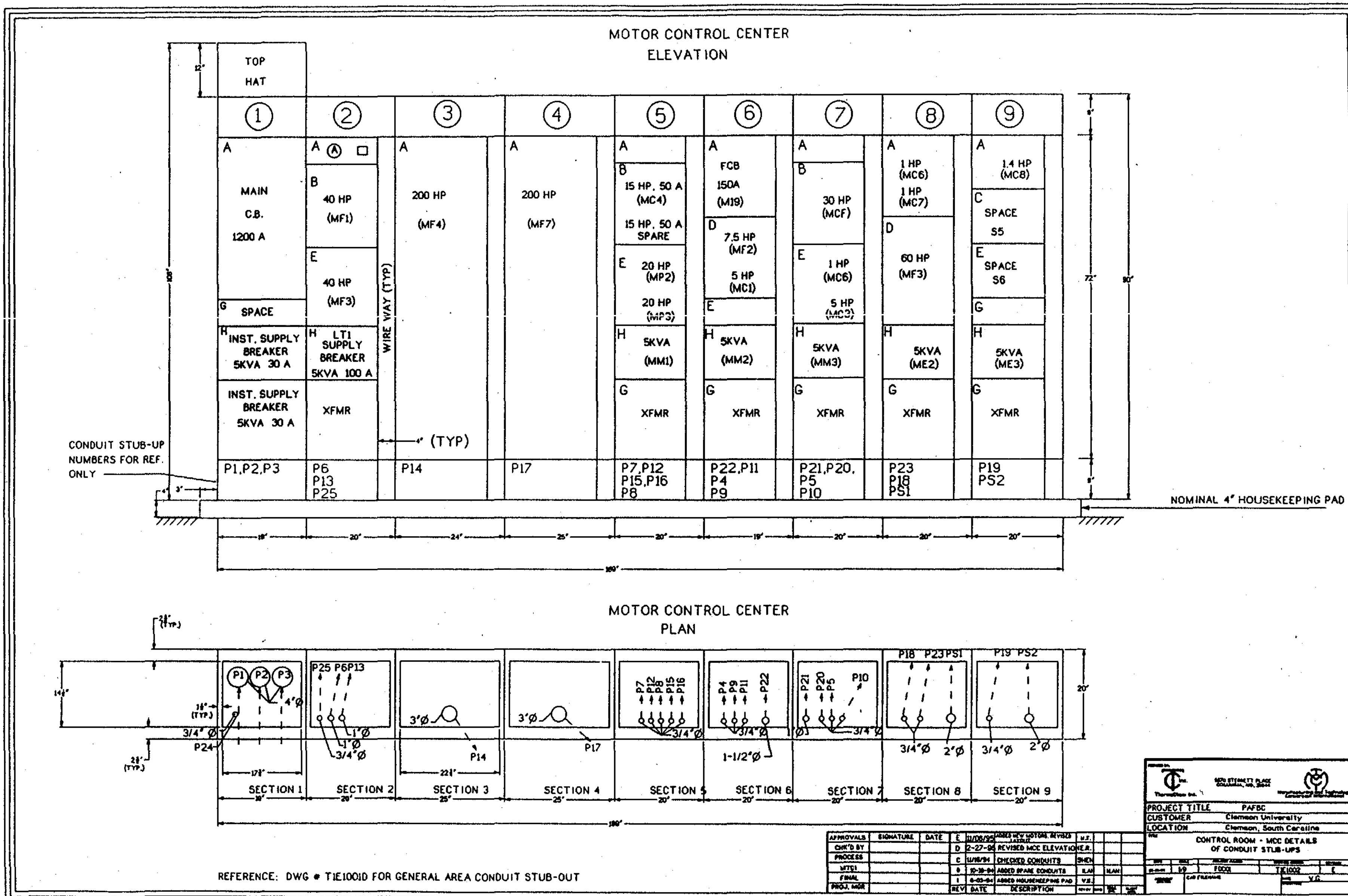


FIGURE 2-55: CONTROL ROOM - MCC DETAILS OF CONDUIT STUB-UPS

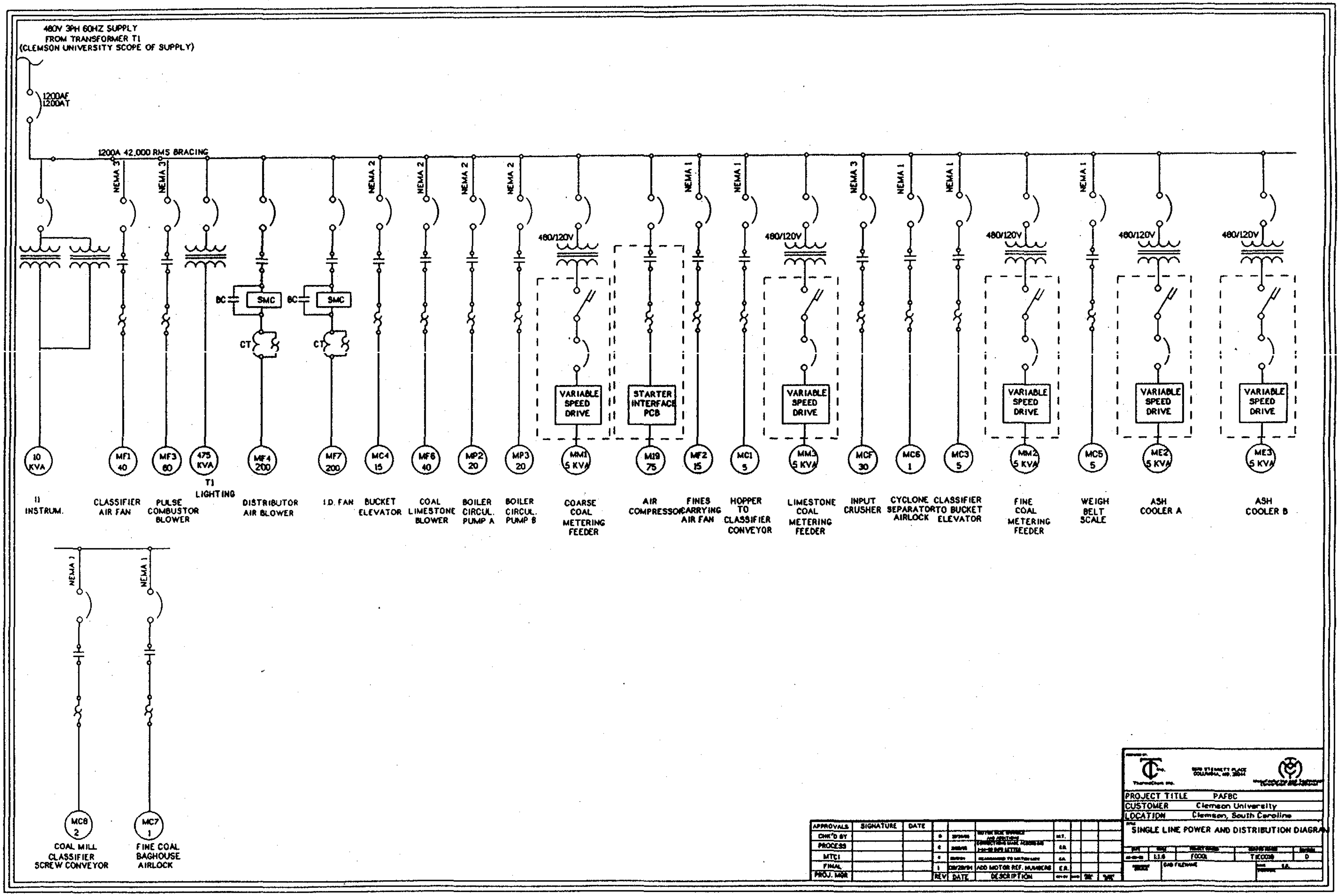


FIGURE 2-56: SINGLE-LINE POWER AND DISTRIBUTION DIAGRAM

LOAD SCHEDULE

ITEM NO.	NEW REF #	OLD REF #	DESCRIPTION	HP	KVA	FLC IN AMPS	STARTING TYPE	CABLE SIZE	CONDUIT SIZE	CONDUIT #
1	MC1	M1	HOPPER TO CLASSIFIER CONVEYOR	5	3.6	7.6	FVNR	3# 12+1# 12 G	3/4"	P4
2	MC3	M2	CLASSIFIER TO B. E. CONVEYOR	5	3.6	7.6	FVNR	3# 12+1# 12 G	3/4"	P5
3	MF1	M3	CLASSIFIER AIR FAN	40	25	52	FVNR	3# 6+1# 10 G	1"	P6
4	MC4	M4	BUCKET ELEVATOR	15	10	21	FVNR	3# 10+1#12 G	3/4"	P7
5	MM1	M5	COARSE COAL METERING FEEDER	2 (DC)	2.4	5	5 KVA 480/120V TR	2#12+1#12 G	3/4"	P8
6	MM2	M6	FINE COAL METERING FEEDER	2 (DC)	2.4	5	5 KVA 480/120V TR	2#12+1#12 G	3/4"	P9
7	MM3	M7	LIMESTONE METERING FEEDER	2 (DC)	2.4	5	5 KVA 480/120V TR	2#12+1#12 G	3/4"	P10
8	MF2	M8	FINES CARRYING AIR FAN	15	10	21	FVNR	3#10+1#12 G	3/4"	P11
9	MF6	M9	COAL LIMESTONE BLOWER	40	25	52	FVNR	3#10+1#12 G	3/4"	P12
10	MF3	M10	PULSE COMBUSTOR BLOWER	60	37	77	FVNR	3#6+1#10 G	1"	P13
11	MF4	M11	DISTRIBUTOR AIR BLOWER (FLUID BED ID FAN)	200	115	240	SOFT START	3#350 MCM+1#4 G	3"	P14
12	MP2	M12	BOILER CIRCULATION PUMP A	20	13	27	FVNR	3#10+1#12 G	3/4"	P15
13	MP3	M13	BOILER CIRCULATION PUMP B	20	13	27	FVNR	3#10G+1#12 G	3/4"	P16
14	MF7	M14	I.D. FAN	200	115	240	SOFT START	3#350MCM+ 1#4 G	3"	P17
15	ME2	M15	ASH COOLER A (240V. 8.5)	2 (DC)	2	5	5 KVA 480/120V TR	2#12+1#12 G	3/4"	P18
16	ME3	M16	ASH COOLER B (240V. 8.5)	2 (DC)	2	5	5 KVA 480/120V TR	2#12+1#12 G	3/4"	P19
17	MC6	M17	CYCLONE SEPARATOR AIRLOCK	1	1	2	FVNR	3#12+1#12 G	3/4"	P20
18	MCF	M18	INPUT CRUSHER	30	19.2	40	FVNR	3#8+1#10 G	1"	P21
19	M19	M19	AIR COMPRESSOR (LOCAL FIELD MOUNTED)	75	46	96	SOFT START	3#1+1#6 G	1 1/2"	P22
20	MC5	M20	WEIGH BELT SCALE	5	3.6	7.6	FVNR	3#12+1#12 G	3/4"	P23
21	I1	I1	INSTRUMENT SUPPLY		5	11	5 KVA 480/120V TR	2#10+1#12 G (THE CONDUIT WILL GO ABOVE THE CEILING)	3/4"	P24
22	LT1	L1	INSTRUMENT SUPPLY (2)		5	11	5 KVA 480/120V TR	2#10+1#12 G	3/4"	P25
23	L1		INCOMING CABLE				INCOMING	3#750MCM + 1 GROUND 4/0	4"	P1
24	L2		INCOMING CABLE				INCOMING	3#750MCM + 1 GROUND 4/0	4"	P2
25	L3		INCOMING CABLE				INCOMING	3#750MCM + 1 GROUND 4/0	4"	P3
27	S1		TRANSFORMER FLOOR MOUNTED (PL-1 PANEL)		37.5	60	37.5 KVA 480/120V TR	2#6+1#10 G (THE CONDUIT WILL BE ABOVE THE CEILING)	2"	PS1
28	S2	MC8	COAL MILL TO CLASSIFIER SCREW CONVEYOR	2	1.4	3	FVNR	2#12+1#12 G	2"	PS1
28	S3	MC7	FINE COAL BAGHOUSE AIRLOCK	1	1	2	FVNR	3#12+1#12 G	3/4"	P20
28	S4		SPACE						2"	PS2
28	S5		SPACE							—
28	S6		SPACE							—
28	S7		SPACE							—
			ROUNDED OFF TOTALS	744	501	1028				

NOTE
The power conduits size is based on Table 38, chapter 9 of NEC for using THHN cable, except, the incoming cables which are of XHHW type.

PROJECT TITLE PAFBC
CUSTOMER Clemson University
LOCATION Clemson, South Carolina
Power Conduit Sizes and Lead Schedule

