# Analysis of Improvements in the FETC Fluid Bed Gasifier

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**Abstract**

Over the past four years, FETC has improved reliability, availability, and performance of the FETC 10 Inch Fluid Bed Gasifier (FBG) operating at 425 psig. To improve the gasifier performance and avoid ash clinker formation, it was necessary to evaluate and improve solid mixing within the fluid bed. It was shown that the jet momentum was the key parameter used to control mixing and that increasing jet momentum resulted in marked improvements in operational reliability. For the first time, FBG tests have demonstrated that increased momentum in the center region of the feed jet, with the coal, markedly improved mixing. Observation of improved mixing was measured indirectly by a combination of decreased peak temperatures, increased average bed temperatures, and/or additional gas-make, and carbon conversion. Tests also confirmed past research that it is necessary to promote movement under the jet inlet in order to establish consistent stable performance. This was accomplished using both underflow fluidizing gas, jets in the bottom cone, and by removing larger more dense solids which accumulate in the bottom of the bed periodically.

**INTRODUCTION**

Gasification and hot gas cleanup are essential components to many advanced power generation technologies. Integrated Gasification Combined Cycle (IGCC) and Advanced Pressurized Fluid Bed Combustion (APFB) processes cleanly and efficiently convert coal into electricity. Only through high component reliability can these technologies become feasible to replace the current installed base of power generation. Kellogg and Foster Wheeler (McClung et al., 1995) are active technology developers of these new power cycles at the Power System Development Facility in Wilsonville, Alabama (Powell et al., 1996) .

The FETC 10-inch Fluidized-Bed Gasifier (FBG) system is a jetting fluid bed reactor similar in design to that of the KRW gasifier (Haldipur et al.,1988; Katta et al., 1985) and the Foster Wheeler carbonizer (Froehlich et al., 1994). This type of fluid bed system has proven ideal for coal conversion processes under reducing conditions because of the capability to handle caking coals. In such systems the coal rapidly mixes with solidified char before it can devolatilize and melt to form carbon agglomerates. Due to mixing the bed temperatures are uniform and stable.

The FBG has been used to gain operational experience for the gasifier and subsystems which operate at elevated pressures. The facility has been run to develop and design components of a novel gasifier, PyGas$^{TM}$ (Brown and Sadowski, 1991); to develop FBG operational control strategy, jet design and operational characteristics; and to prepare bituminous coal char at elevated pressure for further characterization. In addition, this gasifier provides operational data to verify predictive process models, and it was used to test and verify advanced control schemes prior to their application.

The primary, process objective was to provide a high quality, realistic, coal gas by a reliable method to permit downstream testing of hot gas cleanup components and processes. One aspect of the FBG operation of particular interest is the mixing in the jet since this is critical to decaking bituminous coals. Past experience with caking coals had been unsuccessful, and operations with subbituminous coals were often plagued with the production of clinkers. In order to improve reliability and operate with caking coals, it was deemed necessary to improve bed mixing.

From 1993 until 1996, many mechanical, operational, and procedural modifications were made to achieve reliable operations. More specifically, the start-up procedures were modified, the momentum distribution of reactants in the central jet was redistributed, the fluidization gas below the central jet was maintained, cone jets to promote mixing were added, and the solids withdrawal system was improved. In the final year of these performance improvements, a high speed differential probe provided some insight into the coal bed's dynamics.

## EXPERIMENTAL

The FBG consists of a pressurized (425 psig) gasifier and several subsystems, e.g., coal storage conveying system, air preheater, particulate removal system, and an incinerator. The gasifier is a 20-ft high, nominal 80 lb/hr, air-blown and refractory-lined vessel (Figure 1). Coal is fed into the gasifier by ambient (convey) air through the center core of a concentric jet located at the bottom center of the bed. Preheated (reactor) air and steam are premixed and introduced into the bed through the annular core of the jet. This system has produced about 300 lb/hr of flue gas with a heating value of 120 Btu/scf. Typical process conditions are presented in Table 1.

Coal ground to 14 X 60 mesh (1.41 mm to 0.250 mm) is pneumatically conveyed into the bottom of the 3 stage, refractory-lined gasifier along with steam, preheated air and a small amount of nitrogen. Solids from the gasification process are continuously withdrawn from the top and intermittently withdrawn from the bottom of the fluid bed. The product gases are processed