FIGURE 2-57: POWER CONDUIT SIZES AND LEAD SCHEDULE

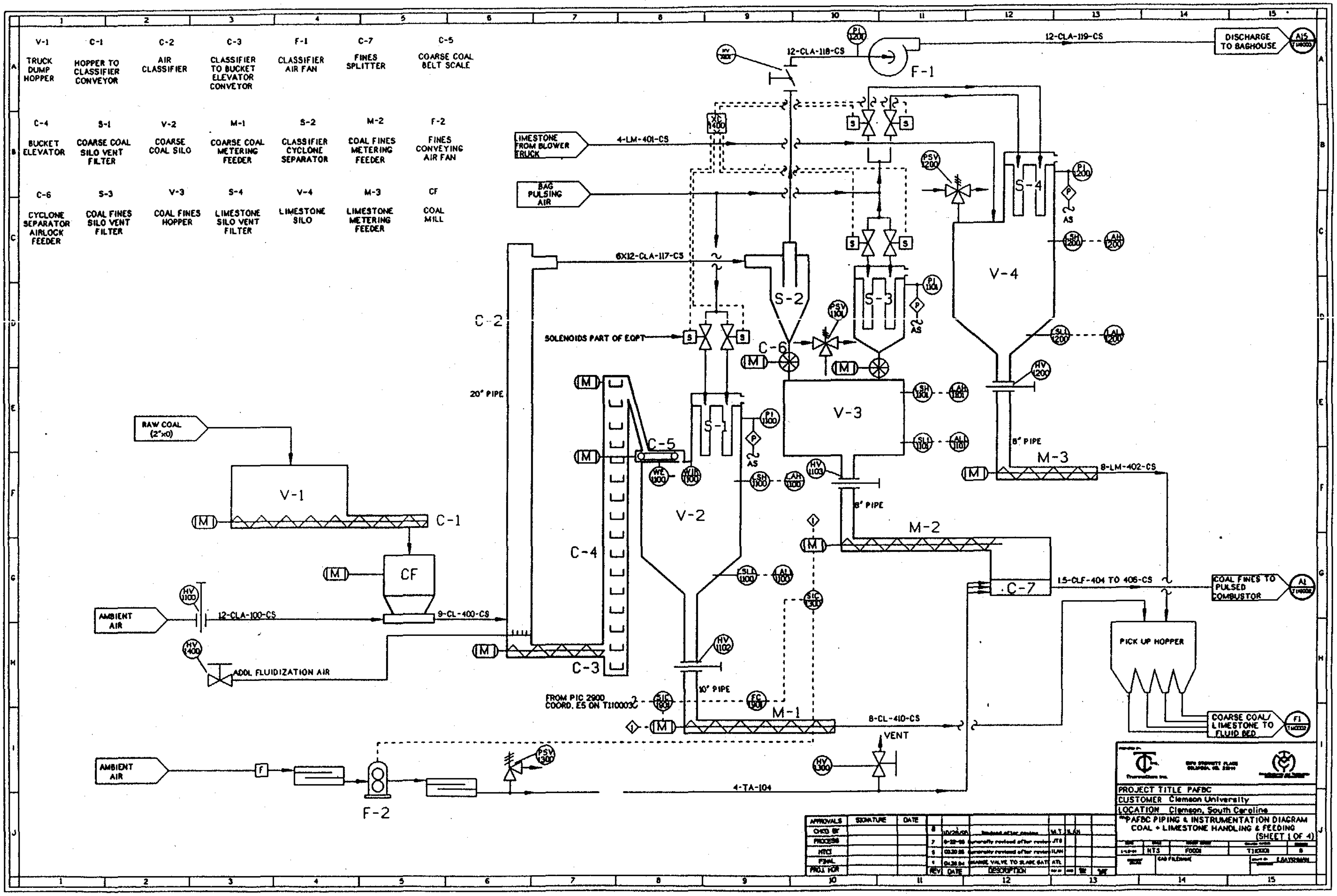


FIGURE 2-58: PAFBC PIPING & INSTRUMENTATION DIAGRAM - COAL + LIMESTONE HANDLING AND FEEDING

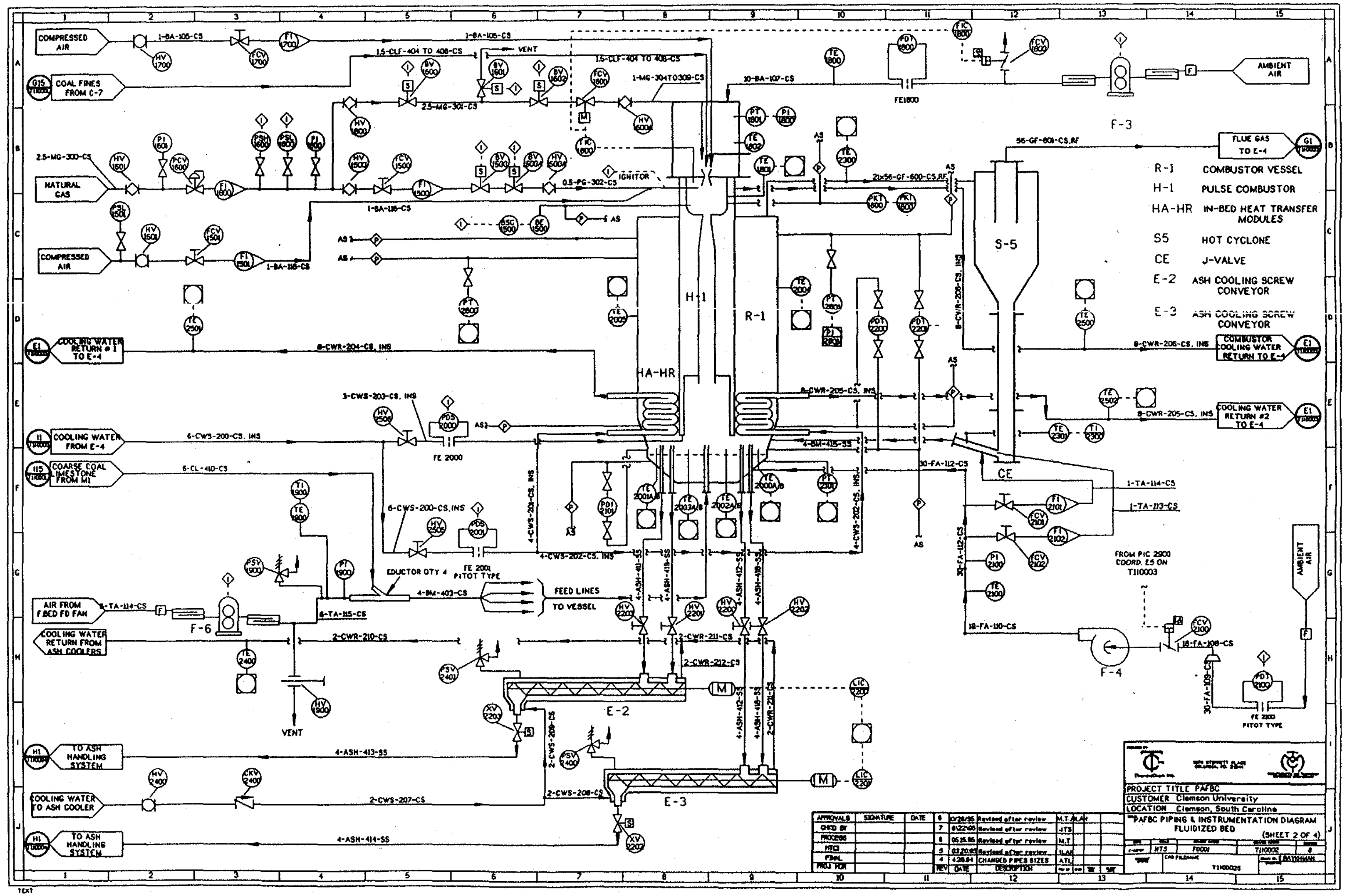


FIGURE 2-59: PAFBC PIPING & INSTRUMENTATION DIAGRAM - FLUIDIZED BED

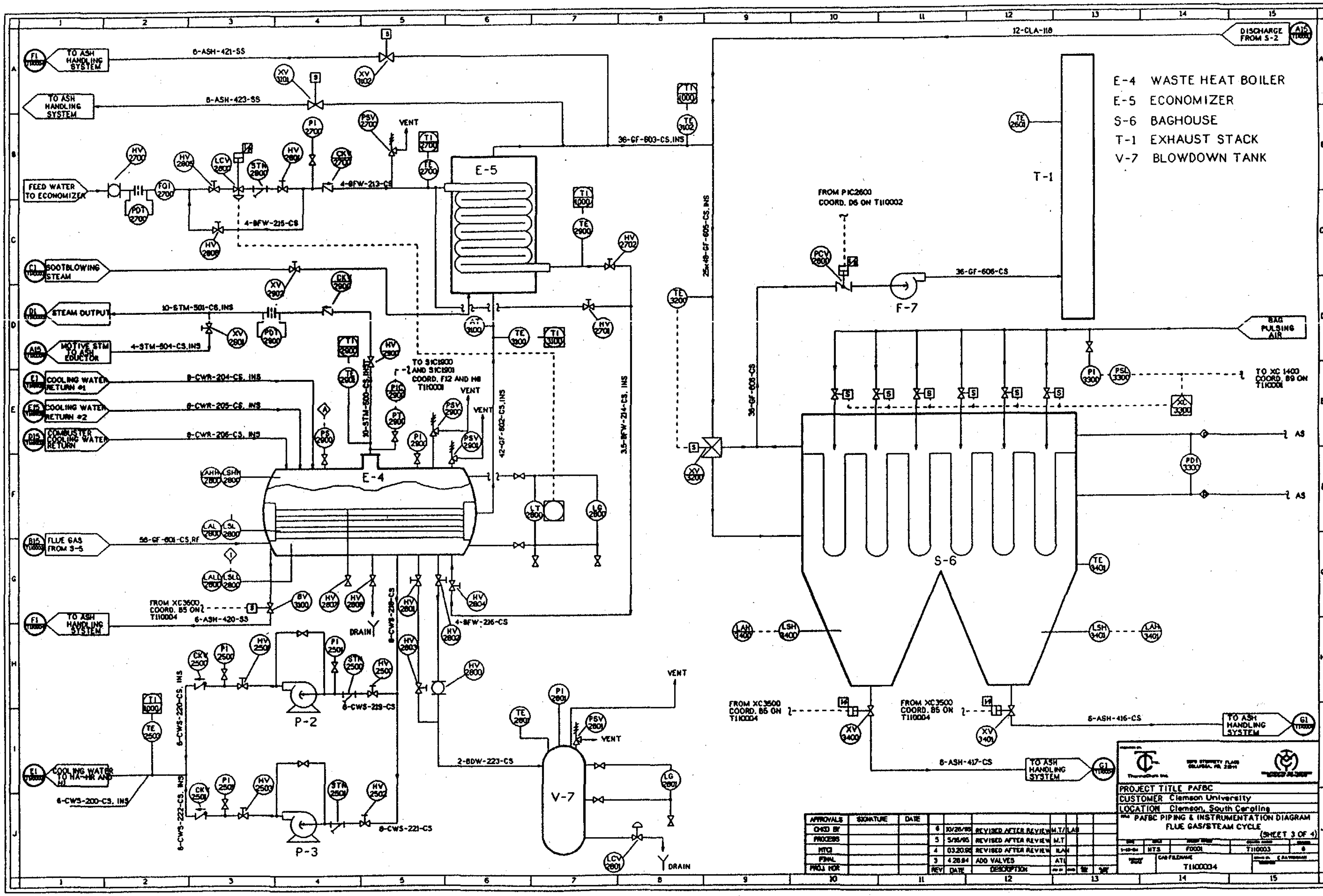


FIGURE 2-60: PAFBC PIPING & INSTRUMENTATION DIAGRAM - FLUE GAS/STEAM CYCLE

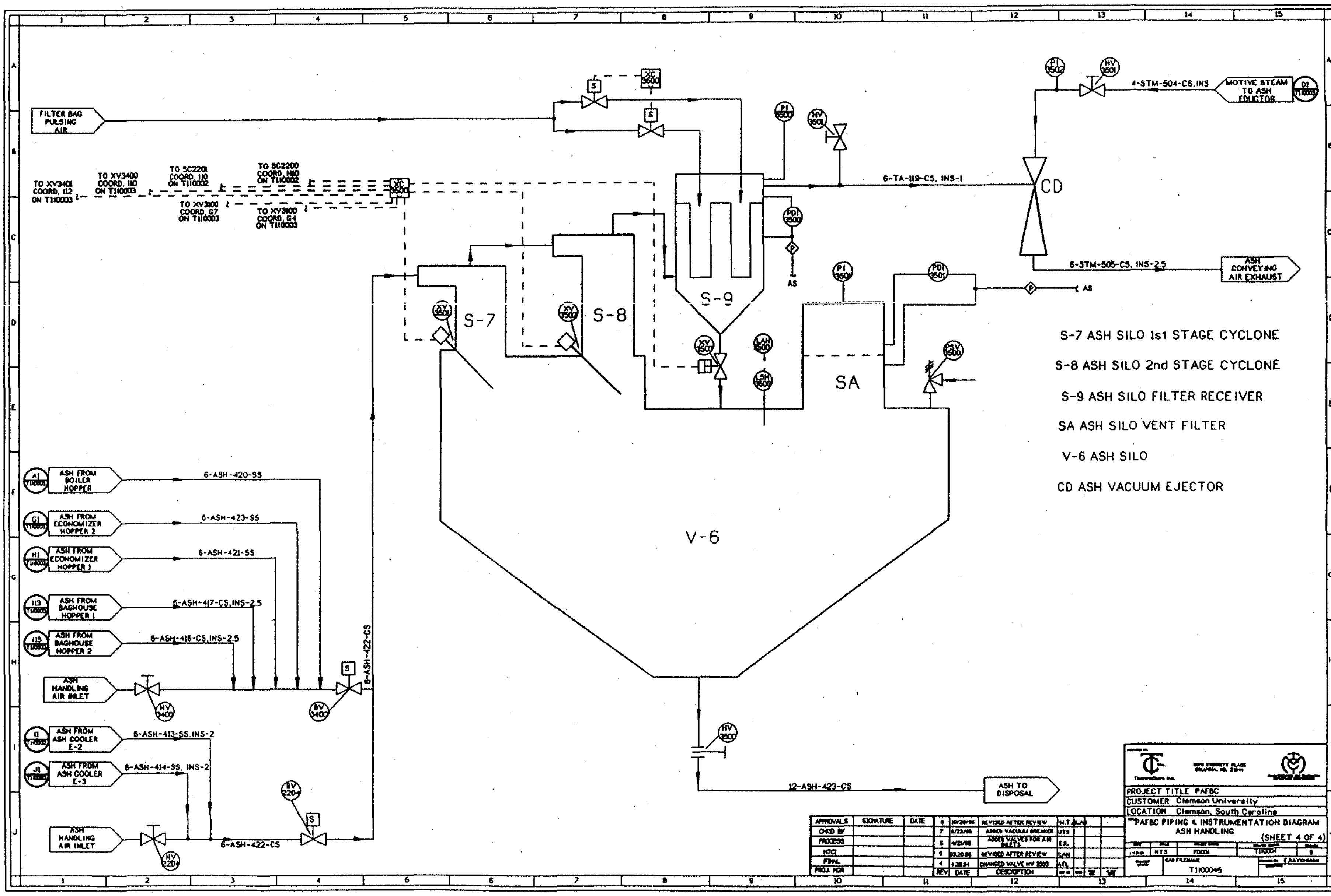


FIGURE 2-61: PAFBC PIPING & INSTRUMENTATION DIAGRAM - ASH HANDLING

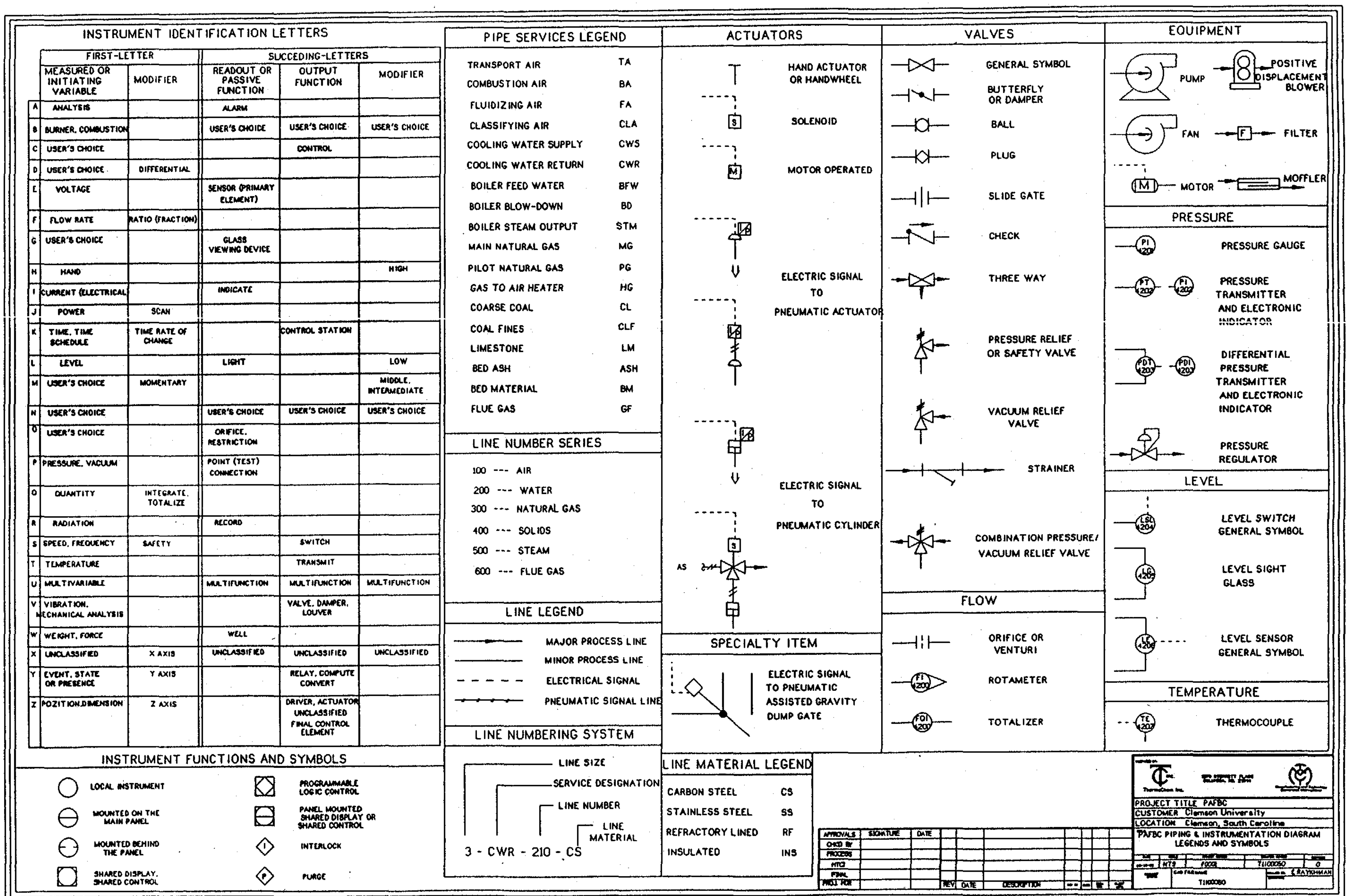
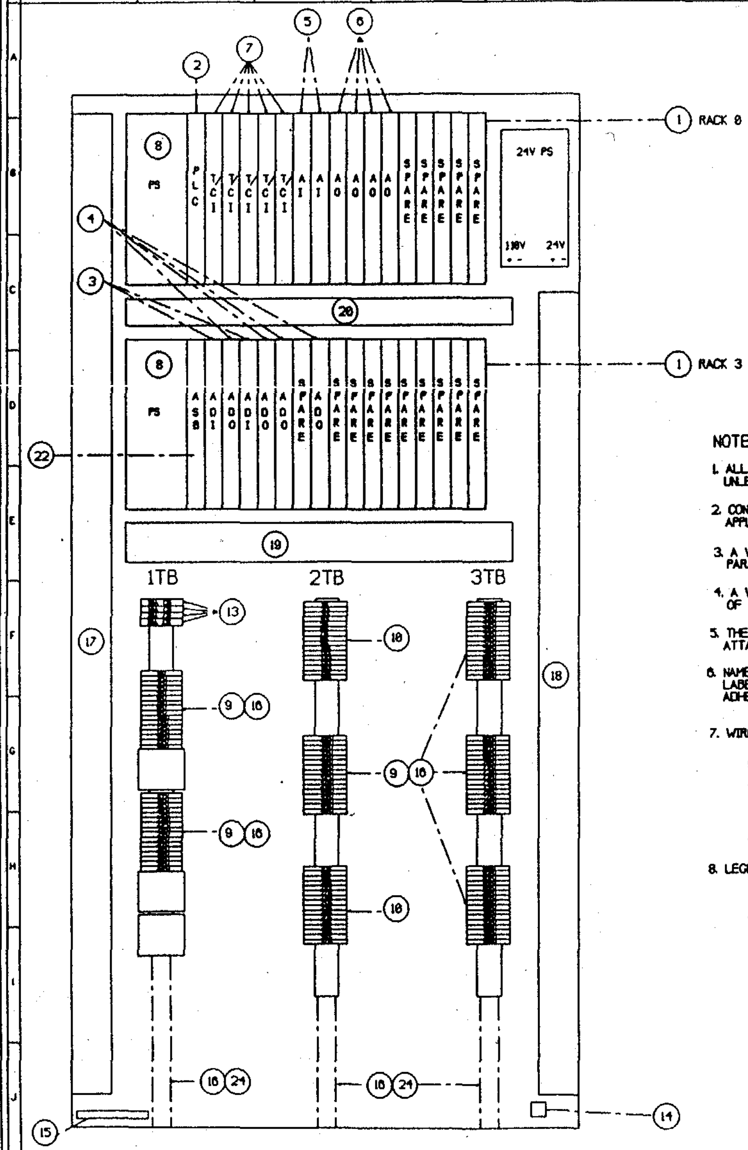


FIGURE 2-62: PAFBC PIPING & INSTRUMENTATION DIAGRAM - LEGENDS AND SYMBOLS



NOTES:

1. ALL CABINET WIRING IS #14 AWG, TYPE THWN, 600V AC, 300V DC UNLESS OTHERWISE NOTED
2. CONTROL PANEL TO BE MANUFACTURED TO NFPA 70, 79 & UL-508 APPLICABLE STANDARDS.
3. A WARNING LABEL SHALL BE PLACED NEXT TO FUSES PER UL-508 PARAGRAPH 66.18
4. A WARNING LABEL SHALL BE PLACED ON THE FRONT DOOR WARNING OF HIGH VOLTAGE
5. THE ENCLOSURE AND PANEL ARE TO BE BONDED TO A GROUND BLOCK ATTACHED TO THE ENCLOSURE
6. NAMEPLATES IDENTIFYING COMPONENTS INSIDE THE PANEL TO BE LABELED AS SHOWN WITH 3-PLY LAMINATED WHITE FACE, BLACK CORE, ADHESIVE MOUNTED WITH 3/16" HIGH LETTERING
7. WIRE INSULATION TO BE COLOR CODED AS FOLLOWS:

BLACK	=120V AC POWER
RED	=120V AC CONTROL
WHITE	=AC NEUTRAL
WHITE	=+24V DC (INSULATED SIGNAL CABLE)
BLACK	=-24V DC (INSULATED SIGNAL CABLE)

8. LEGEND:

AI	=ANALOG INPUT
AO	=ANALOG OUTPUT
DI	=DESCRETE INPUT
DO	=DISCRETE OUTPUT
T/CI	=THERMOCOUPLE INPUT

BILL OF MATERIAL

ITEM	QTY	DESCRIPTION
1	2	CHASSIS, 16 SLOT, ALLEN-BRADLEY CAT. No. 1771-A4B
2	1	PROCESSOR, ALLEN-BRADLEY PLC-5 CAT. No.
3	2	DICRETE INPUT MODULE, ALLEN-BRADLEY CAT. No. 1771-IAD
4	4	DICRETE OUTPUT MODULE, ALLEN-BRADLEY CAT. No. 1771-OAD
5	2	ANALOG INPUT MODULE, ALLEN-BRADLEY CAT. No. 1771-IFE
6	4	ANALOG OUTPUT MODULE, ALLEN-BRADLEY CAT. No. 1771-OFE
7	5	THERMOCOUPLE MODULE,ALLEN-BRADLEY CAT. No.1771-IXE
8	2	POWER SUPPLY, ALLEN-BRADLEY CAT. No.1771-P7
9	96	FUSED TERMINAL BLOCK, ALLEN-BRADLEY CAT. No.1492-H6
10	32	ISOLATION SWITCH BLOCK, ALLEN-BRADLEY CAT. No. 1492-H7
11	48	MULTI-CIRCUIT TERMINAL BLOCK, ALLEN-BRADLEY CAT. No. 1492-WTF3
12	118	STANDARD TERMINAL BLOCK, ALLEN-BRADLEY CAT. No. 1492-W4
13	4	CIRCUIT BREAKER, ALLEN-BRADLEY CAT. No. 1492-CH100
14	1	GROUND BLOCK, #8 AWG CONNECTION LUG
15	1	GROUND BUS BAR, 10H,
16	96	FUSE 2A, BUSSMAN CAT.No. FNM-2
17	1	WIRING DUCT 3"x4" 72" LONG WITH COVER
18	1	WIRING DUCT 3"x4" 59" LONG WITH COVER
19	1	WIRING DUCT 3"x4" 29" LONG WITH COVER
20	1	WIRING DUCT 2"x4" 29" LONG WITH COVER
21	3	DIN MOUNTING RAILS, ALLEN-BRADLEY CAT. No. 1492-DR5
22	1	REMOTE I/O ADAPTER, ALLEN-BRADLEY CAT No. 1771-ASB
23	1	ALLEN-BRADLEY MCC SECTION 90"H x 40"W x 15"D
24	6	END ANCHOR, ALLEN-BRADLEY CAT. No.1492-H47

REFERENCE DRAWINGS:

PROJECT TITLE PAFBC
CUSTOMER CLEMSON UNIVERSITY
LOCATION CLEMSON, SOUTH CAROLINA
PLC
PANEL LAYOUT

FIGURE 2-63: PLC PANEL LAYOUT

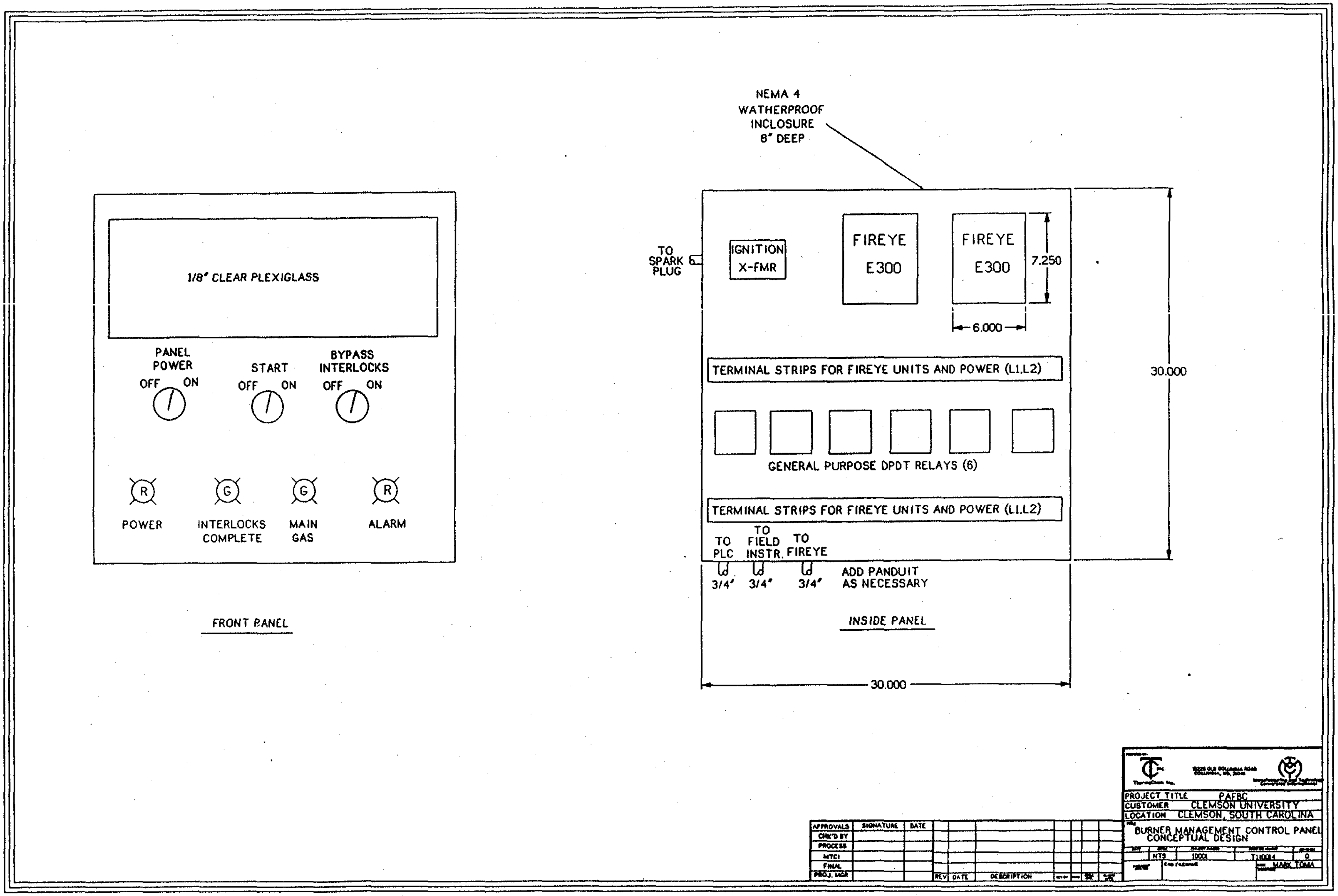


FIGURE 2-64: BURNER MANAGEMENT CONTROL PANEL - CONCEPTUAL DESIGN

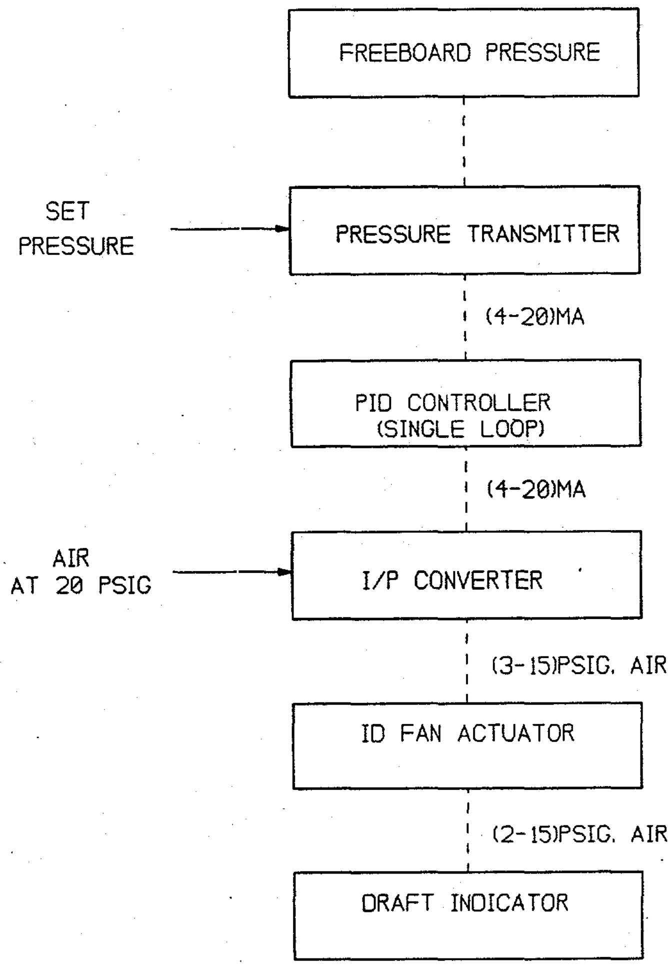


FIGURE 2-65: PAFBC INSTRUMENTATION FUNCTIONAL DIAGRAM - DRAFT CONTROL

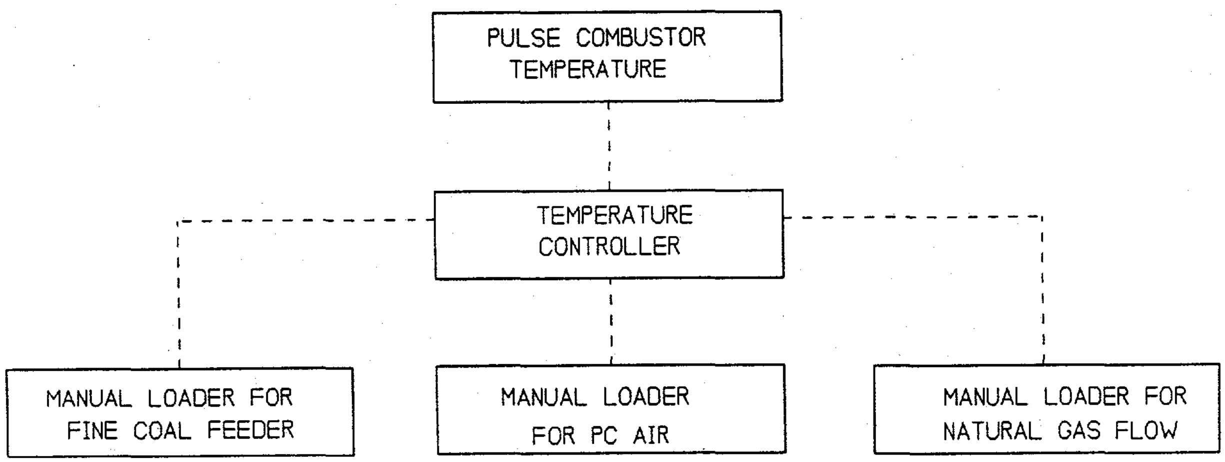


FIGURE 2-66: PAFBC INSTRUMENTATION FUNCTIONAL DIAGRAM - PULSE COMBUSTOR TEMPERATURE CONTROL

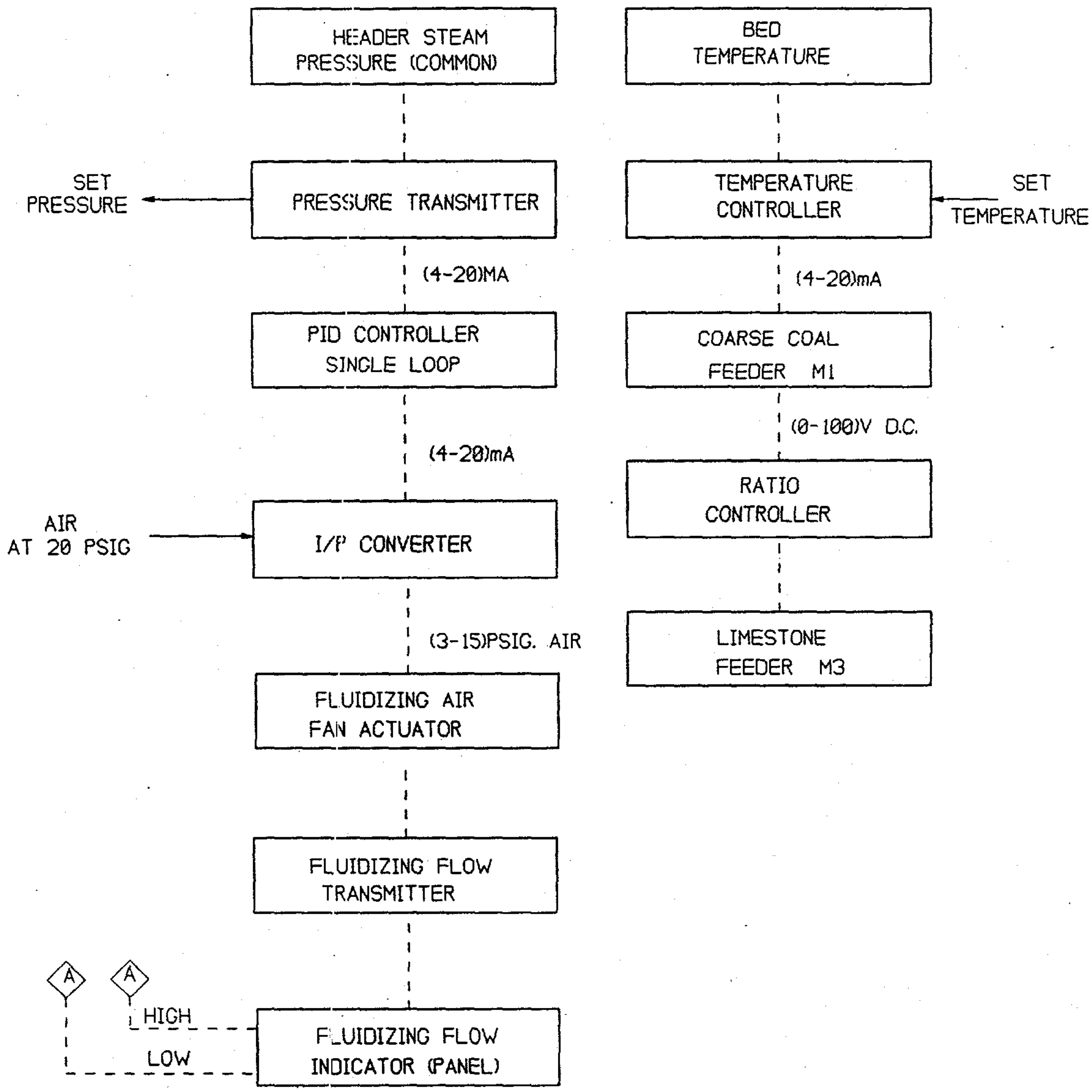


FIGURE 2-67: PAFBC INSTRUMENTATION FUNCTIONAL DIAGRAM - COMBUSTION CONTROL

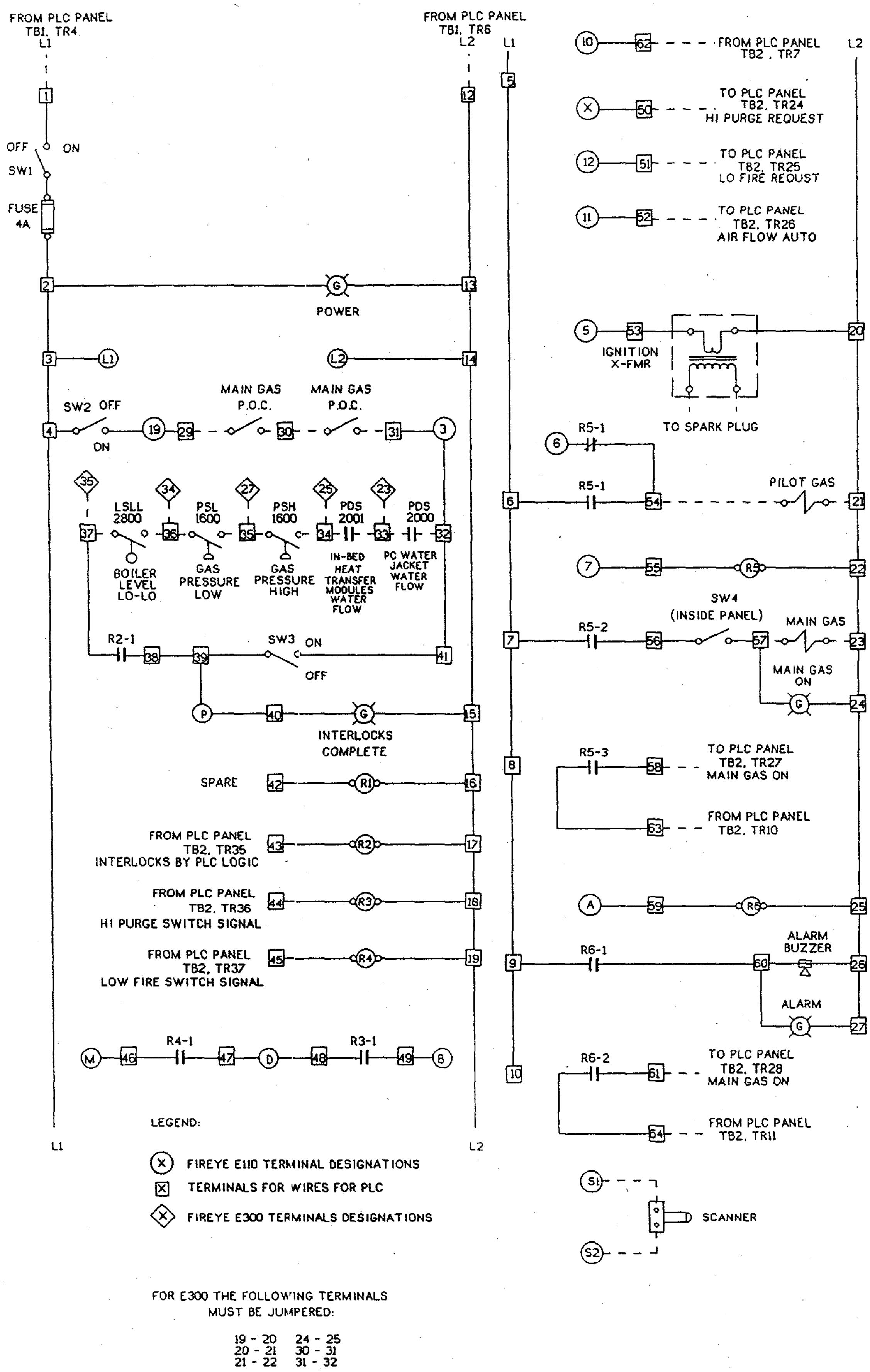


FIGURE 2-68: BURNER MANAGEMENT CONTROL PANEL - ELEMENTARY LOGIC DIAGRAM

Control System

The PAFBC plant is controlled by an Allen Bradley PLC 5/11 connected to a Pentium-based personal computer running Windows '95 as its operating system. RS-232C protocol for on-line/off-line programming, systems monitoring, setpoint control and data acquisition. Another computer and a 33.6 Kbps modem are connected to the PLC using Allen Bradley's Data Highway Plus protocol for remote systems monitoring and data acquisition. The software system was ICOM's Wintelligent Series which is a Windows-based package consisting of five separate modules. The first module is WinLogic 5 which is used for on-line/off-line ladder logic programming of the PLC. The second module is WinLynx which is the underlying communications package that coordinates the Dynamic Data Exchange (DDE) communications to the PLC. WinTrend is the data acquisition package that allows the computer system to store any assigned data points at the user's selected sampling frequency and saves the information in any one of five formats or allows direct printing of the selected data. The last two packages are WinView and View Runtime. These are the MMI packages that allows systems monitoring and parameter changing by the user. WinView is the development software used to design the MMI, whereas the latter is the compiled runtime version. It is the duty of the PLC to monitor all critical functions of the plant and automatically compensate for minor upset conditions, as well as the initiation of a full plant shut down should an emergency condition arise. With the interface computer, an operator has the ability to both monitor plant parameters, alter set points, or even run the system manually.

The PLC looks at all of the important instruments and adjusts setpoints in order to better control the boiler operation. The man machine interface is done through a 80486-based PC, running Rockwell's WinView Runtime. The ladder logic for the PLC was written in Rockwell Software's WinLogic 5. Copies of the ladder logic exit both in the PLC memory as well as on the hard disk of the interface computer. The PLC is programmed to control all control loops, as well as the timing and control sequences for both the Ash Baghouse and the Ash Removal System. Should an emergency condition arise, the PLC is capable of shutting off all fuel and tripping fan interlocks.

Coal/Limestone Feeding

Limestone Flow Rate. The amount of limestone fed to the bed is determined by the value of the Calcium/Sulfur ratio setpoint. The setpoint is used to calculate the correct amount of limestone for the PAFBC unit based on the coal flow rate. Since the ratio is defined as the molar ratio of calcium to sulfur, the percentage of sulfur in the coal must be known in order for the setpoint to function properly. A normal safe ratio is approximately 3.

When the limestone is fed into the hot bed, the Limestone (calcium carbonate) endothermally decays into CaO and CO_2 in this fashion:

$$CaCO_3 = CaO + CO_2 \tag{1}$$

The calcium oxide then interacts with the sulfur to form calcium sulfide in this manner:

$$2CaO + 2S = 2CaS + O_2 \tag{2}$$

The controller determines the amount of sulfur in the coal by multiplying the mass flow rate by the percentage of sulfur present in the coal. The molar flow rate of sulfur is then determined. Since the molar ratio of calcium oxide to sulfur in equation two is one, the molar flow rate of the sulfur is multiplied directly by the calcium/sulfur ratio setpoint.

The setpoint for the calcium/sulfur ratio can be placed on manual, however, the best method is to have it automatically changed by the PLC based on the SO_2 levels of the stack. In this mode the PLC automatically compensates for the fluctuation of the sulfur in the coal and the sulfur capture efficiency of the limestone. With a particular safe SO_2 setpoint, the PLC will help lower the use of excess limestone.

Coal Flow Rate. The coal flow rate is either determined manually or is set to keep the bed temperature at a certain level. As the swings in the temperature go up and down, the coal rate increases for a decrease in temperature and increases for decrease in temperature. However, if the steam demand is higher than the output, the coal rate will be indirectly increased. This indirect increase is caused by the fact that when the boiler pressure drops, the fluidization velocity increases. Because more air is introduced into the bed, the bed cools down, creating the need for a higher coal input.

Forced Draft/Induced Draft Control

Fluidization Velocity. The flow rate for the forced draft fan is determined by the fluidization setpoint. The fluidization velocity is determined by taking the air flow rate from the inlet duct as well as the air flow rate into the pulse combustor, scaling the volume change based on the temperature of the bed, then dividing by the hot volume by 121 ft^2. This value is the average fluidization of the bed. When the boiler is being used in a minimum swing capacity, the fluidization velocity setpoint changes based on the demand on the boiler. When the boiler is set to produce a fixed amount of steam, the fluidization velocity can be set at the required value.

Induced Draft. The PAFBC unit is designed to operate at atmospheric pressures. In order to achieve this, two pressure sensors have been placed in the freeboard of the bubbling fluid-bed chamber. These pressure sensors send a signal back to the PLC, allowing it to maintain a slightly negative freeboard. The freeboard being slightly negative ensures that no hot coal or ash will escape causing both a pollution as well as a safety hazard. As a precaution for the ID fan motor, the current on one of the phases of the motor is being constantly monitored. The PLC will not allow this motor to open its damper far enough to pull more than 300 amps.

Pulse Combustor Controls

Fine Coal Feeding. The fine coal metering feeder is set to deliver the amount of fine coal specified from the fine coal firing rate set point on the pulse combustor. The pulse combustor has a maximum firing rate of 18 MMBtu/hr. For the most part, the pulse combustor is designed to run at full capacity until it has depleted the fine coal in the fine coal silo. The only thing that will affect the fine coal flow rate in this case is the use of any support natural gas which will cause the fine coal to be fired at a slightly lower rate.

Natural Gas Control. The natural gas is designed to be both a start-up fuel as well as a possible support fuel for the pulse combustor. If the combustion chamber starts to cool down due to a decrease in the combustion of the fine coal in the combustion chamber, then natural gas can be set to automatically keep the combustion chamber at a set point. The pulse combustor will then cut back on the amount of fine coal being supplied to the pulse combustor and add natural gas to bring the chamber temperature up to the required level.

Pulse Combustor Main Air Control. The pulse combustor main air is set to maintain the minimum air-to-fuel ratio. As the amount of natural gas and/or fine coal that is supplied to the combustion chamber changes, the required stoichiometric air rate is recalculated and the main air diverter valve is changed to supply more or less air to the combustion chamber as needed. The main air to the pulse combustor is supplied by a roots blower, therefore, the amount of air delivered to the combustion chamber is controlled by a valve controlling a waste gate diverter for the air supply.

Ash System

The ash-removal timing is located in the ladder logic of the PLC. The manual controls for the ash-handling system are on a screen on the main control computer.

Baghouse

The baghouse timing and ladder logic are kept in a subroutine of the PLC. The control of the baghouse is straightforward. There are 36 solenoid valves on the puffing air of which two fire at the same time. There are four sensors for the baghouse. These sensors are used to determine:

- Inlet duct temperature
- Outlet duct temperature
- Differential pressure across the bags
- Pressure of compressed air header

The bypass valve on the baghouse opens under these emergency conditions:

- The inlet temperature is too hot or cold
- The outlet temperature is too hot or cold
- The differential pressure across the bags is too high

A picture of the baghouse structure is provided in *Figure 2-69*.

Pilot Burners

The controls for the pilots are manual. The flame safety systems are Eclipse Fireye Panels with Interlock catchers to save the first trip. The controls for the gas to the pilots are rotameters. There are manual block valves, as well as automatic block valves in the gas line. The pilot for the pulse combustor is controlled by a gate valve. This valve operates as a waste gate in the same manner that the pulse combustor main air control valve works. The blower for the pulse combustor is also a roots blower and therefore there is not a procedure to throttle the blower. The pilot for the bed uses a standard rotameter to control the air rate. In both cases, there are certain ratios and setpoints to follow. Both pilots are strong enough that if the air-to-fuel ratio is off or if the pilot is being fired too hard, the possibility for flame impingement arises.

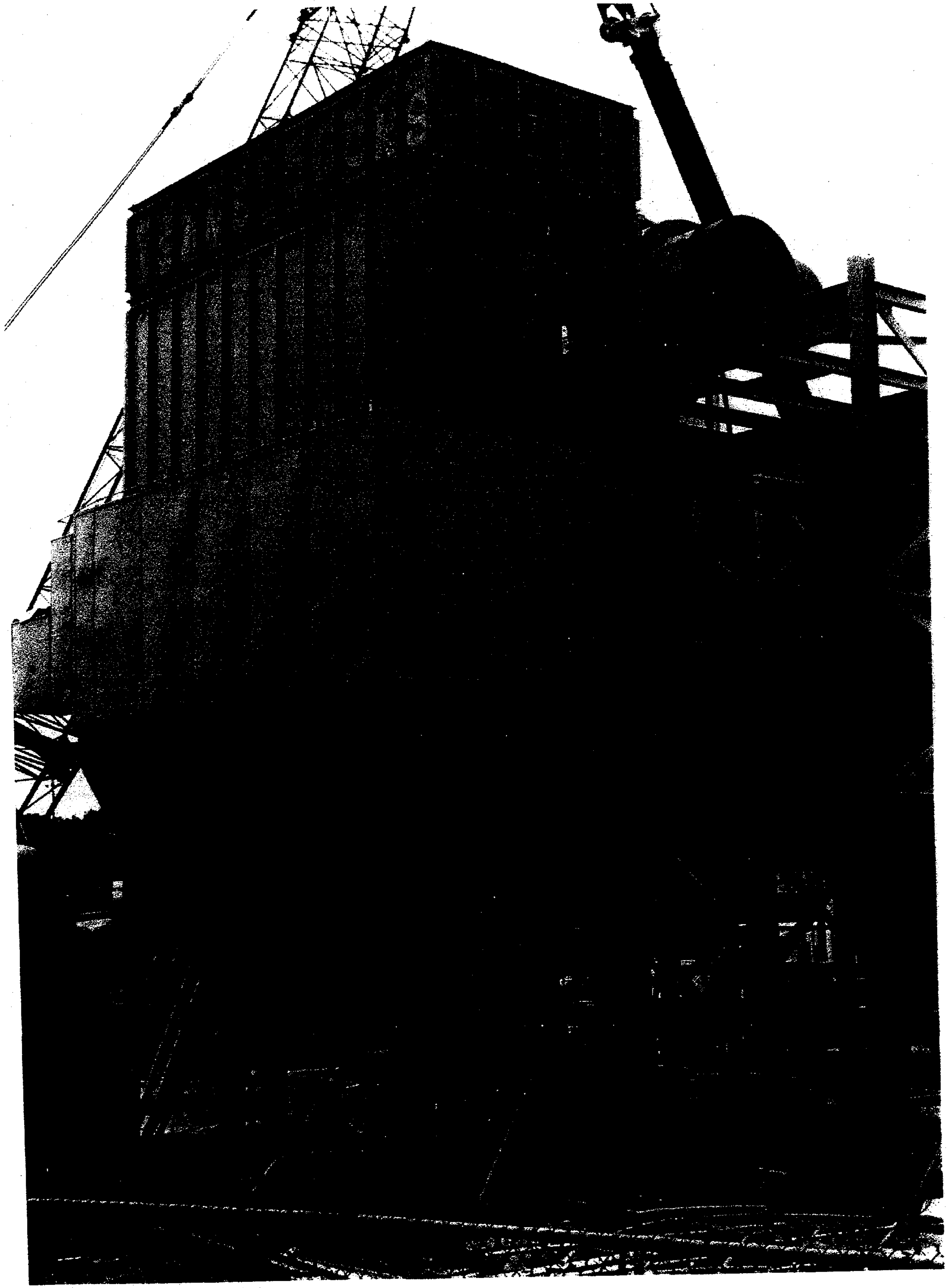


FIGURE 2-69: CLEMSON PAFBC UNIT - BAGHOUSE EXTERNAL VIEW

2.3 SITE DEMONSTRATION TEST PROGRAM

Following the PAFBC unit shakedown at Clemson, South Carolina, ThermoChem began operational testing in March 1997. During this demonstration period, the unit supplied steam for district heating at Clemson University. Although the nominal design firing rate is 72 MMBtu/hr and the nominal design steam generation rate is 60,000 lb/hr at 120 psig, the unit supplied steam with flows ranging from 20,000 to 50,000 lb/hr depending on demand by the University.

It was impossible during the short demonstration period to procure coal within project specifications to achieve design performance. Instead, a low sulfur was used which was readily available at Clemson University. Sulfur dioxide emissions with that coal were at undetectable levels. Limestone was used as a bed material and sorbent at a feed rate of about 900 lb/hr to keep the bed height at nominal design level. Because of low sulfur content in the coal, the lime sulfidation rate was low and the bed elutriation rate was high. This caused overloading of the ash-removal system which would not have occurred utilizing high sulfur coal.

Overloading of the ash-removal system was compounded by poor performance of the J-valve at the bottom of the solids recycling cyclone, which is located at the side of the fluidized bed vessel. This design was chosen to reduce total height of the unit and should be replaced by an overhead cyclone design, or ash should be recycled by a screw feeder installed at the bottom of the cyclone.

To avoid the overloading of the ash-handling system, it was decided to use sand as inert bed material during the demonstration test. Utilization of sand successfully reduced the high carryover into the ash-handling system. A mobile, continuous stack gas monitoring system from Horiba was used during the demonstration tests.

The system is based on cross-modulated, non-dispersive infrared analyzers for NO_x, SO_2, CO_2, CO and HC concentrations and magnetopneumatic analyzer for O_2 concentration in the flue gas. Operational reliability has been improved by the incorporation of microprocessor technology.

Measurement results are displayed on an easy-to-read digital display and saved every minute on a computer. The monitor allows simultaneous sampling, reading and printing of the size concentrations. The ranges on the Horiba monitor are as follows:

CO	0 - 500 ppm, 0 - 2500 ppm
CO_2	0 - 5%, 0 - 25%
O_2	0 - 10%, 0 - 25%
SO_2	0 - 1000 ppm, 0 - 2000 ppm
NO_x	0 - 50 ppm, 0 - 500 ppm
HC	0 - 10 ppm, 0 - 30 ppm, 0 - 100 ppm, 0 - 300 ppm, 0 - 1000 ppm, 0 - 3000 ppm, 0 - 10000 ppm, 0 - 30000 ppm

Repeatability of the measurements are normally 0.5% of full scale and 1.0% of full scale for ranges less than 200 ppm. Drift of zero point is normally 1.0% of full scale per week, and 2% of full scale per week for ranges less than 200 ppm. Drift of span is 2% of full scale per week. The Horiba stack gas analysis system is installed in a trailer and can be moved or transported for mobile use.

The drawings of the mobile unit interior configuration and continuous emissions monitoring (CEM) interconnection configuration are shown in *Figures 2-70* and *2-71*.

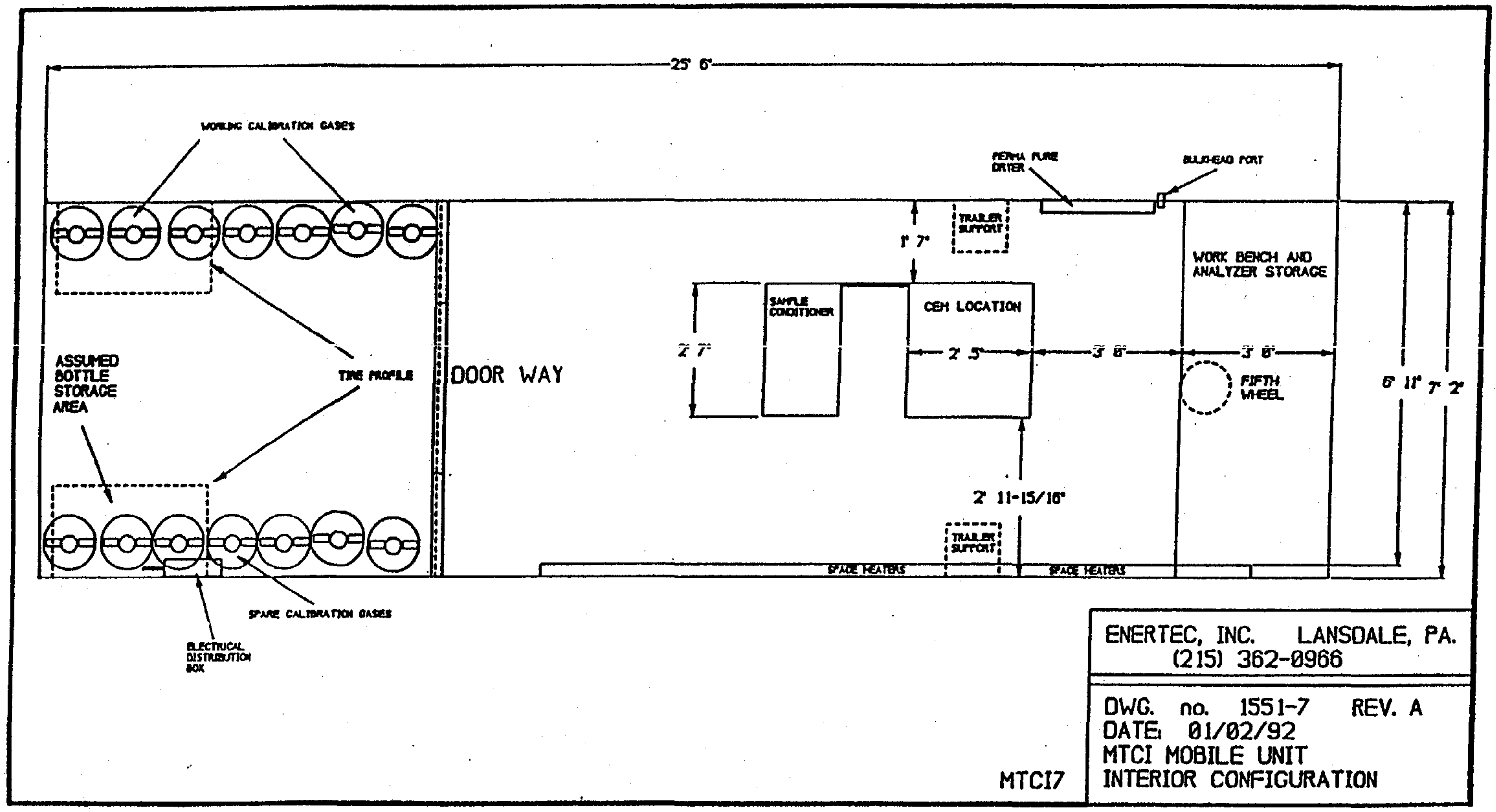


FIGURE 2-70: MTCI MOBILE GAS ANALYSIS UNIT INTERIOR CONFIGURATION

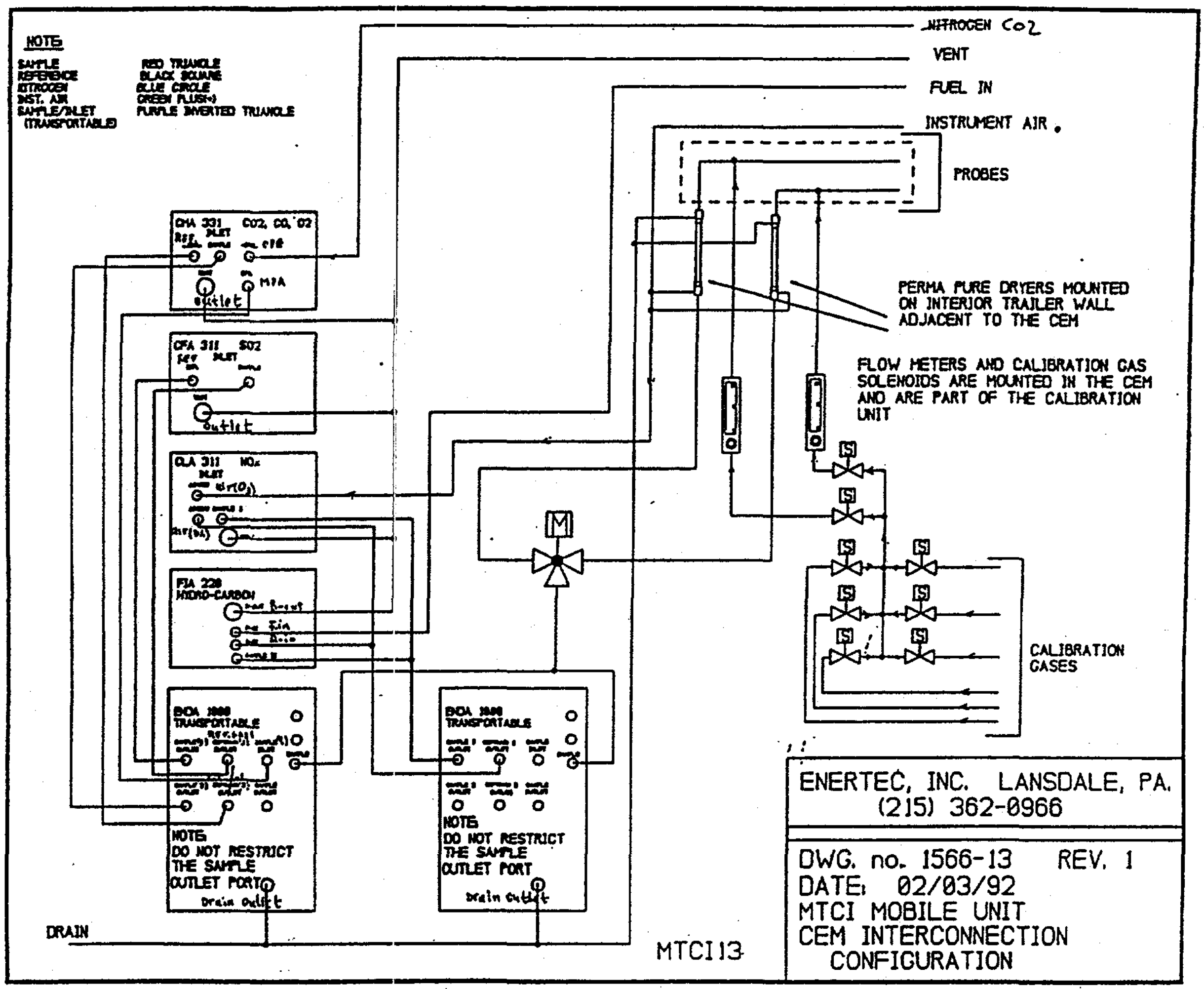


FIGURE 2-71: CEM INTERCONNECTION CONFIGURATION

At about 40,000 lb/hr steam generation rate, the flue gas composition was:

	By Volume:	
Component	Actual	Corrected to 3% O_2
O_2	8.5%	3%
CO	39 ppm	56 ppm
NO_x	52 ppm	75 ppm
SO_2	340 ppm	491 ppm
THC	0 ppm	0 ppm

(Stack opacity was 12%.)

The high level of SO_2 emissions was due to utilizing sand as bed material. Using the sand instead of limestone also caused erosion of the coal/sorbent transportation line.

The PAFBC unit was run with coal feed for 36 hours. The overall performance and gas emissions data for the test duration are presented in the following *30* tables starting on page 105. The data shows good performance with gas at steady-state condition and freeboard temperature at about 1500°F. The data demonstrate that the system would have operated as expected had there not been any coal/ash handling problems.

March 6, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate, MBTU/hr	Coal firing rate, MBTU/hr	Total firing rate, MBTU/hr	Steam flow rate, lbs/hr	Exp.steam flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs xit.temp. F	Stack temp., F
00:13									2627	6097	1.22	424				
01:13									2597	6127	1.22	429				
02:13									2564	6074	1.20	482				
03:13	17053	0	922	17.74	0.00	17.74	14595	14595	2539	6082	1.20	538		281	218	233
04:13	17053	0	922	17.74	0.00	17.74	14438	14399	2539	6012	1.19	559		280	214	229
05:13	17056	0	1097	17.74	0.00	17.74	14618	14439	2575	5775	1.17	555		278	216	229
06:13	17119	0	992	17.80	0.00	17.80	14971	14727	2564	5751	1.16	579		278	214	228
07:47	17144	0	976	17.83	0.00	17.83	15082	15082	2586	5879	1.18	618	746	283	213	227
08:47	17050	0	976	17.73	0.00	17.73	15779	15239	2625	5903	1.19	626	743	282	214	228
09:47	17098	0	975	17.78	0.00	17.78	16046	14903	2627	5989	1.20	632	744	282	214	229
10:47	17130	0	977	17.82	0.00	17.82	10105	10105	2627	5767	1.17	651	871	274	221	233
11:47	17128	0	975	17.81	0.00	17.81	7681	7681	2627	6143	1.23	648	806	275	199	213
12:47	17145	0	976	17.83	0.00	17.83	14840	11382	2589	6285	1.24	619	780	279	217	230
13:47	17046	0	658	17.73	0.00	17.73	14048	12665	2627	7119	1.36	536	749	283	223	236
14:47	16965	0	742	17.64	0.00	17.64	10898	10898	2627	7508	1.41	569	691	290	228	242
15:47	17207	0	976	17.90	0.00	17.90	15102	12162	2627	7408	1.40	594	771	285	228	242
16:47	21008	0	975	21.85	0.00	21.85	15413	15413	2627	6482	1.27	634	853	285	224	238
17:47	17247	0	976	17.94	0.00	17.94	15205	15205	2627	6694	1.30	625	828	285	222	236
18:47	11254	0	975	11.70	0.00	11.70	9612	9612	1522	6203	1.08	329	416	291	217	231
19:47	8160	0	976	8.49	0.00	8.49	3704	3381	2433	6097	1.19	383	748	281	230	244
20:47	14608	0	760	15.19	0.00	15.19	11658	5745	2399	8173	1.47	514	607	288	225	239
21:47	14498	0	650	15.08	0.00	15.08	12781	11791	2376	8150	1.47	532	618	289	224	239
22:47	14241	0	651	14.81	0.00	14.81	4820	4820	2338	8184	1.47	525	608	289	223	237
23:47	16319	0	335	16.97	0.00	16.97	12601	10423	2375	8156	1.47	541	625	291	224	239

March 6, 1997 EMISSIONS DATA

Time	Stack	Stack emissions					
	opacity,	CO	CO2	O2	SO2	NOx	HC
	%	ppm	%	%	ppm	ppm	ppm
00:13							
01:13							
02:13							
03:13	0	11	2.45	16.36	2	17.8	0
04:13	0	11	2.43	16.45	3	17.1	0
05:13	0	14	2.62	16.16	4	18.2	0
06:13	0	13	2.63	16.25	4	18.2	0
07:47	0	19	2.68	16.21	5	18.7	0
08:47	0	18	2.71	16.21	5	17.8	0
09:47	0	12	2.69	16.3	5	18.1	0
10:47	0	19	2.81	16.01	4	16.8	0
11:47	0	16	2.78	16.01	4	16.5	0
12:47	0	11	2.77	16.02	4	17.8	0
13:47	0	8	2.44	16.59	3	15.3	0
14:47	0	9	2.29	16.79	3	14	0
15:47	0	14	2.43	16.48	3	14.3	0
16:47	0	18	2.65	16.07	6	17	32
17:47	0	31	2.64	16.09	3	15.9	1
18:47	0	2	0.09	20.95	3	0.1	1
19:47	0	341	2.62	16.1	3	12.8	0
20:47	0	4	1.93	17.6	4	12.6	0
21:47	0	10	2.02	17.48	4	12.9	0
22:47	0	5	1.97	17.66	5	12.9	0
23:47	0	14	1.9	17.85	5	12.6	0

March 7, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate MBTU/hr	Coal firing rate MBTU/hr	Total firing rate MBTU/hr	Steam flow rate, lbs/hr	Exp.stea flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs xit.temp. F	Stack temp., F
00:47	14428	0	608	15.01	0	15.01	5077	5077	2357	8139	1.46	507	593	291	227	243
01:47	14274	0	399	14.84	0	14.84	11432	3812	2367	8218	1.47	507	593	293	228	244
02:47	14332	0	0	14.91	0	14.91	11328	11328	2448	8241	1.49	503	601	290	229	245
03:47	14583	0	0	15.17	0	15.17	3895	3895	2456	8359	1.50	498	608	291	228	244
04:47	14473	0	0	15.05	0	15.05	7	7	2448	8387	1.51	487	607	290	226	242
05:47	14630	0	0	15.22	0	15.22	30	30	2474	8464	1.52	480	611	293	227	243
06:47	14857	0	0	15.45	0	15.45	70	70	2465	8448	1.52	479	617	294	225	241
07:47	14841	0	0	15.43	0	15.43	28	28	2478	8503	1.53	486	628	293	225	240
08:47	14688	0	0	15.28	0	15.28	4850	2495	2627	8606	1.56	484	636	288	228	244
09:47	14517	0	0	15.10	0	15.10	3131	2861	2627	8519	1.55	474	636	296	233	248
10:47	14154	0	0	14.72	0	14.72	4714	1987	2627	8420	1.54	461	636	298	235	250
11:47	14154	0	0	14.72	0	14.72	1533	1533	2627	8622	1.56	454	639	292	234	249
12:47	14131	0	0	14.70	0	14.70	1906	1906	2627	8692	1.57	440	640	293	236	251
13:47	13959	0	0	14.52	0	14.52	3025	1331	2627	8364	1.53	424	639	290	238	254
14:47	13916	0	0	14.47	0	14.47	11375	5005	2627	8348	1.53	406	638	293	238	253
15:47	13929	0	0	14.49	0	14.49	8217	8217	2627	8359	1.53	393	643	307	247	260
16:47	5318	0	0	5.53	0	5.53	0	0	142	0	0.03	233	391	263	219	217
17:47	0	0	0	0.00	0	0.00	0	0	130	0	0.03	230	422	225	203	179
18:47	0	0	0	0.00	0	0.00	0	0	188	0	0.04	214	380	195	187	163
19:47	0	0	0	0.00	0	0.00	0	0	197	0	0.04	206	361	184	164	145
20:47	0	0	0	0.00	0	0.00	0	0	193	0	0.04	199	337	166	144	130
21:47	0	0	0	0.00	0	0.00	0	0	209	0	0.04	195	319	152	128	116
22:47	0	0	0	0.00	0	0.00	0	0	205	0	0.04	193	310	138	110	103
23:47	0	0	0	0.00	0	0.00	0	0	205	0	0.04	191	299	127	98	92

March 7, 1997 EMISSIONS DATA

Time	Stack opacity, %	Stack emissions					
		CO ppm	CO2 %	O2 %	SO2 ppm	NOx ppm	HC ppm
00:47	0	7	1.87	17.93	5	11.5	0
01:47	100	5	1.84	18.06	5	11.7	0
02:47	57	4	1.83	18.08	5	11.4	0
03:47	54						
04:47	100						
05:47	100						
06:47	100						
07:47	0						
08:47	0						
09:47	0						
10:47	0						
11:47	0						
12:47	0						
13:47	0						
14:47	0						
15:47	0						
16:47	0						
17:47	0						
18:47	0						
19:47	0						
20:47	0						
21:47	0						
22:47	0						
23:47	0						

March 8, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate MBTU/hr	Coal firing rate MBTU/hr	Total firing rate MBTU/hr	Steam flow rate, lbs/hr	Exp.steam flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs exit.temp. F	Stack temp., F
00:47							0		201	0	0.04	189	287	118	91	84
01:47	0	0	0	0.00	0.00	0.00	0	0	221	0	0.04	186	277	112	87	78
02:47	0	0	0	0.00	0.00	0.00	0	0	205	0	0.04	185	267	107	81	73
03:47	0	0	0	0.00	0.00	0.00	0	0	217	0	0.04	183	258	103	78	69
04:47	0	0	0	0.00	0.00	0.00	0	0	205	0	0.04	181	249	100	76	66
05:47	0	0	0	0.00	0.00	0.00	0	0	201	0	0.04	179	242	98	75	65
06:47	0	0	0	0.00	0.00	0.00	0	0	213	0	0.04	176	232	102	79	66
07:47	0	0	0	0.00	0.00	0.00	0	0	205	0	0.04	174	225	103	80	67
08:47	0	0	0	0.00	0.00	0.00	0	0	197	0	0.04	173	219	103	80	67
09:47	0	0	0	0.00	0.00	0.00	0	0	159	0	0.03	171	213	104	82	68
10:47	0	0	0	0.00	0.00	0.00	0	0	154	0	0.03	169	208	104	83	71
11:47	0	0	0	0.00	0.00	0.00	0	0	136	0	0.03	171	205	99	81	72
12:47	0	0	0	0.00	0.00	0.00	0	0	154	1806	0.28	170	200	97	83	75
13:47	0	0	0	0.00	0.00	0.00	0	0	159	0	0.03	169	198	98	86	78
14:47	0	0	0	0.00	0.00	0.00	0	0	136	0	0.03	167	194	98	87	82
15:47	0	0	0	0.00	0.00	0.00	0	0	154	0	0.03	166	190	98	87	82
16:47	0	0	0	0.00	0.00	0.00	0	0	169	0	0.03	164	187	98	88	85
17:47	0	0	0	0.00	0.00	0.00	0	0	164	0	0.03	162	185	97	87	83
18:47	605	0	0	0.63	0.00	0.63	0	0	174	0	0.03	161	182	97	85	81
19:47	108	0	0	0.11	0.00	0.11	0	0	193	0	0.04	160	180	95	84	78
20:47	93	0	0	0.10	0.00	0.10	0	0	197	0	0.04	158	178	93	81	75
21:47	11	0	0	0.01	0.00	0.01	0	0	209	0	0.04	157	176	91	78	72
22:47	0	0	0	0.00	0.00	0.00	0	0	201	0	0.04	155	174	89	76	70
23:47	0	0	0	0.00	0.00	0.00	0	0	217	0	0.04	154	172	87	74	68

March 8, 1997 EMISSIONS DATA

Time	Stack opacity, %	Stack emissions					
		CO ppm	CO2 %	O2 %	SO2 ppm	NOx ppm	HC ppm
00:47	0						
01:47	0						
02:47	0						
03:47	0						
04:47	0						
05:47	0						
06:47	0						
07:47	0						
08:47	0						
09:47	0						
10:47	0						
11:47	0						
12:47	0						
13:47	0						
14:47	0						
15:47	0						
16:47	0						
17:47	0						
18:47	0						
19:47	0						
20:47	0						
21:47	0						
22:47	0						
23:47	0						

March 9, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate MBTU/hr	Coal firing rate MBTU/hr	Total firing rate MBTU/hr	Steam flow rate, lbs/hr	Exp.steam flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs exit.temp. F	Stack temp., F
00:47							0		209	0	0.04	152	170	85	72	66
01:47	0	0	0	0.00	0.00	0.00	0	0	225	0	0.04	151	169	83	70	64
02:47	0	0	0	0.00	0.00	0.00	0	0	232	0	0.04	150	168	80	68	62
03:47	0	0	0	0.00	0.00	0.00	0	0	229	0	0.04	149	166	78	66	60
04:47	0	0	0	0.00	0.00	0.00	0	0	236	0	0.04	147	164	77	65	58
05:47	0	0	0	0.00	0.00	0.00	0	0	232	0	0.04	144	163	75	64	57
06:47	0	0	0	0.00	0.00	0.00	0	0	236	0	0.04	143	162	74	63	56
07:47	0	0	0	0.00	0.00	0.00	0	0	209	0	0.04	142	160	75	64	56
08:47	0	0	0	0.00	0.00	0.00	0	0	209	0	0.04	141	159	76	66	57
09:47	0	0	0	0.00	0.00	0.00	0	0	307	0	0.05	124	157	76	69	58
10:47	0	0	0	0.00	0.00	0.00	0	0	201	1330	0.22	121	147	88	69	108
11:47	3	0	0	0.00	0.00	0.00	0	0	169	1753	0.27	122	152	95	88	95
12:47	52	0	0	0.05	0.00	0.05	0	0	159	0	0.03	122	152	93	88	91
13:47	167	0	0	0.17	0.00	0.17	0	0	169	0	0.03	122	152	94	90	91
14:47	312	0	0	0.32	0.00	0.32	0	0	164	0	0.03	122	151	93	88	91
15:47	669	0	0	0.70	0.00	0.70	0	0	169	0	0.03	122	150	92	87	90
16:47	928	0	0	0.96	0.00	0.96	0	0	174	0	0.03	123	149	92	87	90
17:47	689	0	0	0.72	0.00	0.72	0	0	188	0	0.04	123	149	93	88	85
18:47	545	0	0	0.57	0.00	0.57	0	0	193	0	0.04	123	148	92	87	81
19:47	348	0	0	0.36	0.00	0.36	0	0	201	0	0.04	123	148	90	85	76
20:47	145	0	0	0.15	0.00	0.15	0	0	205	0	0.04	123	147	88	83	73
21:47	81	0	0	0.08	0.00	0.08	0	0	197	0	0.04	123	146	86	80	69
22:47	43	0	0	0.05	0.00	0.05	0	0	213	0	0.04	123	146	84	79	68
23:47	46	0	0	0.05	0.00	0.05	0	0	210	0	0.04	123	145	83	77	68

March 09, 1997 EMISSIONS DATA

Time	Stack opacity,	Stack emissions					
		CO	CO2	O2	SO2	NOx	HC
	%	ppm	%	%	ppm	ppm	ppm
00:47	0						
01:47	0						
02:47	0						
03:47	0						
04:47	0						
05:47	0						
06:47	57						
07:47	0						
08:47	0						
09:47	0						
10:47	7						
11:47	9						
12:47	9						
13:47	12						
14:47	9						
15:47	9						
16:47	9						
17:47	9						
18:47	9	0	0.07	20.46	0	0	0
19:47	9	0	0.07	20.43	1	0	0
20:47	9	0	0.07	20.48	1	0	0
21:47	9	0	0.08	20.5	1	0	0
22:47	9	0	0.08	20.51	2	0	0
23:47	8	1	0.08	20.5	2	0	0

March 10, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate MBTU/hr	Coal firing rate MBTU/hr	Total firing rate MBTU/hr	Steam flow rate, lbs/hr	Exp.steam flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs exit.temp. F	Stack temp., F
00:47							0		205	0	0.04	122	144	82	77	68
01:47	93	0	0	0.10	0.00	0.10	0	0	205	0	0.04	123	144	80	75	67
02:47	69	0	0	0.07	0.00	0.07	0	0	201	0	0.04	123	143	78	73	66
03:47	41	0	0	0.04	0.00	0.04	0	0	205	0	0.04	122	143	77	72	65
04:47	25	0	0	0.03	0.00	0.03	0	0	201	0	0.04	122	143	76	72	65
05:47	15	0	0	0.02	0.00	0.02	0	0	217	0	0.04	122	142	75	71	65
06:47	14	0	0	0.01	0.00	0.01	0	0	205	0	0.04	122	141	74	70	64
07:47	24	0	0	0.02	0.00	0.02	0	0	221	0	0.04	122	140	73	69	64
08:47	26	0	0	0.03	0.00	0.03	0	0	205	0	0.04	121	139	73	68	68
09:47	62	0	0	0.06	0.00	0.06	0	0	201	0	0.04	121	139	74	69	73
10:47	165	0	0	0.17	0.00	0.17	0	0	82	0	0.02	121	138	76	72	76
11:47	319	0	0	0.33	0.00	0.33	0	0	123	0	0.03	121	138	78	76	77
12:47	469	0	0	0.49	0.00	0.49	0	0	179	0	0.03	120	137	79	79	79
13:47	528	0	0	0.55	0.00	0.55	0	0	174	0	0.03	120	137	81	81	82
14:47	3998	0	0	4.16	0.00	4.16	0	0	184	0	0.03	120	390	169	148	161
15:47	8558	0	0	8.90	0.00	8.90	0	0	1329	8064	1.30	117	423	237	203	210
16:47	3704	0	0	3.85	0.00	3.85	0	0	1318	8133	1.31	134	274	196	157	146
17:18	2746	0	0	2.86	0.00	2.86	0	0	188	0	0.04	171	419	247	219	223
18:18	9725	0	0	10.11	0.00	10.11	0	0	1583	8820	1.44	140	473	283	222	237
19:18	13752	0	0	14.30	0.00	14.30	5048	390	1590	8778	1.44	163	681	284	234	250
20:18	17173	0	13	17.86	0.00	17.86	0	2389	2576	8196	1.50	288	693	287	231	247
21:18	16904	0	304	17.58	0.00	17.58	4122	2158	2620	8342	1.52	375	683	288	232	248
22:18	16832	0	581	17.51	0.00	17.51	0	2921	2610	8241	1.51	435	673	288	232	247
23:18	16696	0	581	17.36	0.00	17.36	4428	3079	2627	8190	1.51	496	667	286	230	246

March 10, 1997 EMISSIONS DATA

Time	Stack	Stack emissions					
	opacity,	CO	CO2	O2	SO2	NOx	HC
	%	ppm	%	%	ppm	ppm	ppm
00:47	0	1	0.08	20.48	2	0	0
01:47	8	1	0.08	20.48	2	0	0
02:47	8	1	0.08	20.48	2	0	0
03:47	8	1	0.08	20.45	2	0	0
04:47	8	1	0.08	20.45	2	0	0
05:47	8	1	0.09	20.47	2	0	0
06:47	8	1	0.08	21	6	0	0
07:47	8	1	0.08	21.03	3	0	0
08:47	8	1	0.09	21.01	2	0	0
09:47	9	1	0.08	20.96	2	0	0
10:47	8	1	0.08	20.92	1	0	0
11:47	8	1	0.08	20.85	1	0	0
12:47	8	0	0.07	20.82	0	0	0
13:47	8	0	0.07	20.81	0	0	0
14:47	32	3	1.35	18.44	0	4	0
15:47	25	2	1.3	18.51	0	3.5	0
16:47	16	1	0.09	20.76	0	0	0
17:18	32	9	1.38	18.36	0	1.7	0
18:18	16	2	1.39	18.34	0	3.4	0
19:18	16	9	2.2	16.81	0	5.4	0
20:18	16	13	2.21	16.81	1	5.4	0
21:18	16	22	2.14	16.98	1	4.8	0
22:18	16	37	2.17	16.96	1	4.8	0
23:18	16	62	2.2	16.92	1	4.6	0

March 11, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate MBTU/hr	Coal firing rate MBTU/hr	Total firing rate MBTU/hr	Steam flow rate, lbs/hr	Exp.steam flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs exit.temp. F	Stack temp., F
00:18							1617		2562	8081	1.48	559	661	286	229	245
01:18	16867	0	329	17.54	0.00	17.54	4353	3705	2585	8139	1.49	593	693	280	225	240
02:18	16821	0	155	17.49	0.00	17.49	4428	3567	2573	8167	1.50	565	665	288	228	243
03:18	16825	0	581	17.50	0.00	17.50	0	3875	2537	8081	1.48	569	659	286	228	244
04:18	17036	0	475	17.72	0.00	17.72	5113	4630	2522	7912	1.45	575	657	289	228	244
05:18	17089	0	581	17.77	0.00	17.77	3333	4266	2506	7912	1.45	566	640	282	227	243
06:18	17083	0	581	17.77	0.00	17.77	4501	4301	2502	7983	1.46	561	630	294	229	245
07:18	16771	0	581	17.44	0.00	17.44	3430	3610	2510	7889	1.45	562	625	284	228	244
08:18	7043	0	248	7.32	0.00	7.32	6517	3673	2498	7818	1.44	564	624	284	230	246
09:18	0	0	0	0.00	0.00	0.00	13234	8946	2501	7818	1.44	560	627	283	236	252
10:18	0	0	0	0.00	0.00	0.00	13381	9040	2514	7835	1.44	555	625	289	238	253
11:18	0	0	0	0.00	0.00	0.00	14640	10345	2508	6879	1.31	563	722	301	240	254
12:19	0	0	209	0.00	0.00	0.00	16005	14468	2565	6561	1.27	622	751	279	231	247
13:18	5838	0	0	6.07	0.00	6.07	5716	11187	2578	6575	1.28	638	769	279	235	250
14:18	17287	0	81	17.98	0.00	17.98	10096	7577	2525	6694	1.29	635	791	291	237	252
15:19	17731	0	534	18.44	0.00	18.44	8781	9826	2575	6756	1.30	605	784	279	234	249
16:19	17479	0	526	18.18	0.00	18.18	8043	10267	2581	7948	1.47	611	830	284	231	246
17:19	17539	0	396	18.24	0.00	18.24	7754	5464	2579	6035	1.20	635	936	288	232	247
18:19	17509	0	17	18.21	0.00	18.21	7185	6852	2516	5911	1.18	550	855	289	235	249
19:19	17546	0	546	18.25	0.00	18.25	6467	6613	2565	6180	1.22	531	851	290	234	249
20:19	11311	0	394	11.76	0.00	11.76	0	4235	221	0	0.04	423	593	258	210	205
21:19	824	0	0	0.86	0.00	0.86	0	0	1413	6066	1.04	329	545	291	203	218
22:19	14505	0	0	15.09	0.00	15.09	5995	385	2565	6805	1.31	487	711	286	228	243
23:19	17394	0	428	18.09	0.00	18.09	4644	3945	2601	6792	1.31	505	681	292	233	249

March 11, 1997 EMISSIONS DATA

Time	Stack	Stack emissions					
	opacity,	CO	CO2	O2	SO2	NOx	HC
	%	ppm	%	%	ppm	ppm	ppm
00:18	16	89	2.21	16.89	2	4.8	0
01:18	16	79	2.46	16.45	2	5.4	0
02:18	16	68	2.25	16.88	3	5	0
03:18	16	135	2.26	16.85	3	5	4
04:18	16	194	2.29	16.81	3	5	5
05:18	16	257	2.24	16.95	3	4.6	1
06:18	58						
07:18	12						
08:18	16	141	2.23	17.16	0	8.7	0
09:18	19	67	2.21	17.1	0	9	0
10:18	12	31	2.15	17.14	0	9.5	0
11:18	12	61	2.28	16.83	0	9.8	0
12:19	13	51	2.71	15.96	0	12	6
13:18	55	361	2.76	15.75	0	12.8	0
14:18	12	198	2.71	15.82	0	13.4	0
15:19	12	201	2.45	16.28	0	12.8	0
16:19	13	113	2.67	15.84	0	14.6	0
17:19	12	95	2.82	15.53	0	15.9	0
18:19	12	59	2.54	16.1	0	14.6	0
19:19	12	92	2.54	16.12	0	14.2	0
20:19	12	4	0.13	20.79	0	1	2
21:19	16	79	1.75	17.78	0	5.7	0
22:19	16	36	2.44	16.5	0	14.2	0
23:19	16	90	2.25	16.93	0	12.5	0

March 12, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate MBTU/hr	Coal firing rate MBTU/hr	Total firing rate MBTU/hr	Steam flow rate, lbs/hr	Exp.steam flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs exit.temp. F	Stack temp., F
00:19							2286		2627	6743	1.31	517	658	288	233	249
01:19	15994	0	526	16.63	0.00	16.63	6517	3912	2627	6913	1.33	536	643	291	232	248
02:19	16114	0	526	16.76	0.00	16.76	8965	2435	2627	6729	1.30	556	644	292	230	247
03:19	16466	0	526	17.12	0.00	17.12	0	4598	2566	6680	1.29	565	637	284	228	245
04:19	16560	0	526	17.22	0.00	17.22	2556	3495	2590	6763	1.30	575	632	288	228	245
05:19	17232	0	526	17.92	0.00	17.92	3430	5374	2542	6743	1.30	586	746	285	231	246
06:19	17431	0	526	18.13	0.00	18.13	6467	10242	2550	6825	1.31	587	637	291	232	246
07:19	17468	0	526	18.17	0.00	18.17	4714	8394	2528	6756	1.30	584	641	286	227	243
08:19	17303	0	526	17.99	0.00	17.99	4042	5358	2516	6582	1.27	586	628	287	228	243
09:19	17115	0	526	17.80	0.00	17.80	5995	11450	2505	6652	1.28	589	623	279	231	247
10:19	17227	0	520	17.92	0.00	17.92	14993	9115	2519	6417	1.25	592	629	284	235	250
11:19	17229	0	145	17.92	0.00	17.92	5773	7977	2473	6666	1.28	588	622	285	235	250
12:19	17018	0	66	17.70	0.00	17.70	13718	11573	2502	6933	1.32	586	630	283	237	252
13:19	16849	0	427	17.52	0.00	17.52	14094	14706	2503	6388	1.24	587	691	279	239	254
14:19	16732	0	455	17.40	0.00	17.40	5601	8211	2503	5365	1.10	624	767	279	234	250
15:19	16864	0	263	17.54	0.00	17.54	6049	5891	2509	5304	1.09	638	906	279	229	244
16:19	22735	0	263	23.64	0.00	23.64	9038	8315	2512	5400	1.11	702	1036	285	233	247
17:19	22238	475	300	23.13	6.23	29.36	7711	8840	2521	5313	1.10	683	849	278	231	247
18:19	17813	0	318	18.53	0.00	18.53	3960	6676	2560	4003	0.92	667	841	276	225	242
19:19	18024	0	494	18.74	0.00	18.74	6811	6963	2627	5287	1.11	622	881	282	224	238
20:19	20933	0	526	21.77	0.00	21.77	9966	7323	2627	4284	0.97	673	953	284	220	237
21:19	20831	715	526	21.66	9.36	31.03	33261	13732	2530	6082	1.21	998	1543	310	236	248
22:19	23661	1410	744	24.61	18.47	43.08	33999	31706	2586	5180	1.10	1254	1524	294	239	254
23:19	13315	1555	775	13.85	20.37	34.22	34863	30666	1732	6439	1.15	1158	1794	305	236	249

March 12, 1997 EMISSIONS DATA

Time	Stack opacity,	Stack emissions					
		CO	CO2	O2	SO2	NOx	HC
	%	ppm	%	%	ppm	ppm	ppm
00:19	16	13	2.12	17.2	0	12.3	0
01:19	16	11	2.1	17.21	0	11.8	0
02:19	16	35	2.17	17.1	0	11.7	0
03:19	16	39	2.21	17.04	0	11.8	0
04:19	16	387	2.28	16.9	0	10	0
05:19	16	268	2.4	16.77	0	10.6	2
06:19	55						
07:19	16	459	2.44	16.66	0	10.4	0
08:19	16	254	2.5	16.6	0	10.7	0
09:19	16	297	2.54	16.51	0	11.2	0
10:19	16	321	2.54	16.46	0	11	0
11:19	16	245	2.45	16.58	0	10.9	0
12:19	16	173	2.45	16.51	24	10.9	767
13:19	16	126	2.6	16.18	0	12.3	169
14:19	16	90	2.82	15.7	0	13.2	0
15:19	16	119	2.83	15.64	2	13.7	59
16:19	16	514	3.21	14.63	519	12.6	1029
17:19	0	496	2.83	15.5	318	13.1	1029
18:19	16	233	2.93	15.35	0	14.2	13
19:19	16	481	2.78	15.48	609	10.6	1029
20:19	16	514	3.23	14.68	488	10.3	1029
21:19	17	177	8.33	8.76	0	38.5	75
22:19	16	514	9.96	6.16	163	33.2	1029
23:19	16	12	9.98	7.37	11	76.8	0

March 13, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate MBTU/hr	Coal firing rate MBTU/hr	Total firing rate MBTU/hr	Steam flow rate, lbs/hr	Exp.steam flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs exit.temp. F	Stack temp., F
00:19							0		1287	8297	1.34	654	667	258	206	224
01:19	12816	0	118	13.33	0.00	13.33	11798	11005	2596	5628	1.15	573	793	263	214	230
02:19	17564	0	526	18.27	0.00	18.27	14415	15630	2627	5611	1.15	604	835	285	223	237
03:19	17714	0	570	18.42	0.00	18.42	7796	7931	2627	5519	1.14	580	827	285	230	245
04:19	17659	0	176	18.37	0.00	18.37	8593	7326	2627	5686	1.16	564	818	287	228	243
05:19	17609	0	0	18.31	0.00	18.31	7583	7163	2627	5735	1.17	552	810	288	229	244
06:19	17492	0	0	18.19	0.00	18.19	6617	7039	2627	5628	1.15	536	807	285	229	244
07:19	17331	0	0	18.02	0.00	18.02	6262	6610	2627	5743	1.17	517	800	288	229	244
08:19	17403	0	0	18.10	0.00	18.10	7047	6770	2627	5611	1.15	515	805	287	228	244
09:19	17441	0	0	18.14	0.00	18.14	1617	6409	2627	4240	0.96	488	936	283	226	242
10:19	17563	0	0	18.27	0.00	18.27	15861	11161	2627	3921	0.91	356	928	278	222	237
11:19	17715	0	125	18.42	0.00	18.42	16107	10674	2627	3944	0.92	352	811	274	216	229
12:19	17715	0	486	18.42	0.00	18.42	7230	8280	2627	4465	0.99	504	826	281	224	238
13:19	17715	0	559	18.42	0.00	18.42	6209	6247	2627	4517	1.00	496	825	279	221	236
14:19	17719	0	925	18.43	0.00	18.43	7230	6117	2608	3991	0.92	530	943	279	217	232
15:19	17715	0	1163	18.42	0.00	18.42	3333	6293	2584	4084	0.94	587	889	282	221	236
16:19	17713	0	1040	18.42	0.00	18.42	7453	6777	2606	4185	0.95	596	883	280	222	236
17:19	17611	0	969	18.32	0.00	18.32	9217	10044	2591	4026	0.93	622	908	289	219	235
18:19	17556	0	839	18.26	0.00	18.26	4501	5788	2610	6240	1.24	594	783	278	227	242
19:19	17628	0	443	18.33	0.00	18.33	6262	6302	2586	4174	0.95	672	884	281	220	237
20:19	17598	0	650	18.30	0.00	18.30	1400	6708	2556	6097	1.21	551	785	277	225	239
21:19	17401	0	651	18.10	0.00	18.10	7001	5538	2523	6112	1.21	604	787	291	229	244
22:03	12522	0	470	13.02	0.00	13.02	7230	4038	2520	6059	1.20	606	781	282	227	242
23:03	17300	0	651	17.99	0.00	17.99	4042	5538	2528	6104	1.21	607	769	281	228	243

March 13, 1997 EMISSIONS DATA

Time	Stack	Stack emissions					
	opacity,	CO	CO2	O2	SO2	NOx	HC
	%	ppm	%	%	ppm	ppm	ppm
00:19	17	219	0.26	20.66	0	0.7	54
01:19	16	13	2.49	16.5	0	9.6	0
02:19	15	132	2.77	15.98	0	9.3	0
03:19	15	119	2.75	16.06	0	9.2	0
04:19	15	94	2.75	16.07	0	9.8	0
05:19	15	72	2.72	16.13	0	9.5	0
06:19	15						
07:19	15	35	2.67	16.29	0	10	0
08:19	15	48	2.67	16.33	0	9.5	0
09:19	15	19	3	15.68	0	11.4	0
10:19	0	62	3.12	15.43	0	10.1	1
11:19	0	47	3.07	15.5	0	10	0
12:19	0	55	2.96	15.67	0	9.5	0
13:19	0	51	2.96	15.65	0	9.8	0
14:19	100	55	3.15	15.25	0	10.4	0
15:19	100	153	3.25	15.05	0	9.8	0
16:19	100	177	3.16	15.21	0	9.5	0
17:19	100	134	3.16	15.21	0	9.6	0
18:19	100	325	2.67	16.1	0	8.4	0
19:19	100	253	3.14	15.19	0	8.5	0
20:19	100	297	2.64	16.14	0	8.1	0
21:19	100	300	2.58	16.2	0	7.8	0
22:03	100	380	2.57	16.18	0	7.8	0
23:03	18	326	2.57	16.17	0	7.8	0

March 14, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate MBTU/hr	Coal firing rate MBTU/hr	Total firing rate MBTU/hr	Steam flow rate, lbs/hr	Exp.steam flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs exit.temp. F	Stack temp., F
00:03							5829						767	291	227	242
01:03	0	0	651	0.00	0.00	0.00	8244	5387	2547	6020	1.20	609	764	292	226	242
02:03	0	0	650	0.00	0.00	0.00	7453	5341	2504	6004	1.19	612	760	281	226	242
03:03	0	0	651	0.00	0.00	0.00	15080	5556	2554	6035	1.20	611	762	280	226	242
04:03	0	0	650	0.00	0.00	0.00	6859	5490	2552	5911	1.18	612	757	289	226	242
05:03	0	0	651	0.00	0.00	0.00	5773	5325	2569	5911	1.19	613	761	281	225	242
06:03	0	0	650	0.00	0.00	0.00	10384	5582	2552	5997	1.19	613	758	281	231	246
07:03	0	0	615	0.00	0.00	0.00	8244	5597	2570	6028	1.20	626	770	290	231	246
08:03	0	0	112	0.00	0.00	0.00	3131	5942	2580	3599	0.87	685	906	271	215	233
09:03	0	0	388	0.00	0.00	0.00	0	6112	2563	3663	0.88	653	908	272	217	234
10:03	0	0	388	0.00	0.00	0.00	8084	5248	2513	6089	1.20	602	802	294	233	247
11:03	0	0	388	0.00	0.00	0.00	4122	4848	2546	6743	1.30	590	774	293	235	251
12:03	0	0	407	0.00	0.00	0.00	5176	4957	2567	6832	1.31	596	772	285	233	248
13:03	0	0	463	0.00	0.00	0.00	7837	4919	2554	6120	1.21	618	797	289	229	245
14:03	1681	0	615	1.75	0.00	1.75	4782	0	2538	6028	1.20	617	782	279	233	248
15:13	20151	0	912	20.96	0.00	20.96	5716	0	2552	6043	1.20	618	770	287	232	247
16:13	17427	0	793	18.12	0.00	18.12	8706	0	2618	6143	1.22	614	770	293	231	246
17:13	17737	0	526	18.45	0.00	18.45	6262	0	2526	4229	0.95	668	783	292	230	246
18:13	20840	640	411	21.67	8.38	30.05	27150	121	2620	4959	1.07	1063	1454	301	236	249
19:13	9553	2018	438	9.93	26.43	36.37	38062	33622	1136	8935	1.41	1041	1894	317	253	267
20:13	17107	1841	689	17.79	24.11	41.91	38387	40223	2615	6839	1.33	1356	1508	311	249	265
21:13	17426	1823	884	18.12	23.88	42.00	36178	38308	2603	6933	1.34	1382	1481	303	243	261
22:13	17492	1865	913	18.19	24.43	42.62	38071	39348	2601	7106	1.37	1384	1485	311	249	265
23:13	17704	1866	945	18.41	24.44	42.86	37976	37806	2627	7093	1.37	1366	1500	307	243	259

March 14, 1997 EMISSIONS DATA

Time	Stack	Stack emissions					
	opacity,	CO	CO2	O2	SO2	NOx	HC
	%	ppm	%	%	ppm	ppm	ppm
00:03	0	366	2.57	16.09	0	7.8	0
01:03	18	427	2.58	16.03	0	7.5	0
02:03	28	356	2.55	16.03	0	7.8	0
03:03	23	316	2.53	16.03	0	7.8	0
04:03	20	308	2.53	16	0	7.9	0
05:03	25	286	2.52	16.02	0	7.9	0
06:03	67	302	2.63	15.79	0	8.1	0
07:03	22	353	2.67	15.77	0	9.6	0
08:03	25	190	3.12	14.97	0	10.3	0
09:03	9	119	3.1	15.03	0	10.7	0
10:03	28	165	2.56	16.1	0	9.2	0
11:03	13	240	2.42	16.31	0	8.7	0
12:03	20	332	2.42	16.25	0	8.5	0
13:03	0	321	2.58	15.93	0	9	0
14:03	9	292	2.64	15.82	0	9.3	0
15:13	3	296	2.6	15.82	0	9.5	0
16:13	0	479	2.53	15.92	0	8.9	0
17:13	0	298	3.01	15.01	11	9.2	103
18:13	1	514	9.31	5.95	96	29.3	679
19:13	2	16	9.01	8.82	0	112.1	5
20:13	18	62	9.06	8.26	0	60.4	0
21:13	12	75	8.88	8.72	0	80.9	0
22:13	4	82	8.79	8.87	0	105	0
23:13	2	87	8.21	9.41	0	101.8	0

March 15, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate MBTU/hr	Coal firing rate MBTU/hr	Total firing rate MBTU/hr	Steam flow rate, lbs/hr	Exp.steam flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs exit.temp. F	Stack temp., F
00:13							42775		2627	7106	1.37	1392	1518	312	249	265
01:13	17735	1870	970	18.44	24.50	42.95	43950	42166	2611	7191	1.38	1400	1516	313	249	265
02:13	17710	1868	969	18.42	24.47	42.89	44298	42314	2604	7351	1.40	1463	1527	336	265	279
03:13	19987	437	969	20.79	5.73	26.51	20306	29572	2569	4413	0.98	686	925	309	238	252
04:13	17715	0	969	18.42	0.00	18.42	10446	16885	2564	4455	0.98	670	864	298	232	246
05:13	17705	0	968	18.41	0.00	18.41	9358	8897	2567	4475	0.99	612	907	291	227	243
06:13	17720	0	969	18.43	0.00	18.43	8084	8012	2623	6852	1.32	564	752	298	232	251
07:13	17715	0	969	18.42	0.00	18.42	7047	7511	2627	6839	1.32	576	710	291	235	253
08:13	17684	0	969	18.39	0.00	18.39	8124	6934	2627	6784	1.31	592	712	298	234	252
09:13	17617	0	969	18.32	0.00	18.32	6262	6861	2627	6610	1.29	601	721	288	237	252
10:13	17619	0	969	18.32	0.00	18.32	5940	6771	2627	6337	1.25	600	709	287	236	251
11:13	17617	0	969	18.32	0.00	18.32	4644	6570	2627	6120	1.22	625	731	284	234	250
12:13	25925	0	969	26.96	0.00	26.96	6156	6947	2627	6188	1.23	630	730	294	236	252
13:13	20259	0	969	21.07	0.00	21.07	7275	6911	2627	5919	1.19	635	735	293	243	256
14:13	18712	0	969	19.46	0.00	19.46	4714	6240	2627	6012	1.21	620	718	288	243	256
15:13	17651	0	943	18.36	0.00	18.36	6763	5435	2627	6180	1.23	613	705	288	238	253
16:13	18808	0	775	19.56	0.00	19.56	2425	5535	2627	6059	1.21	618	710	286	236	251
17:13	19781	0	775	20.57	0.00	20.57	7754	6039	2627	5981	1.20	621	712	296	272	267
18:13	18025	0	775	18.75	0.00	18.75	7669	6157	2627	6051	1.21	626	716	294	260	264
19:13	17802	0	775	18.51	0.00	18.51	7409	6491	2602	6104	1.22	630	719	287	246	252
20:13	17715	0	775	18.42	0.00	18.42	11602	7206	2627	6059	1.21	638	727	288	238	245
21:13	17705	0	775	18.41	0.00	18.41	6811	6860	2627	6112	1.22	634	722	286	231	240
22:13	17715	0	775	18.42	0.00	18.42	4277	7118	2627	6097	1.22	633	728	298	211	222
23:13	17705	0	775	18.41	0.00	18.41	1617	6663	2627	6203	1.23	632	732	292	223	238

March 15, 1997 EMISSIONS DATA

Time	Stack	Stack emissions					
	opacity,	CO	CO2	O2	SO2	NOx	HC
	%	ppm	%	%	ppm	ppm	ppm
00:13	10	85	8.7	9.14	0	110.9	0
01:13	9	92	8.64	9.21	0	116.2	0
02:13	62	87	11.83	7.79	0	148.4	0
03:13	0	56	2.78	15.9	0	7.9	0
04:13	7	96	2.85	15.9	0	7.8	0
05:13	9	73	2.95	15.82	0	7.8	0
06:13	12						
07:13	13	514	2.35	16.91	1	7.3	0
08:13	100	337	2.46	16.8	1	8.2	0
09:13	100	413	2.61	16.56	1	8.5	0
10:13	100	255	2.6	16.55	0	8.9	0
11:13	100	265	2.68	16.31	0	9	0
12:13	100	334	2.67	16.28	0	8.7	1
13:13	78	238	2.83	15.89	0	9.5	0
14:13	0	193	2.76	15.92	0	9	0
15:13	8	157	2.65	16.07	0	6.8	0
16:13	7	353	2.66	16	0	6.7	0
17:13	39	467	2.76	15.81	0	5.9	4
18:13	100	514	2.37	15.88	0	1.4	7
19:13	100	514	2.04	16.03	0	-0.4	8
20:13	100	170	0.9	17.82	0	-0.4	15
21:13	100	120	0.77	18.29	0	-0.3	0
22:13	32	114	0.78	18.32	0	-0.3	1
23:13	15	140	0.79	18.32	1	-0.3	0

March 16, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate MBTU/hr	Coal firing rate MBTU/hr	Total firing rate MBTU/hr	Steam flow rate, lbs/hr	Exp.steam flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs exit.temp. F	Stack temp., F
00:13							7275		2627	6104	1.22	627	733	288	227	241
01:13	17725	0	775	18.43	0.00	18.43	9531	6708	2627	6496	1.27	633	749	291	224	239
02:13	17710	0	770	18.42	0.00	18.42	8283	7031	2627	6255	1.24	629	761	288	227	243
03:13	13892	0	0	14.45	0.00	14.45	0	5302	290	0	0.05	478	588	263	202	207
04:13	0	0	0	0.00	0.00	0.00	0	0	278	0	0.05	423	536	220	182	146
05:13	0	0	0	0.00	0.00	0.00	0	0	272	0	0.05	396	488	189	160	121
06:13	0	0	0	0.00	0.00	0.00	0	0	272	0	0.05	380	453	168	143	104
07:13	0	0	0	0.00	0.00	0.00	0	0	278	0	0.05	366	425	152	129	91
08:13	0	0	0	0.00	0.00	0.00	0	0	229	0	0.04	359	402	139	118	85
09:13	0	0	0	0.00	0.00	0.00	0	0	197	1330	0.22	352	381	129	109	86
10:13	0	0	0	0.00	0.00	0.00	0	0	169	0	0.04	344	362	122	102	85
11:13	0	0	0	0.00	0.00	0.00	0	0	197	305	0.08	337	346	115	97	83
12:13	0	0	0	0.00	0.00	0.00	0	0	239	1727	0.28	330	332	110	95	81
13:13	0	0	0	0.00	0.00	0.00	0	0	232	0	0.04	323	319	107	95	80
14:13	0	0	0	0.00	0.00	0.00	0	0	243	0	0.05	316	307	104	93	80
15:13	1748	0	0	1.82	0.00	1.82	0	0	236	0	0.04	309	296	102	90	80
16:13	2160	0	0	2.25	0.00	2.25	0	0	239	0	0.05	303	285	101	90	82
17:13	2442	0	0	2.54	0.00	2.54	0	0	236	0	0.04	296	275	100	90	82
18:13	567	0	0	0.59	0.00	0.59	0	0	256	0	0.05	290	266	99	87	78
19:13	0	0	0	0.00	0.00	0.00	0	0	246	0	0.05	264	210	125	112	89
20:13	0	0	0	0.00	0.00	0.00	0	0	266	0	0.05	249	182	135	122	97
21:13	0	0	0	0.00	0.00	0.00	0	0	260	0	0.05	241	169	142	129	101
22:13	0	0	0	0.00	0.00	0.00	0	0	272	0	0.05	234	159	146	129	102
23:13	0	0	0	0.00	0.00	0.00	0	0	275	0	0.05	227	149	150	129	103

March 16, 1997 EMISSIONS DATA

Time	Stack	Stack emissions					
	opacity,	CO	CO2	O2	SO2	NOx	HC
	%	ppm	%	%	ppm	ppm	ppm
00:13	19	107	0.79	18.46	1	0	0
01:13	20	142	0.81	18.47	1	0	0
02:13	19	112	0.81	18.54	2	0	0
03:13	13	29	0.44	20.08	2	0	0
04:13	18	3	0.14	21.2	2	0	0
05:13	18	3	0.14	21.25	3	0	0
06:13	20	3	0.14	21.29	3	0.1	0
07:13	23						
08:13	100	4	0.14	20.91	6	0.3	0
09:13	100	4	0.14	20.84	3	0	0
10:13	100	3	0.13	20.76	3	0	0
11:13	100	3	0.13	20.68	2	0	0
12:13	100	3	0.13	20.6	2	0	0
13:13	100	3	0.12	20.5	1	0	0
14:13	100	3	0.12	20.42	0	0	0
15:13	67	2	0.11	20.37	0	0	0
16:13	0	2	0.11	20.35	0	0	0
17:13	100	2	0.11	20.38	0	0	0
18:13	9	2	0.11	20.39	0	0	0
19:13	14	2	0.11	20.37	0	0	0
20:13	16	2	0.11	20.42	0	0	0
21:13	56	2	0.12	20.48	0	0	0
22:13	16	2	0.12	20.54	0	0	0
23:13	15	2	0.13	20.59	0	0	0

March 17, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate MBTU/hr	Coal firing rate MBTU/hr	Total firing rate MBTU/hr	Steam flow rate, lbs/hr	Exp.steam flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs exit.temp. F	Stack temp., F
00:13							0		272	0	0.05	221	142	152	125	102
01:13	0	0	0	0.00	0.00	0.00	0	0	272	0	0.05	216	134	152	124	101
02:13	0	0	0	0.00	0.00	0.00	0	0	269	0	0.05	210	127	151	125	99
03:13	0	0	0	0.00	0.00	0.00	0	0	278	0	0.05	205	122	148	123	97
04:13	0	0	0	0.00	0.00	0.00	0	0	284	0	0.05	200	116	145	120	95
05:13	0	0	0	0.00	0.00	0.00	0	0	272	0	0.05	196	112	143	116	91
06:13	0	0	0	0.00	0.00	0.00	0	0	272	0	0.05	191	108	140	113	89
07:13	0	0	0	0.00	0.00	0.00	0	0	272	0	0.05	188	105	136	108	86
08:13	0	0	0	0.00	0.00	0.00	0	0	213	0	0.04	185	104	132	104	85
09:13	0	0	0	0.00	0.00	0.00	0	0	164	1221	0.20	187	116	112	91	78
10:13	0	0	0	0.00	0.00	0.00	0	0	179	0	0.03	187	125	100	82	75
11:13	591	0	0	0.61	0.00	0.61	0	0	188	610	0.12	186	122	101	84	76
12:13	1242	0	0	1.29	0.00	1.29	0	0	232	1806	0.29	183	120	99	85	77
13:13	1	0	0	0.00	0.00	0.00	0	0	221	0	0.04	181	118	98	86	78
14:13	0	0	0	0.00	0.00	0.00	0	0	239	0	0.04	179	116	99	87	80
15:13	0	0	0	0.00	0.00	0.00	0	0	221	0	0.04	176	115	98	87	82
16:13	0	0	0	0.00	0.00	0.00	0	0	236	0	0.04	174	113	98	88	83
17:13	0	0	0	0.00	0.00	0.00	0	0	232	0	0.04	172	112	97	86	82
18:13	0	0	0	0.00	0.00	0.00	0	0	236	0	0.04	169	111	91	82	76
19:13	0	0	0	0.00	0.00	0.00	0	0	246	0	0.04	167	110	86	79	72
20:13	0	0	0	0.00	0.00	0.00	0	0	246	0	0.04	165	109	84	77	69
21:13	0	0	0	0.00	0.00	0.00	0	0	246	0	0.04	162	108	81	74	65
22:13	0	0	0	0.00	0.00	0.00	0	0	243	0	0.04	160	107	78	72	63
23:13	0	0	0	0.00	0.00	0.00	0	0	246	0	0.04	158	107	76	70	62

March 17, 1997 EMISSIONS DATA

Time	Stack opacity, %	Stack emissions					
		CO ppm	CO2 %	O2 %	SO2 ppm	NOx ppm	HC ppm
00:13	15	2	0.13	20.64	0	0	0
01:13	15	2	0.13	20.7	0	0	0
02:13	16	2	0.13	20.75	0	0	0
03:13	16	2	0.14	20.78	0	0	0
04:13	15	2	0.14	20.8	0	0	0
05:13	17	3	0.14	20.82	0	0	0
06:13	15	2	0.14	20.87	0	0	0
07:13	18	2	0.14	20.88	0	0	0
08:13	100	2	0.14	20.88	0	0	0
09:13	100	3	0.14	20.84	0	0	0
10:13	100	2	0.13	20.65	0	0	0
11:13	100	2	0.1	20.5	0	0	0
12:13	100	2	0.1	20.37	0	0	0
13:13	0	2	0.11	20.33	0	0	0
14:13	7	2	0.11	20.31	0	0	0
15:13	10	2	0.09	20.67	0	0	0
16:13	14	2	0.1	20.85	0	0	0
17:13	6	1	0.1	20.86	0	0	0
18:13	9	1	0.1	20.85	0	0	0
19:13	10	1	0.1	20.83	0	0	0
20:13	9	1	0.1	20.81	0	0	0
21:13	10	2	0.1	20.82	0	0	0
22:13	10	1	0.1	20.84	0	0	0
23:13	10	2	0.1	20.87	0	0	0

March 18, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate MBTU/hr	Coal firing rate MBTU/hr	Total firing rate MBTU/hr	Steam flow rate, lbs/hr	Exp.steam flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs exit.temp. F	Stack temp., F
00:13							0		253	0	0.04	156	105	75	69	61
01:13	0	0	0	0.00	0.00	0.00	0	0	256	0	0.05	154	105	73	67	61
02:13	0	0	0	0.00	0.00	0.00	0	0	260	0	0.05	152	103	72	66	60
03:13	0	0	0	0.00	0.00	0.00	0	0	246	0	0.04	150	103	70	65	60
04:13	0	0	0	0.00	0.00	0.00	0	0	256	0	0.04	147	101	69	64	59
05:13	0	0	0	0.00	0.00	0.00	0	0	256	0	0.04	145	100	68	63	58
06:13	0	0	0	0.00	0.00	0.00	0	0	243	0	0.04	144	100	67	63	59
07:13	0	0	0	0.00	0.00	0.00	0	0	256	0	0.04	142	99	67	62	57
08:13	0	0	0	0.00	0.00	0.00	0	0	243	0	0.04	140	98	66	62	58
09:13	0	0	0	0.00	0.00	0.00	0	0	232	0	0.04	138	97	67	62	57
10:13	0	0	0	0.00	0.00	0.00	0	0	239	3675	0.55	133	97	84	72	80
11:32	0	0	0	0.00	0.00	0.00	0	0	221	0	0.04	112	128	103	91	98
12:32	0	0	0	0.00	0.00	0.00	0	0	201	0	0.04	110	125	95	85	88
13:32	0	0	0	0.00	0.00	0.00	0	0	217	0	0.04	110	123	90	82	83
14:32	0	0	0	0.00	0.00	0.00	0	0	201	0	0.04	110	122	88	82	83
15:32	0	0	0	0.00	0.00	0.00	0	0	201	0	0.04	110	123	86	82	84
16:32	0	0	0	0.00	0.00	0.00	0	0	213	0	0.04	110	122	86	81	82
17:32	0	0	0	0.00	0.00	0.00	0	0	221	0	0.04	109	119	86	82	82
18:32	0	0	0	0.00	0.00	0.00	0	0	221	0	0.04	108	116	85	80	78
19:32	0	0	0	0.00	0.00	0.00	0	0	221	0	0.04	107	115	84	78	75
20:32	0	0	0	0.00	0.00	0.00	0	0	246	0	0.04	106	114	82	77	72
21:32	0	0	0	0.00	0.00	0.00	0	0	243	0	0.04	106	113	81	75	69
22:32	0	0	0	0.00	0.00	0.00	0	0	239	0	0.04	106	112	79	73	68
23:32	0	0	0	0.00	0.00	0.00	0	0	250	0	0.04	105	111	78	72	67

March 18, 1997 EMISSIONS DATA

Time	Stack opacity, %	Stack emissions					
		CO ppm	CO2 %	O2 %	SO2 ppm	NOx ppm	HC ppm
00:13	10	2	0.11	20.89	0	0	0
01:13	9	2	0.1	20.88	0	0	0
02:13	10	2	0.11	20.9	0	0	0
03:13	11	2	0.11	20.9	0	0	0
04:13	10	2	0.11	20.92	0	0	0
05:13	9	1	0.11	20.91	0	0	0
06:13	10	2	0.12	20.97	0	0	0
07:13	1	2	0.11	20.97	2	0	0
08:13	17	2	0.12	20.97	0	0	0
09:13	11	2	0.12	20.95	0	0	0
10:13	100	1	0.11	20.92	0	0	0
11:32	0	2	0.11	20.85	0	0	0
12:32	0	1	0.1	20.79	0	0	0
13:32	0	1	0.1	20.72	0	0	0
14:32	0	1	0.1	20.71	0	0	0
15:32	0	1	0.1	20.75	0	0	0
16:32	0	0	0.1	20.75	0	0	0
17:32	0	0	0.09	20.75	0	0	0
18:32	0	1	0.09	20.75	0	0	0
19:32	0	1	0.09	20.69	0	0	0
20:32	0	1	0.09	20.71	0	0	0
21:32	0	1	0.09	20.72	0	0	0
22:32	0	1	0.09	20.75	0	0	0
23:32	0	1	0.1	20.75	0	0	0

March 19, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate MBTU/hr	Coal firing rate MBTU/hr	Total firing rate MBTU/hr	Steam flow rate, lbs/hr	Exp.steam flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs exit.temp. F	Stack temp., F
00:32							0		246	0	0.04	105	110	77	72	66
01:32	0	0	0	0.00	0.00	0.00	0	0	250	0	0.04	104	109	76	71	65
02:32	0	0	0	0.00	0.00	0.00	0	0	239	0	0.04	104	108	75	70	64
03:32	0	0	0	0.00	0.00	0.00	0	0	239	0	0.04	103	107	74	69	64
04:32	0	0	0	0.00	0.00	0.00	0	0	250	0	0.04	103	106	73	68	63
05:32	0	0	0	0.00	0.00	0.00	0	0	256	0	0.04	102	106	72	67	61
06:32	0	0	0	0.00	0.00	0.00	0	0	243	0	0.04	102	105	71	66	61
07:32	0	0	0	0.00	0.00	0.00	0	0	256	0	0.04	101	104	71	65	61
08:32	0	0	0	0.00	0.00	0.00	0	0	243	0	0.04	101	103	68	64	61
09:32	0	0	0	0.00	0.00	0.00	0	0	246	0	0.04	100	102	69	65	62
10:32	0	0	0	0.00	0.00	0.00	0	0	239	0	0.04	99	101	69	65	57
11:32	0	0	0	0.00	0.00	0.00	0	0	232	0	0.04	98	99	69	65	55
12:32	0	0	0	0.00	0.00	0.00	0	0	236	0	0.04	111	125	70	65	57
13:32	0	0	0	0.00	0.00	0.00	0	0	256	0	0.04	114	129	71	65	57
14:32	0	0	0	0.00	0.00	0.00	0	0	1667	6729	1.17	101	177	152	120	143
15:32	9427	0	339	9.80	0.00	9.80	0	0	2593	7306	1.38	354	529	236	194	196
16:32	16962	0	636	17.64	0.00	17.64	0	4	2627	7313	1.38	529	621	342	305	293
17:32	17810	0	637	18.52	0.00	18.52	0	4843	2627	7093	1.35	549	630	286	226	241
18:32	18239	0	637	18.97	0.00	18.97	0	769	2588	6980	1.33	554	630	290	231	245
19:32	17631	0	637	18.34	0.00	18.34	2139	10372	2564	6900	1.32	560	630	293	232	246
20:32	16610	0	572	17.27	0.00	17.27	0	1703	2566	5710	1.16	584	708	282	229	244
21:32	24085	683	20	25.05	8.95	34.00	24332	7749	2562	5718	1.16	839	1325	313	233	245
22:32	19668	1605	541	20.46	21.02	41.47	27686	36039	2247	6652	1.25	1200	1285	301	237	253
23:32	19069	1465	581	19.83	19.19	39.02	44180	26349	2568	6652	1.30	1300	1501	314	241	257

March 19, 1997 EMISSIONS DATA

Time	Stack	Stack emissions					
	opacity,	CO	CO2	O2	SO2	NOx	HC
	%	ppm	%	%	ppm	ppm	ppm
00:32	0	1	0.1	20.75	0	0	0
01:32	0	1	0.09	20.75	0	0	0
02:32	0	1	0.1	20.71	0	0	0
03:32	0	1	0.1	20.75	0	0	0
04:32	0	1	0.1	20.75	0	0	0
05:32	0	2	0.1	20.75	0	0	0
06:32	0						
07:32	0	1	0.1	20.84	0	0	0
08:32	9	2	0.1	20.83	0	0	0
09:32	0	1	0.1	20.82	0	0	0
10:32	0	2	0.1	20.81	0	0	0
11:32	0	2	0.1	20.78	0	0	0
12:32	0	1	0.1	20.75	0	0	0
13:32	0	1	0.1	20.71	0	0	0
14:32	59	25	0.32	20.29	0	0.4	0
15:32	100	45	1.77	17.26	0	10.6	0
16:32	45	79	1.91	16.78	0	13.1	0
17:32	13	64	1.84	16.85	0	13.1	0
18:32	7	55	1.86	16.82	0	13.5	0
19:32	8	43	1.87	16.79	0	15	0
20:32	0	60	2.12	16.23	1	15.3	6
21:32	11	514	6.01	8.46	166	95	1029
22:32	0	154	4.09	13.89	0	108.7	11
23:32	0	37	7.39	8	0	86.8	11

March 20, 1997 PERFORMANCE DATA

Time	Nat. gas flow rate, ft^3/hr	Coal feed rate, lbs/hr	Sorbent feed rate, lbs/hr	Nat. gas firing rate MBTU/hr	Coal firing rate MBTU/hr	Total firing rate MBTU/hr	Steam flow rate, lbs/hr	Exp.steam flow rate, lbs/hr	PC air, SCFM	Fluidiz. air, SCFM	Bed velocity, ft/sec	Bed temp., F	Freebrd temp., F	Ecnmzr temp., F	Baghs exit.temp. F	Stack temp., F
00:32							45015		2606	6893	1.34	1407	1483	312	242	258
01:32	17051	1946	596	17.73	25.49	43.22	47563	45871	2583	6811	1.32	1460	1551	315	243	260
02:32	16487	1776	707	17.15	23.27	40.42	42897	46380	2571	7420	1.41	1413	1508	324	249	266
03:32	16502	1695	858	17.16	22.20	39.37	43382	43419	2588	6987	1.35	1411	1488	313	242	258
04:32	16471	1727	858	17.13	22.63	39.76	43003	44098	2560	6927	1.34	1398	1479	306	239	256
05:32	15583	1698	858	16.21	22.24	38.45	44121	43105	2540	6791	1.31	1403	1479	308	238	255
06:32	16107	1895	737	16.75	24.83	41.58	47398	45566	2561	7569	1.43	1416	1486	314	244	260
07:32	15247	2028	517	15.86	26.57	42.42	47769	44855	2561	7740	1.45	1452	1530	324	250	267
08:32	14782	2080	710	15.37	27.25	42.62	34713	42833	2551	7060	1.35	1259	1349	304	241	258
09:32	18312	1746	0	19.04	22.88	41.92	43913	0	2563	7382	1.40	1366	1542	317	248	264
10:32	0	1943	233	0.00	25.45	25.45	43329	0	2562	7287	1.39	1352	1504	319	250	267
11:32	0	2131	443	0.00	27.92	27.92	47315	0	2531	7912	1.47	1441	1525	323	255	273
12:32	0	2138	490	0.00	28.01	28.01	45949	0	1667	8584	1.44	1530	1558	305	247	266
13:32	14339	1346	526	14.91	17.64	32.55	25994	35884	2627	8714	1.58	758	1081	305	249	268
14:32	13503	1707	526	14.04	22.36	36.41	38472	36441	1622	8303	1.40	1352	1454	308	247	264
15:32	9392	1949	526	9.77	25.54	35.30	37708	37531	1501	8070	1.35	1320	1485	310	246	263
16:32	9388	2131	526	9.76	27.91	37.68	40499	40447	1469	7918	1.32	1348	1480	308	246	263
17:32	9173	2232	526	9.54	29.24	38.78	43449	40897	1461	7994	1.33	1373	1490	308	245	262
18:32	8553	2236	526	8.90	29.29	38.18	45312	43063	1456	8392	1.39	1366	1497	310	243	259
19:32	8641	2247	526	8.99	29.43	38.42	40386	43204	1451	8403	1.39	1313	1472	313	244	261
20:32	8796	2309	526	9.15	30.25	39.40	40940	41681	1445	8348	1.37	795	1479	310	241	258
21:32	8806	2328	526	9.16	30.50	39.66	34496	42634	1688	8671	1.46	1381	927	295	236	253
22:32	14487	2324	526	15.07	30.44	45.51	50119	45985	2627	7080	1.35	622	1464	316	247	264
23:32	11933	1477	281	12.41	19.35	31.76	25499	38038					797	271	217	236

March 20, 1997 EMISSIONS DATA

Time	Stack	Stack emissions					
	opacity,	CO	CO2	O2	SO2	NOx	HC
	%	ppm	%	%	ppm	ppm	ppm
00:32	0	40	8.28	7.62	0	106.2	2
01:32	14	44	9.14	7.14	0	162.8	0
02:32	0	56	6.78	9.43	0	232.1	0
03:32	10	35	7.38	8.61	14	133.2	0
04:32	8	32	7.5	8.64	107	115.1	0
05:32	17	33	7.82	7.92	212	80	0
06:32	12						
07:32	15	31	7.96	8.31	265	113.4	0
08:32	13	113	4.65	13.1	121	57.8	4
09:32	12	20	7.27	8.96	262	92.5	0
10:32	11	25	7.59	8.79	317	70	0
11:32	12	22	7.68	8.85	304	108.7	0
12:32	10	22	8.53	8.27	316	153.4	0
13:32	10	514	2.71	15.03	103	20.6	1029
14:32	11	46	7.65	9.14	300	60.7	0
15:32	11	42	8.09	8.55	347	49.3	0
16:32	53	39	8.21	8.47	352	51.4	0
17:32	11	36	8.6	8.1	365	55.6	0
18:32	12	38	8.19	8.75	323	61.5	0
19:32	12	41	8.34	8.23	358	51.2	0
20:32	12	38	8.96	7.09	408	49.5	0
21:32	14	158	3.53	14.73	152	54.3	2
22:32	15	58	8.19	8.46	279	46.8	3
23:32	15	514	2.12	16.39	68	15.6	250

SECTION 3.0

MARKET ANALYSIS

The impact of the Coal Fired Pulse Combustion technology pioneered by MTCI and ThermoChem on overall energy usage in the United States and in world markets through the year 2030 could be enormous. At the current state of its development, coal-fired pulse combustors will shortly be marketable in several forms. This market analysis is focused on the potential industrial and commercial markets for the PAFBC.

The Pulsed Atmospheric Fluidized Bed Combustion (PAFBC) technology will have extensive applications both in industrial and large-scale commercial applications in both its adiabatic and non-adiabatic versions. In its non-adiabatic (steam raising) version, it has the potential to replace oil, gas or existing coal-fired boilers in the entire range from 10,000 pph of steam to 250,000 pph or more.

In its adiabatic (hot gas raising) version, it has immediate applications in the coal mining industry where it offers the industry the opportunity of using low-grade coals and problematic coal fines to dry more marketable coals cheaply. It has widespread potential as a means for repowering oil- or gas-fired boiler systems which are failing to meet EPA standards. It offers industry an inexpensive means to dry or dewater troublesome sludges of all kinds, including municipal sludges, and to provide hot air for industrial drying processes in many industries including the cement industry, the lumber and wood products industries, the food industry and many more.

The initial commercial versions of this technology will be fabricated with modules sized with 8 ft. x 8 ft. bed area fluid beds or multiples thereof. Each module will generate approximately 33 MMBtu/hr of steam in the non-adiabatic version, and 16 MMBtu/hr of hot air in the adiabatic version. The marketing strategy envisions initial market entry using the 8 ft. x 8 ft. modules to establish marketplace credibility for the technology, and maximize the benefits of "economies of scale" and mass production of modules, and at the same time to take advantage of established coal using and handling infrastructure.

The first immediate applications will be in providing cheap hot gas for coal drying in the coal mining industry (adiabatic) and in providing a primary heating source for very large buildings with more than 750,000 square feet of floor area requiring approximately 33,000 pph steam or more in the commercial market (non-adiabatic). Once the technology is established at that scale, marketing strategy will begin to target the segment of the commercial boiler market comprised of buildings ranging in size from 50,000 square feet to 750,000 square feet. Smaller units will be designed to satisfy that portion of the market. Module sizes will include a 4 ft. x 4 ft. bed area generating 8,000 pph, and a 6 ft. x 6 ft. bed area generating 18,000 pph steam.

Expected market applications for the technology include the following:

➢ ***Non-Adiabatic*** ***(steam raising)***

- Replacement of oil, gas, propane, and existing coal-fired boilers in the 10,000 pph to 250,000 pph steam range in the commercial and industrial markets.
- Greenfield installations of new boilers in the same range.
- Replacement of existing steam boilers in district steam heating distribution networks.
- Greenfield installations for district steam heating expansion projects.
- New boilers for industrial cogeneration plants.

➢ ***Adiabatic*** ***(hot gas raising)***

- Replacement of existing thermal dryer heat sources in the coal mining industry.
- Greenfield installations of new thermal dryers. In many situations the prohibitive cost of oil or gas for thermal drying has removed thermal drying from being a viable option for mining companies. By using their low low-grade coal and coal fines as a heat source, many mines will be able to increase their output and revenues without being subject to the rise and fall of oil and gas prices. This will require the interfacing of the PAFBC with existing or current dryer equipment.

- Installation of PAFBC units for repowering of single or multiple oil or gas-fired or coal-fired boilers which are in a state of non-compliance with EPA standards controlling atmospheric release of NO_x and SO_2. The PAFBC would be centrally placed to feed super-hot gas to the boiler heat exchangers which could be augmented by the firing of oil or gas in the individual boilers. This will require interfacing of the export hot gas ducts of the PAFBC with existing boiler installations.

In addition to the above, there are hundreds of other possible applications in the dryer and oven markets including the following:

- Delacquering furnaces and solution heat treating lines in the aluminum industry.
- Tenter dryers in the textile industry.
- Preheating in many industries.
- Core and mold baking in the foundry industry.
- Plastic curing.
- Annealing.
- Aging, baking, drying, curing processes in many industries.
- Roasting and sterilizing processes.
- Sludge dewatering in rotary dryers of paper mill sludges, municipal sludges and many other industrial sludges.
- Rotary dryers in the rubber industry.
- Indirect heat rotary calcination applications including:
    - Activating wood charcoal
    - Reducing mineral high oxides to low oxides
    - Calcination of silica gel
    - Reduction of metal oxides
    - Oxidizing and "burning-off" or organic impurities
    - Reclamation of foundry sand from the shell-molding process.

Industrial Markets

The first target is the replacement of existing boiler systems and/or hot gas industrial process systems which are currently using coal and which are either approaching the end of their useful life or in a state of non-compliance with existing EPA air-quality regulations. In these cases, the infrastructure is already in place. There is an existing coal supply, an established transportation system, and bulk coal storage and handling facilities. These will tend to be concentrated along the Eastern Seaboard and in the Mid-West. Prominent among these will be coal-mining operations presently using thermal dryers and district heating systems. In a similar category will be the installation of new greenfield boilers and/or hot gas industrial process systems and coal-mining thermal dryers in the same general areas of the country where coal usage in large facilities is an accepted norm rather than an exception. Also in this category will be the repowering of single or multiple boiler systems which are in non-compliance with EPA standards and which are presently using oil or gas, but which are in an area where coal usage is common.

The development of new industrial markets will require introducing the concept of using coal in a wide variety of industrial processes which presently use oil or gas to produce hot gas for aging, drying, baking, curing, oxidizing, calcining, preheating, dewatering, sterilizing, annealing, etc. This market will evolve during the second phase of marketing strategy.

One of the immediate applications for the PAFBC in its adiabatic version will be the immediate replacement of existing thermal dryer heat sources in the coal-mining industry in coal preparation plants. At the present time, according to the 1991 edition of the Keystone Coal Industry Manual, and a mine census summary published in the November 1989 issue of *Coal,* by Mark Sprouls, publisher, there are approximately 416 coal preparation plants in the United States. Table 3-1 shows the distribution of coal preparation plants by state (United States) and province (Canada).

TABLE 3-1:

COAL PREPARATION PLANTS BY STATE AND PROVINCE

Alabama	15	Kansas	2	Pennsylvania:	
Alberta	5	Kentucky	66	Anthracite	35
British Columbia	8	Maryland	1	Bituminous	0
Colorado	5	Missouri	5	Tennessee	3
Georgia	1	New Mexico	1	Texas	1
Illinois	36	Nova Scotia	1	Utah	6
Indiana	2	Ohio	1	Virginia	34
Iowa	3	Oklahoma	1	Washington	3
				West Virginia	95

Capacity data was provided for 257 plants. Of these, 58 percent have design capacities of 500 tph or more (*Table 3-2* below). Most of those big plants fall into the 500 tph to 999 tph range. Only 13 percent of the plants produce 1,000 tph or more.

TABLE 3-2:

PREPARATION PLANT CAPACITY

Number of Plants*	Capacity (tph)
4	> 2,000
12	1,500 - 1,999
43	1,000 - 1,499
89	500 - 999
69	250 - 499
40	< 250

* Of those reporting.
Source: *Coal* magazine, November 1989.

Of the total number of U.S. preparation plants, somewhere between 99 and 150 plants presently operate dryers. The ability of the PAFBC to use relatively low-grade coals to provide hot gas to dry better grade coals should prove to be very attractive to the mining industry. It will offer the possibility of limiting the growth of existing gob piles, while at the same time significantly increasing mine output in an environmentally acceptable manner. Ultimately, it will also provide a means to put to profitable use the lower grade coal inventories which gob piles represent. This will involve the development of a practical method for recycling gob pile coals, and the initial marketing effort will be aimed at thermal dryer replacement with PAFBC technology wherever possible.

Island Creek Coal Corporation has several coal preparation plants in the United States which are presently using thermal dryers in addition to the Alpine plant which is the host site for the Phase III Field Test portion of this project. They have a total of seven thermal dryers using conventional fluidized beds as a heat source, burning high Btu coals which could otherwise be sold for a profit. The attraction of the PAFBC adiabatic technology to Island Creek Coal Corporation is that it allows the use of refuse coal, which is unmarketable, to dry better quality coals which are marketable, while at the same time reducing air pollutants. Island Creek can see the potential for the PAFBC to make a major contribution towards corporate profitability once it has been successfully demonstrated. Their willingness to proceed with installation in other plants is strictly dependent upon the success of the subject demonstration project.

Island Creek has several other plants in the United States at which the addition of greenfield coal-fired coal dryers might prove attractive once the economics have been demonstrated.

The industrial sector as a whole is the largest end-use consumer of energy. The total industrial sector energy use is projected to be 18.4 quadrillion Btu in 1995, and 30.3 quads in 2030. In 1985, total energy consumption in the industrial sector was 16.4 quads *(Source: DOE/Revised NES No Further Policy Action Case - 12-14-90 and CCT-IV PON)*.

In a study undertaken in 1986 by Burns & Roe, a methodology for determining the population and size of industrial boilers (< 50 MMBtu) and direct-fired combustors (< 100 MMBtu/hr) was devised. This in turn was based on a study by PEI Associates, Inc. and P.W.

Spaite Company in 1985. However, for the purposes of this report, the values for potential market sector will be based on those reported values as shown in *Table 3-3*. This table shows the overall boiler and direct-fired combustor populations and energy consumption revealed by the study.

TABLE 3-3:

TOTAL ENERGY DISTRIBUTION
(Trillion Btu)

Direct Fired Combustors (< 100 MMBtu/hr)

Size Range (MMBtu)	Population (Number)	Energy (10^{12} Btu)
1 - 9	14,882	447.19
10 - 24	3,816	391.53
25 - 49	1,529	350.59
50 - 99	959	448.47
Total	*21,186*	*1,638.78*

Boilers (50 MMBtu/hr)

Size Range (MMBtu)	Population (Number)	Energy (10^{12} Btu)
1 - 9	34,678	105.00
10 - 24	13,211	140.00
25 - 49	7,193	246.00
Total	*55,082*	*491.00*

Total represents sum of distillates, residual and natural gas.

Source: Burns and Roe Services Corporation, "Marketing and Equipment Performance Analysis for the Application of Coal-Based Fuels/Advanced Combustion Systems," March 1986.

If it is assumed that this same distribution will prevail as new units are installed, then it is possible to break down by size the growth in the industrial sector as projected by DOE and thereby identify the potential markets for coal-fired PAFBCs replacing gas and oil in the 10 to 100 MMBtu/hr range. By the year 2030, the possible replacement is 1.8 quads. In addition, the DOE "No Further Policy Action Case" projects a growth of 2.4 quads in coal use which also represents a potential market for PAFBCs.

Although coal was once the predominant fuel in the residential, commercial and industrial markets, it has not occupied that position for decades. Although there was once a fairly extensive coal distribution system servicing those market sectors, it no longer exists except in the more heavily industrialized areas of the country. A market entry scenario for a coal technology should therefore primarily be concerned with the largest (most dense) geographic areas of population, industry, commerce, coal production, heating demand, and, finally, a tradition of continued coal use or acceptance of coal use. Obviously, more detailed analysis of the factors can probably locate centers of opportunity in other areas of the country where the demographics are supportive of coal utilization.

Commercial Markets

This market segment includes that portion of the 6,000 or so commercial buildings in the United States which are larger than 500,000 square feet and which presently are using coal as their heating fuel of choice, or which are presently using oil or gas but are in an area of the country where coal usage is an accepted norm. This will encompass replacement of existing units and greenfield applications.

The future commercial market is composed of buildings ranging in size from 50,000 square feet to 750,000 square feet. This segment will also be addressed in the second phase of marketing strategy and will require a considerable shift in marketplace mentality, or drastic price rises in competing fuels before it will gain widespread acceptance.

In mid-1986, there were 4,154,000 commercial buildings in the United States. The total floor space in these buildings was 58.2 billion square feet. These buildings consumed energy for six end uses: space heating, water heating, cooling, manufacturing, and electricity generation. The type, location, age, and size of the buildings affected the amount and the source of energy used.

EIA reports that the source of energy supplied to commercial buildings also varies with building size. Electricity was supplied to 95.8 percent of all buildings containing 5,000 square feet or less. It was supplied to 100 percent of all buildings of more than 200,000 square feet. Natural gas was used in 49.1 percent of the buildings containing 5,000 square feet or less and in 75.9 percent of buildings containing more than 200,000 square feet. A negligible amount of district steam or hot water was used in buildings containing 5,000 square feet or less, but district steam or hot was used in 17.3 percent of all buildings of more than 200,000 square feet, which amounted to 19.5 percent of the total floor space in that size range.

The majority of commercial buildings are small in area. Over 70 percent contain less than 10,000 square feet and over 47 percent are smaller than 5,000 square feet.

A simple attempt at disaggregating the commercial sector on the basis of square footage and the average energy demand to provide average sizes in equivalent Btu/hr steam systems was made. An assumption was made using an estimated average peak demand value of 54 Btu/hr per square foot in determining the size of peak heating demand for a building of average square footage in each size range. Actual regional values range from 75 Btu/hr in the Northeast to 30 Btu/hr in the South. The number of buildings includes only those buildings using oil or gas for space heating.

This analysis indicates that buildings with over 100,000 square feet represent a significant potential market for coal-fired PAFBC systems. The 8 ft. x 8 ft. PAFBC is sized to generate approximately 33 MMBtu/hr; the 6 ft. x 6 ft. - 18 MMBtu/hr; the 4 ft. x 4 ft. - 8 MMBtu/hr. The marketing approach for the PAFBC in the commercial market will be initially to target the segment containing the largest buildings with square footages in excess of 500,000 square feet using the 8 ft. x 8 ft. non-adiabatic module as the vehicle for market entry. This size range will provide optimum benefits of economies of scale in terms of unit cost, and very attractive cost savings to the end user in fuel costs, and there should as a result be less of a need for market preconditioning to the concept of using clean coal technology instead of conventional fuels.

There are approximately 6,000 buildings in the United States presently using oil or gas for heat which fall within the target size range. A significant number of these buildings are in the government sector where government priorities will tend to encourage clean coal technology as it evolves. The PAFBC would be applicable to existing buildings as their present boiler systems reach 30 or more years of use, and would be immediately applicable to new construction projects in this size range. The salability of the technology will inevitably relate to the availability and proximity of coal suppliers, and in consequence, it is anticipated that the primary immediate commercial market area will be in the North-Eastern, Mid-Western and Mid-Atlantic state regions of the United States.

Market Summary

The analysis presented above indicates that a market for the PAFBC exists in the industrial and commercial sectors if the technology is demonstrated to be technically and economically feasible. The economic analysis presented in the previous section indicated that the PAFBC has the potential to be economically competitive with conventional technologies. The field tests planned for Phase II are designed to demonstrate the technical performance and verify the economic projections for the technology.

In marketing a new coal-based technology, several institutional barriers must be overcome. We have developed a marketing strategy to focus our initial commercialization efforts in markets in which coal is considered to be a desirable fuel. For this reason, the initial application of using the PAFBC to supply heat for coal drying is highly desirable - the customer obviously has a desire to demonstrate that coal is an economically and environmentally attractive fuel. Once this application has demonstrated success, we can expand into more traditional markets for process heat and for steam.

SECTION 4.0

ENGINEERING AND ECONOMIC EVALUATION

4.1 ENGINEERING EVALUATION

Bubbling atmospheric fluid-bed combustors have a number of technical attributes that hold a great deal of promise in combustion of high sulfur, high ash coals having fuel-bound nitrogen, with inherent pollution control capabilities.

The fundamental concept of a bubbling atmospheric fluid-bed combustion is to burn crushed coal at a moderate temperature in a bed comprised of particles that are made of a sulfur sorbent. The ability to burn the coal at a moderate temperature (1550°F to 1650°F) in a fluid bed is aimed at control of NO_x formation from fuel-bound nitrogen and thermal sources. Limestone, generally used as the bed material, calcines at the operating temperature providing a calcium oxide sorbent that captures sulfur oxides produced from burning the sulfur-containing coal fuels.

Heat removal from the bed to control the bed temperature to the desired range is indeed opportune with bubbling fluid beds due to the generally high heat transfer available between the bed material and surfaces immersed in the bed. This provides for relatively small heat transfer surface to extract useful heat from the combustion process.

As the development of atmospheric fluidized beds proceeded, it became evident that coal fines tend to quickly elutriate from the fluid bed that is normally operating at a fluidization air velocity suitable for the bulk of the crushed coal particle size (say, 3/8" x 28 mesh) and the corresponding selection of limestone particle size. Generally, the higher the fluidization velocity, the higher the heat release rate per square foot of bed area. The higher heat release rates are desirable to reduce the specific capital cost of the equipment per Btu/hr of firing rate. Nevertheless, limitation on fluidization velocity was encountered because of residence time, bed height, particle size distribution and related bed stability considerations. Practical considerations and developmental experience gave rise to bubbling fluid beds that operate within the range of 3 to 5 feet in height with a large freeboard required above the bubbling bed.

The freeboard was provided to permit sufficient residence time at a suitable temperature for the combustion of fines that are found with crushed coal as received at the fluid-bed combustor fuel inlet. Coal screening and classification became necessary to remove excessive amounts of fines prior to feeding the fuel to a fluid-bed combustor so as to maintain the operation of the equipment within the environmental performance constraints. This, however, tends to compromise the economics of the operation significantly. Most, if not all, bubbling atmospheric fluid bed manufacturers only warrantee performance with fuel fines content (28 mesh x 0) less than 7 to 10 percent of the total feed.

While bubbling fluid-bed combustors are designed with an appropriate freeboard height and volume with suitable operating temperatures, practical problems still accrue due to the presence of elutriated fines in the freeboard.

The carbon burnout of the fines is not the primary problem because such fines can be captured in a cyclone and/or a baghouse and recycled into the bed, preferably educted with air in an underbed feed, to improve the ultimate carbon burnout. Nevertheless, production of excessive amounts of CO can be experienced from fines burning in the freeboard having insufficiently high gas temperature. Thus, heat removal from the freeboard must be mindful of such considerations. This is particularly important because *gas flow is only once-through the entire system*.

Should the rate of elutriation of fines increase, however, temperatures in the freeboard could rise sufficiently to levels that promote NO_x production in the flue gas even though under such conditions the CO content in the flue tends to decrease. Further increase in the rate of fines elutriation could once more promote production of CO with a modest reduction in NO_x production from combustion in the freeboard.

In addition to the above, fines burning in the freeboard release sulfur oxides downstream of the sorbent bed. Some sulfur capture by sorbent fines elutriating form the bed is encountered but generally not sufficient. This is particularly the case when the coal being fed to the combustor contains excessive amounts of sulfur-laden fines.

Therefore, the design of the freeboard height and volume as well as the allowable heat loss (or removal) from the freeboard must be mindful of the fines combustion considerations as well as the need for a sufficient particle disengagement height for the return of larger particles ejected by the bubbling action at the surface of the bed.

Practical experience with bubbling fluid beds further indicated that gas bubble growth in the bed does occur and ultimately causes channeling of the gas flow through the bed. This tends to compromise both the combustion process efficiency (high excess air detected as high O_2 concentration in the flue) and low sorbent utilization efficiency with sulfur oxides breaking through.

Upstream of the bubbling bed (not downstream of the sorbent) fines combustion maximizes sulfur capture efficiency of the system and sorbent utilization since gas flow through the system is once through.

The PAFBC technology is a hybrid combustion system that employs both a pulse coal combustor and a bubbling fluid bed. In the PAFBC system, the fines (28 mesh x 0) are introduced to the combustion chamber of the pulse combustor. The combustion of the fines in the pulse coal combustor occurs at a very high heat release rate of 2 to 6 MMBtu/hr.cu.ft. The pulse combustor maintains the stoichiometry substantially constant within its design turndown ratio of firing. The pulse coal combustor is the primary new technology in the hybrid PAFBC system.

The dynamic pressure oscillations emanating from the pulse combustor tailpipe impart a forced oscillation to the fluid bed at frequencies in the range of 50 to 60 Hz. This forced oscillation causes large bubbles normally formed in a bubbling bed to break up into smaller bubbles, minimizing gas breakthrough and higher sulfur capture and combustion efficiency in the bed at higher fluidization velocities (and hence higher throughput).

In addition to the above, the fluid-bed combustor design employs an expanded freeboard cross-sectional area to permit shorter freeboard height but with adequate freeboard volume.

Temperature, turbulence and residence time for the pulse combustor and the bubbling fluid-bed freeboard are quite different, as shown below:

	Pulse Combustor	AFBC Freeboard
Temperature	> 1093°C or 2000°F (high)	843°C or 1550°F (low)
Turbulence	Very high (oscillatory)	Moderate (plug flow with backmixing)
Gas Residence Time	10 to 100 milliseconds	2 to 3 seconds

Since the PAFBC unit employs both the pulse combustor and the AFBC technologies, it can handle the full-size range of coarse and fines. The oscillating flow field in the pulse combustor provides for high interphase and intraparticle mass transfer rates. Therefore, the fuel fines essentially burn under kinetic control. Due to the reasonably high temperature, combustion of fuel fines is substantially complete at the exit of the pulse combustor. The additional residence time of 1 to 2 seconds in the freeboard of the PAFBC unit then ensures high carbon conversion and, in turn, high combustion efficiency which was 92 to 97 percent.

The overall heat transfer coefficient in the water-jacketed pulse combustor tailpipes is of the same order as that for tubes immersed in the dense fluidized bed. The replacement of the inefficient heat exchanger in the freeboard of a conventional BFBC by the water-jacketed pulse combustor tailpipes significantly decreases the heat transfer surface area requirement and cost.

4.2 ECONOMIC EVALUATION

It is generally known that large coal-fired facilities are economically superior to those fired with more conventional industrial fuels (oil and natural gas) since the savings in fuel cost realized by the coal-fired system is sufficiently large to recover the additional capital investment required for the more complex coal-fired system.

The firing capacity of the PAFBC unit is 72 MMBtu/hr with a steam generation rate of about 60,000 lb/hr at 120 psig pressure. Following is a review of the major components and systems which are included in the PAFBC unit:

- Coal/Limestone Preparation and Feed System,
- Pulse Combustor,
- Fluidized Bed,
- Boiler Parts, and
- Ash-Handling System.

Coal/Limestone Preparation and Feed System:

1) Coal preparation system:
   Includes coal hopper, coal mill, classifier and bucket elevator with three screw conveyors from hopper to mill, from mill to classifier, and from classifier to bucket elevator.

2) Coarse coal feed system:
   Includes weigh belt, silo and metering screw.

3) Fine coal feed system:
   Includes blower, three eductors, silo and metering screw.

4) Limestone feed system:
   Includes pipeline for pressurized blower truck connection, silo and metering screw.

Pulse Combustor:

The pulse combustor has three aerovalves, combustion chamber, and six tailpipes. The combustion chamber and tailpipes are all water-jacketed and constitute part of the entire PAFBC boiler surface.

Fluidized Bed:

The fluidized bed vessel is constructed in three sections:

1) Air plenum/distributor,
2) Bed with cooling coils, and
3) Freeboard with gas exit on the roof.

The recycling cyclone is a part of the fluidized bed.

Boiler Parts:

The boiler parts include the boiler and economizer.

Ash-Handling System:

The ash-handling system includes two water-cooled stainless steel screw conveyors receiving ash from discharge ports in the bed, ash dropout parts from the boiler chute, two economizer chutes, and two baghouse chutes. It also includes vacuum conveying system and ash silo.

Capital investment data are presented in *Table 4-1*. Economic analysis of the PAFBC system versus the oil-fired boiler system is presented in *Table 4-2*.

TABLE 4-1:
CAPITAL INVESTMENT FOR PAFBC UNIT

COST OF PURCHASED EQUIPMENT

Item	Purchased Cost
PAFBC, including cyclone	$394,000
Boiler	88,400
Economizer	45,280
Coal hopper	6,300
Coal mill	16,000
Classifier	12,500
Screw conveyors	36,000
Bucket elevator	58,500
Coarse coal feed system	15,000
Limestone feed system	15,000
Fine coal feed system	10,300
Coarse coal silo	54,300
Limestone silo	26,300
Fine coal silo	16,200
Ash silo	34,100
Ash-handling system	32,750
Air compressor with dryer	16,810
boiler water pumps	17,046
Fans	48,950
Blowers	16,350
Total	**$960,086**

TABLE 4-1:

CAPITAL INVESTMENT FOR PAFBC UNIT
(Continued)

TOTAL CAPITAL INVESTMENT

Item	Purchased Cost
Direct Costs:	
Purchased equipment (PE)	$960,086
Installation (39% of PE)	374,434
Instrumentation & Controls (13% of PE)	124,811
Piping - installed (31% of PE)	297,627
Electrical - installed (10% of PE)	96,009
Yard improvements (10% of PE)	96,009
Total Direct Costs	**$1,948,976**
Indirect Costs:	
Engineering and supervision (32% of PE)	307,228
Construction Expenses (34% of PE)	326,429
Total Indirect Costs	**$ 633,657**
Total Direct & Indirect Costs	**$2,582,633**
Contractors (5% of Direct & Indirect)	129,132
Contingency (10% of Direct & Indirect)	258,263
Fixed Capital Investment	2,970,028
Working Capital (18% of Fixed Capital Invest.)	534,605
TOTAL CAPITAL INVESTMENT	**$3,504,633**

TABLE 4-2:

ECONOMIC ANALYSIS OF
PAFBC SYSTEM vs. OIL-FIRED BOILER SYSTEM

ECONOMIC ANALYSIS OF PAFBC SYSTEM VERSUS OIL FIRED BOILER SYSTEM

Assumptions - PAFBC system:

Plant life (yrs)	15
Operating days/yr	330
Coal feed rate (lb/hr)	5376
Coal cost ($/ton)	$30
Coal cost escalation	3%
Limestone feed rate (lb/hr)	1906
Limestone cost ($/ton)	$15
Limestone cost escalation	2%
Ash generation rate (lbs/hr)	2166
Ash disposal cost ($/ton)	$8
Ash disposal escalation	5%
Maintenance cost (%)	2%
Maintenance escalation	3%

Heating value BTU/lb)	12500
Heat input (MMBTU/hr)	67.2
Water usage (gal/hr)	8640
Water cost ($/kgal)	$0.50
Water cost escalation	3%
Electricity usage (KWh/hr	375
Electricity cost ($/KWh)	$0.05
Electricity escalation	5%
Operating labor (#/shift)	1
Shifts/day	3
Labor cost ($/man yr)	$40,000
Labor escalation	4%

Assum. oil-fired boiler system:

Fuel cost ($/MMBTU	$4.00
Fuel escalation	5%

Capital cost analysis:

PAFBC System	$3,504,633
Oil-Fired Boiler	$405,169
Cost differential	$3,099,464

Return on investment analysis:

Payback (yrs)	2.73
ROI % (100 equity)	41%

Annual O&M costs -PAFBC System:

	year 1	year 2	year 3	year 4	year 5	year 6	year 7	year 8	year 9	year 10	year 11	year 12	year 13	year 14	year 15
Coal	$639	$658	$678	$698	$719	$740	$763	$785	$809	$833	$858	$884	$911	$938	$966
Limestone	$113	$115	$118	$120	$123	$125	$128	$130	$133	$135	$138	$141	$144	$146	$149
Ash disposal	$69	$72	$76	$79	$83	$88	$92	$97	$101	$106	$112	$117	$123	$129	$136
Water	$34	$35	$36	$37	$39	$40	$41	$42	$43	$45	$46	$47	$49	$50	$52
Electricity	$149	$156	$164	$172	$181	$190	$199	$209	$219	$230	$242	$254	$267	$280	$294
Operating labor	$120	$125	$130	$135	$140	$146	$152	$158	$164	$171	$178	$185	$192	$200	$208
Maintenance	$70	$72	$74	$77	$79	$81	$84	$86	$89	$91	$94	$97	$100	$103	$106
Total O&M cost	$1,193	$1,234	$1,275	$1,318	$1,363	$1,409	$1,457	$1,507	$1,559	$1,612	$1,668	$1,725	$1,785	$1,847	$1,911

Annual O&M cost - Oil-Fired Boiler:

	year 1	year 2	year 3	year 4	year 5	year 6	year 7	year 8	year 9	year 10	year 11	year 12	year 13	year 14	year 15
Fuel cost	$2,129	$2,235	$2,347	$2,464	$2,588	$2,717	$2,853	$2,996	$3,145	$3,303	$3,468	$3,641	$3,823	$4,014	$4,215
Operating labor	$40	$42	$43	$45	$47	$49	$51	$53	$55	$57	$59	$62	$64	$67	$69
Water	$38	$39	$40	$42	$43	$44	$45	$47	$48	$50	$51	$53	$54	$56	$57
Electricity	$50	$53	$55	$58	$61	$64	$67	$70	$74	$78	$81	$86	$90	$94	$99
Maintenance	$8	$8	$8	$9	$9	$9	$10	$10	$10	$10	$11	$11	$11	$12	$12
Total O&M cost	$2,265	$2,377	$2,494	$2,618	$2,747	$2,883	$3,025	$3,175	$3,332	$3,497	$3,670	$3,852	$4,043	$4,243	$4,453
PAFBC O&M savings	$1,072	$1,143	$1,219	$1,299	$1,384	$1,473	$1,568	$1,668	$1,773	$1,885	$2,002	$2,127	$2,258	$2,396	$2,542

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APPENDIX 1:

PAERC Equipment Specifications

APPENDIX 1: PAFBC EQUIPMENT SPECIFICATIONS

PAFBC SYSTEM EQUIPMENT LIST

SECTION 100--COAL AND LIMESTONE HANDLING AND FEEDING

EQUIP#	#	ITEM DESCRIPTION	CAPACITY (EACH)	DIMENSIONS (EACH)
V 1	1	TRUCK DUMP HOPPER		
C 1	1	HOPPER TO CLASSIFIER CONVEYOR	15 TONS/HR @ 50 #/CF (1400 CFH)	
C 2	1	AIR CLASSIFIER	15 TONS PER HOUR COAL, 2800 CFM AIR	20" OD PIPE, 25' HIGH
C 3	1	CLASSIFIER-TO-BUCKET ELEVATOR CONVEYOR	15 TONS PER HOUR COAL MAX.	
F 1	1	CLASSIFIER AIR FAN	2800 CFM, 20" WG	
E 1	1	CLASSIFIER AIR HEATER	13 CFM NAT. GAS, 320 F AIR TEMP RISE	
C 4	1	BUCKET ELEVATOR	15 TONS PER HOUR	
C 5	1	ELEVATOR-TO-SILO CONVEYOR	15 TONS PER HOUR MAX	
S 1	1	COARSE COAL SILO VENT FILTER		
V 2	1	COARSE COAL SILO	1730 CUBIC FT	12' DIAM, 25' HIGH, 60 DEGREE HOPPER

M 1	1	COARSE COAL METERING FEEDER	4500 LB/HR, 50 PCF	
S 2	1	CLASSIFIER CYCLONE SEPARATOR	2800 CFM, 90 FPS INLET VELOCITY	
C 6	1	CYCLONE SEPARATOR AIRLOCK FEEDER		
S 3	1	COAL FINES SILO VENT FILTER		
V 3	1	COAL FINES HOPPER	600 CUBIC FT	7.5' X 9' X 9' HOPPER
S 4	1	LIMESTONE SILO VENT FILTER		
V 4	1	LIMESTONE SILO	460 CUBIC FT	8' DIAM, 14' HIGH, 60 DEGREE HOPPER
M 2	1	COAL FINES METERING FEEDER	1500 LB/HR, 45 PCF	
M 3	1	LIMESTONE METERING FEEDER	2000 LB/HR, 70 PCF	
C 7	1	FINES SPLITTER	1-3" PIPE INLET, 6-1.25" PIPE OUTLETS	
F 2	1	FINES CONVEYING AIR FAN	170 CFM, 130" WG	
F 6	1	COAL/LIMESTONE CONVEYING AIR BLOWER	670 CFM, 4 PSI	

SECTION 200--COMBUSTION AIR FEED

EQUIP#	#	ITEM DESCRIPTION	CAPACITY (EACH)	DIMENSIONS (EACH)
F 3	1	PULSE COMBUSTOR BLOWER	15,000 #/HR AIR @ 40" WG	
F 4	1	DISTRIBUTOR AIR BLOWER A	22,500 #/HR AIR @ 75" WG	
F 5	1	DISTRIBUTOR AIR BLOWER B	22,500 #/HR AIR @ 75" WG	

SECTION 300--COMBUSTOR & HIGH TEMP PARTICULATE REMOVAL

EQUIP#	#	ITEM DESCRIPTION	CAPACITY (EACH)	DIMENSIONS (EACH)

H 2	1	DISTRIBUTOR SECTION	43,000 lb/hr, 1.5 psig, 392 bubble caps	
	1	BED SECTION	10'x10' to 12'x12', 50", 18 in-bed stm-raising modules	
	1	FREEBOARD SECTION	12'x12', 10' high	
H 1	1	PULSE COMBUSTOR	9.5' high, 34" OD jacket, 6-6" cs sch 40 tailpipes	
S 5	1	REFRACTORY-LINED HOT CYCLONE	55000 ACFM @ 1550 F, 90 FPS inlet vel.	

SECTION 400--CONVECTIVE PASS EVAPORATOR AND ECONOMIZER

EQUIP#	#	ITEM DESCRIPTION	CAPACITY (EACH)	DIMENSIONS (EACH)
E 4	1	BOILER SECTION	18,722 lb/hr 175 psig sat. stm	
E 5	1	ECON. SECTION	54,000 lb/hr fdwtr, 170 F in, 275 F out	
P 2,3	2	BOILER CIRC. WTR PUMPS	1,025 gpm each, 175 psig inlet, 23 feet TDH	

SECTION 500--LOW TEMP PARTICULATE REMOVAL & STACK

EQUIP#	#	ITEM DESCRIPTION	CAPACITY (EACH)	DIMENSIONS (EACH)
S 6	1	BAGHOUSE	25000 ACFM, 5 GR/ACF IN, 0.02 GR/ACF OUT	
F 7	1	ID FAN	25000 ACFM, 14" WG PRESSURE RISE	
T 1	1	STACK		

SECTION 600--ASH HANDLING AND DISPOSAL

EQUIP#	#	ITEM DESCRIPTION	CAPACITY (EACH)	DIMENSIONS (EACH)
E 2,3	2	ASH COOLERS	1400#/hr @ 62#/cf ea, 1550F in, 400F out	
P 1	1	ASH COOLER WTR PUMP	10,000 lb/hr (20 gpm), 50 psig	
C 9,A	2	BED DRAIN DUMP FEEDERS	1400#/hr ash @ 62 #/cf ea.	
C B,C	2	BAGHOUSE ASH DUMP FEEDERS	1400 lb/hr @ 40 #/cf	
S 7	1	ASH SILO 1ST STAGE SYCLONE		
S 8	1	ASH SILO 2ND STAGE CYCLONE		
C D	1	ASH VACUUM EJECTOR	2000 lb/hr stm @ 40 psig	
V 6	1	ASH DISPOSAL BIN	3000 CUBIC FT	16' DIAM, 25' HIGH, 55 DEGREE HOPPER
S 9	1	ASH SYSTEM FILTER RECEIVER		
S A	1	ASH BIN VENT FILTER		
V 7	1	ASH CONDITIONER	500 cfh @ 62 #/cf	

APPENDIX 2:

Plant Start-Up and Operation Procedures

Introduction

In order to properly discuss the Start-Up of the PAFBC unit, the application of Safety Interlocks must be discussed. The majority of the safety interlocks are hardwired in the MCC, there are however, some software interlocks controlled by the PLC. The MCC has two sets of interlocks. The first set of interlocks controls the motors governing the Silo loading processes, while the second set of interlocks governs the blowers related to the operation of the fluid bed and the pulse combustor. The software interlocks are mainly for the flame safety management system. However, under certain emergencies the PLC will trip the FD Fan shutting off all the blowers except the ID fan, and 10 seconds later the PLC will trip the ID fan. These conditions are low boiler level and low water flow rate to the tube modules or the bed. The PLC will shut the fuel off certain sub systems without shutting the plant down if it is safe to continue to operate the system without that particular sub system.

Whenever the pulse combustor is on it is crucial to maintain at least a minimum flow rate from the Forced Draft Fan. The reason for this is two fold. This first reason is that the air is needed to keep the distributor plate cool if there is no bed. Secondly, if there is not a minimum velocity through the bubble caps, then bed will backsift into the windbox. The pulse combustor should never be pulsing while there is no bed in the reactor either. The reason for this is the vibrational force imposed by the undampened pulse combustor has been known to back out bolts on the underside of the combustor, loosening flanges and valves.

Solids Loading

Introduction

Loading the coal into the silos can sometimes be slightly tricky. On a warm dry day, loading the coal into the silos should be relatively easy. Normally, when the coal leaves the bottom of the pickup hopper, then the coal loading process to the silos is normally relatively easy. In cases when the coal is wet and is sticking however, the pick-up hopper has two vibrators and some air jets in order to get the coal to move onto the hopper to coal mill screw conveyor. In some cases, if the coal is relatively wet, coal fines will build up and eventually block the coal chute into the coal mill. The door in the coal chute should be opened periodically in order to remove metal fragments on the magnet, so it is always a good idea to check to see if a cake is developing that is blocking the coal mill inlet chute. Never open any of the doors to the coal mill while it is in operation. Also the coal mill will not start if there is a load on the rollers so should a cake develop it must be removed and not pushed in such a manner that the rollers will have to roll over it as the mill spools up.

Loading Coarse Coal and Fine Coal simultaneously

When loading coarse coal into the coarse coal silo and fine coal into the fine coal silo, follow this procedure.

Make sure that the Coarse Coal Silo Bin Vent Filter (S-1) is pulsing by establishing air flow to the filter, and energizing breaker #.
Make sure that the Fine Coal Silo Bin Vent Filter (S-3) is pulsing by establishing air flow to the filter.
Make sure the fine coal bypass switch is in the off position. If the fine coal bypass is engaged, disengage it.
Make sure inlet damper to Classifier I.D. Fan (F-1) is closed. If the damper is open, close it.
Turn on Classifier I.D. Fan (F-1).
Turn on Rotary Air Lock Feeder under Fine Coal Baghouse.
Turn on Rotary Air Lock Feeder under Fine Coal Cyclone.
Open inlet damper to Classifier I.D. Fan (F-1) to the desired position.
Turn on Coarse Coal Weigh Belt Feeder (C-5)
Turn on Bucket Elevator (C-4)
Turn on Classifier to Bucket Elevator Screw Conveyor (C-3)
Turn on Coal Mill to Classifier Screw Conveyor
Turn on Coal Mill (C-F)
Turn on Hopper to Coal Mill Screw Conveyor (C-1)
Begin Loading Coal Hopper.

Loading Coarse Coal only

During times when the fine coal silo might be full, or when there is no need to feed fines into the pulse combustor, than all of the coal can be loaded into the coarse coal hopper. The following procedure works well. However, if the coal is relatively dry it may be necessary to follow the previous procedure and keep the damper for the Classifier I.D. Fan closed all of the way. The little bit of suction imposed by the Classifier ID Fan when the damper is completely closed is enough to keep coal dust from leaking out of holes in the screw conveyors. Otherwise, when choosing to load only coarse coal into the coarse coal silo and to not load fine coal into the fine coal silo, follow this procedure.

Make sure that the Coarse Coal Silo Bin Vent Filter (S-1) is pulsing by establishing air flow to the filter, and energizing breaker #.
Make sure the fine coal bypass switch is in the on position. If the fine coal bypass is disengaged, engage it.
Turn on Coarse Coal Weigh Belt Feeder (C-5)
Turn on Bucket Elevator (C-4)
Turn on Classifier to Bucket Elevator Screw Conveyor (C-3)
Turn on Coal Mill to Classifier Screw Conveyor
Turn on Coal Mill (C-F)
Turn on Hopper to Coal Mill Screw Conveyor (C-1)
Begin Loading Coal Hopper.

Limestone Loading procedure

The limestone is brought in to the plant via a pneumatically equipped delivery truck. Limestone is blown directly form the truck through a pipe and into the Limestone Silo. To prepare the Limestone Silo for loading, follow these steps:

Make sure that the Limestone Silo Bin Vent Filter (S-?) is pulsing by establishing air flow to the filter, and energizing breaker #.

Make sure that the Limestone Silo Bin Vent Filter (S-?) is pulsing by establishing air flow to the filter.

Allow driver of pneumatic limestone truck to begin emptying his load into the silo.

Initial Start-up (no bed)

Filling the Boiler Drum

It is imperative that the boiler level reach the low level before the circulating pumps are turned on. Also, all valves in the boiler water circulation line must be open before a recirculation pump is turned on. Both pumps should have their valves open, so that if one of the pumps should fail, the next one can be quickly started. Minimizing the time that the tubes are in a hot bed without water circulating through them is of the utmost priority. Following this next sequence will ensure the safest possible method for filling the boiler and then starting boiler circulation. Do not under any circumstances attempt to operate a Boiler Recirculation Pump Dry!

Open all boiler recirculation valves.

Open all manual feedwater valves on the PAFBC site.

Make sure that the economizer by-pass valve is closed.

On the Economizer screen open the Boiler Feedwater Control Valve.

Open the Boiler Feedwater Valve inside the Clemson Power House.

Make sure that water is flowing into the in-bed tube modules as well as the Pulse Combustor Water Jacket. This must be checked before the apparatus fill up, since there are no accessible drains on the Pulse Combustor Water Jacket.

Continue filling the boiler until the normal level has been reached.

Turn on one of the boiler recirculation pumps.

Fan/Blower Startup sequence

The fan and blower motors must be started in this order. As a safety precaution, this is the only order in which the fan and blower motors will start in. Both the I.D. Fan (F-7) and the F.D. Fan (F-4), are equipped with soft start MCC panels. These panels allow for the slow startup of the large motors without blowing the fuses or the motors. The F.D. Fan motor will pull over 1,200 amps for a brief period of time, this is perfectly normal. Altering the Ramping Cycle on this motor will cause the heaters to trip if the motor is started on a cool day or night. Should the heater trip, it is recommended that the MCC panel is allowed to cool for 15 minutes before an attempt is made to restart the F.D. Fan (F-4). If the heaters in the MCC panel are not allowed to cool, starting the fan will be impossible.

Close Damper to I.D. Fan (F-7).
Close Damper to F.D. Fan (F-4).
Turn on I.D. Fan (F-7).
When I.D. Fan has completely spooled up start F.D Fan (F-4).
When F.D Fan (F-4) has completely spooled up start.
Coal/Limestone Blower ().
Start Fine Coal Blower (F-2)
Start Pulse Combustor Pilot Blower ()
Start Pulse Combustor Main Air Blower (F-3)

Bed Loading Procedure

In order to dampen the acoustic reverberations of the pulse combustor, it is necessary to load a couple inches of bed before going to main gas on the pulse combustor.

Turn on Coal/Limestone blower.
Open up Hand Valves in Coal/Limestone line below combustor.
Open up Hand Valves above each eductor on the Coal/Limestone Mixing Hopper.
Open Forced Draft Fan Inlet Damper to achieve a bed velocity of .75 ft/s.
Open Hand Valve above Limestone Metering Feeder.
Turn Limestone Metering Feeder on 75 percent.

Pulse Combustor Start-up

The pulse combustor chamber must be heated up slowly overtime to prevent the cracking of the refractory. Ideally, the refractory should be brought to a temperature of 350 F over several hours, then held for one hour. After the water has been driven from the refractory in this manner, the pulse combustor can be taken up 1 MMBTU/hr, wait until it is stable, then increase the firing rate again. The Air Flow rate should be adjusted in order to keep the air plenum temperature below 350 F.

When starting up the pulse combustor pilot, and determining air to fuel ratios for the pilot, follow this chart closely. <u>Over firing the pilot can cause flame impingement!</u>

Make sure all blowers are on and that the Forced Draft Fan is delivering a Bed Velocity of roughly 0.75 ft/s.
Open all manual block valves on natural gas line to the Pulse Combustor or pilot.
Slowly open main gas valve.
Turn on Power to the Pulse Combustor Flame Safety Panel.
Turn the ignition switch.
After purging sequences light pilot with a gas flow rate of ? and an air flow rate of ?.

Slowly bring the pilot up to full firing rate increasing the firing rate .5 MMBTU/hr every time the temperature stabilizes.
After the Pilot has reached the maximum setting on the chart light main gas at 1 MMBTU/hr, and slowly increase until pulsations occur.
After you have reached full firing rate and the bed temperature is as high as it will go, then light the in bed gas injectors.

Ignition of imbedded gas injectors

The fluid bed combustor has gas injectors installed near the bubble caps. This gas is used to start-up the bed, and can be used to produce steam at roughly 30,000 lbs/hr with the pulse combustor running at its full firing rate, should something prevent coal from entering the reactor. By the time the pulse combustor has reached 18 MMBTU/hr the refractory in the bubbling fluid bed should be dry. Therefore, when igniting the in bed gas injectors it is only necessary to light the pilot and then bring the main gas online. When igniting the imbed gas injectors use this procedure:

Like turn them on and stuff

Starting Bed on Coal

The bed must be up to 1000 F before coal is injected into the bed. Once coal is ignited the bed temperature will jump drastically. So in the beginning the coal must be increased slowly, as the reaction to a slight increase can be delayed.

Make sure the hopper valves are open.
Make sure that the feed line valves to the bed are open.
Open hand valve above Coal Metering Feeder.
Manually increase the Coal Feed to a small flow rate.

Starting Pulse Combustor on Fine Coal

In order to fire fine coal into the pulse combustor, the bed temperature must be 1500 degrees. Cold, unburned coal fines in the freeboard represent a possible explosion hazard.

Open the valve to the coal metering feeder.
Turn on the fine coal metering feeder.

Switching Steam to Clemson Main Header

As the pulse combustor heats up it will start building pressure up in the boiler. As the pressure builds up it is important that some of the steam is vented off in order to deairate the steam supply. As the pressure starts to build open up the 10 inch main steam valve on the PAFBC site. This will allow the steam to get to the vacuum eductor. As the pressure builds up, the supply valve to the eductor should be opened up every once an awhile for a couple minutes until the boiler is close to the correct pressure. This will also warm up the steam supply line, and to equalize the pressure across the 10-inch steam supply valve located in the Clemson powerhouse. When the pressure is close to 125 psi the 10 inch steam valve in the Clemson Powerhouse must be opened.

Cold Restart (Bed below 1000)

Introduction

If the plant has been down for a lengthy period of time and the bed has cooled down then this procedure should be implemented. The steps are fairly similar to starting the boiler without a bed in place. However, there are a few new considerations that need to be looked after concerning the bed loading system in particular. Other subsytems have slight modifications or omissions.

Boiler Feed Water

In many instances the boiler still maintains a normal level. In fact, if the bed is still over 300 F then the boiler should still be recirculating in order to keep the in bed tubes from being damaged. If the bed is really cold and the boiler has been drained follow the boiler filling procedure as listed in the Initial Startup Procedure. Otherwise follow this procedure:

Make sure that the manual feed water lines are open to except water from the main boiler house.
Open the boiler feed water to the plant inside the main boiler house.
Make sure all circulation valves are open .
Turn on one of the boiler recirculation pumps.

Fan/Blower Start-Up Sequence

Because of the interlocks, the fans must be started in the same sequence as they were previously. However, certain precautions must be taken in order to insure that the bed feed lines do not become plugged. It is important in this case to make sure that the hand valves below the bed for the coal/limestone feed lines have been closed. These should have been closed when the bed was shutdown. If this was not the case close the valves and check to see if the lines are plugged. If the line is plugged to the elbow then remove the flange on the end and clear the plugged lines before restarting. The other concerns regarding the two 200 hp motors are still valid. When starting up the motors on a cold restart, follow this procedure:

Make sure that the hand valves below the fluid bed on the coal/Limestone Feed Lines are closed.
Check Feed lines for pluggage by tapping on the pipes.
Unclog any plugged feed lines.
Close Damper to I.D. Fan (F-7).
Close Damper to F.D. Fan (F-4).
Turn on I.D. Fan (F-7).

When I.D. Fan has completely spooled up start F.D Fan (F-4).
When F.D Fan (F-4) has completely spooled up start.
Coal/Limestone Blower ().
Start Bed Loading Procedure.
Start Fine Coal Blower (F-2)
Start Pulse Combustor Pilot Blower ()
Start Pulse Combustor Main Air Blower (F-3)

Bed Loading Procedure

Although the bed is loaded and will dampen the acoustic reverberations it is important to still take care of the bed loading system first. Since the hand valves were closed the roots blower has been under stress due to its in ability to deliver a large enough flow rate to its outlet. Therefore, it is imperative to get the air running into the bed as soon as possible. Follow this sequence for opening the valves up on the pneumatic transfer lines.

Open the butterfly valve allowing air to the eductor on one of the lines.
Open the hand valve below the eductor allowing the air to flow into the bed.
Read the orifice plate meter to see that air is flowing through the eductor. (The differential pressure should be roughly 4 inches of water column)
To insure that the air is flowing into the bed check to make sure the eductor is pulling a vacuum.
If line is plugged close off the air to the eductor and blow the line clean using the compressed air, then reopen the butterfly valve to supply the eductor.
When air is flowing into the bed through a feed line proceded to the next line and repeat procedure till all lines are open.

Pulse Combustor Start-Up

In the case of a cold start, as long as the pulse combustor hasn't been exposed to moisture for any length of time, it is all right to bring the combustion chamber up slowly at roughly 100 F an hour till a temperature of 400 F is reached. After that bringing up the pulse combustor 1 MMBTU/hr, holding until the combustion chamber temperatures stabilize and then increase the firing rate until the maximum firing rate of 18 MMBTU/hr has been achieved. When on manual air controls the air flow should be adjusted to keep the air plenum temperature below 350 F. When starting up the pulse combustor pilot, and determining air to fuel ratios, follow this chart closely. <u>Over firing the pilot will cause flame impingement!</u>

Make sure all blowers are on and that the Forced Draft Fan is delivering a Bed Velocity of roughly 0.75 ft/s.
Open all manual block valves on natural gas line to the Pulse Combustor or pilot.
Slowly open main gas valve.
Turn on Power to the Pulse Combustor Flame Safety Panel.
Turn the ignition switch.

After purging sequences light pilot with a gas flow rate of ? and an air flow rate of ?.
Slowly bring the pilot up to full firing rate increasing the firing rate .5 MMBTU/hr every time the temperature stabilizes.
After the Pilot has reached the maximum setting on the chart light main gas at 1 MMBTU/hr, and slowly increase till pulsations occur.

Ignition of imbed gas injectors

After the pulse combustor has reached roughly 10 MMBTU/hr, and the bed still is not up to 1000 F then lighting the gas injectors in the bed to raise the bed temperature can be initiated.

Starting Bed on Coal

The bed must be up to 1000 F before coal is injected into the bed. Once coal is ignited the bed temperature will jump drastically. So in the beginning the coal must be increased slowly, as the reaction to a slight increase can be delayed.

Make sure the hopper valves are open.
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