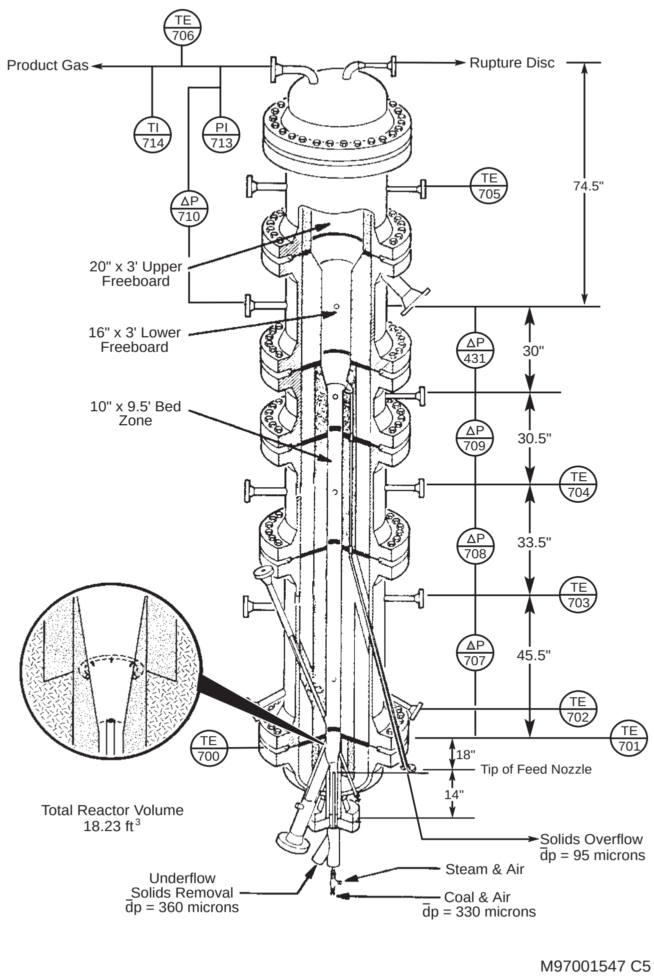


**Figure 1**. Schematic of the FBG and key instrumentation and feed locations.

through two cyclone separators and heat traced to the inlet of a candle filter vessel in an adjacent structure.

The fluid bed gasifier was operated at various reactant ratios in an attempt to improve the operation of the gasifier and to investigate the effects of small deviations in reactor inputs such as air, steam, coal, and diluent nitrogen on measured parameters.  Physical locations of reactant streams entering the gasifier were also varied throughout the testing to improve gas and solid mixing in the fluid bed.  Improvements in operation were viewed from the standpoint of attaining "clinker free operation" and  improving the gas heating value and carbon conversion.  Detailed analysis of 11 test runs identified more than 35 steady state periods.  Observations, process measurements, and calculations used to assess the gasifier operation and the effectiveness of the changes were a)clinker formation, b) carbon conversion, c) gas heating value, d) bed temperature, and e) bed density.

Table 1.  Typical Process Conditions for Fluid Bed Gasifier.

| Description of Process Variables | Range | |
|---|---|---|
| Coal Feed Rate | 70-80 lb/h | 32-36 kg/h |
| Particle Size Range | | <500 microns |
| Total Air Flow Rate | 150-200 lb/h | 68-91 kg/h |
| Steam Flow Rate | 50-57 lb/h | 23-26 kg/h |
| Gas Mass Throughput | 300-350 lb/h | 136-159 kg/h |
| Bed Velocity | 0.3-1.0 ft/s | 0.09-0.3 m/s |
| Average Bed Temperature | 1,550°F | 843°C |
| Reactor Pressure | 425 psig | 3,032 kPa |

The following adjustments were tested in 1994 and 1995 and refined in 1996 to evaluate the effect on mixing in the jet: 1) the addition of steam and nitrogen into four new jet locations in the conical section of the gasifier, 2) the addition of underflow nitrogen to facilitate char withdrawal, 3) the redistribution of air and steam in the convey and process inlets in the nozzle, 4) the readjustment of total coal and gas flow rate, 5) the retargeted air/coal ratio, and 6) the retargeted ratio of steam/air. The latter adjustments involved parametric studies, ie. variations in the reactant ratios, and is not the focus of this paper.

As a first step to evaluating gasifier process improvements, the gasifier was instrumented in 1994 to gain a better understanding of mixing between the char and injected coal above the inlet of the fluid bed. Temperature profiles near the jet were introduced to infer the degree that hot combusting coal and bed char are mixing. Three thermocouples inserted in stainless steel sheaths penetrated 3" and 2" from the wall towards the center of the 10" diameter vessel at an elevation of two to three feet above the nozzle (Figure 1). The temperatures in the bed were monitored at the wall, at the top of the conical section (18" from the nozzle tip), and every few feet along the length of the gasifier. In addition, differential pressure measurements were made along the length of the fluid bed (Figure 1).

## RESULTS AND DISCUSSION

### GASIFIER PERFORMANCE

The FBG's performance can be measured by several factors: carbon conversion (%wt based on carbon in coal), high heating value of the total product gas (HHV, Btu/scf), temperatures in the bed and in the combustion zone (°F), gas make (scfh), and its availability. Availability was defined as the percent of time spent gasifying coal based on the total time possible excluding that time required for start-up and shutdown. Increasing these factors, except perhaps the temperatures, would increase the FBG's performance. Operating at higher temperature, while not strictly improving performance, would broaden the operating range achievable. To obtain a high availability by avoiding the clinker formation, the operating temperature, either in the bed or in the bottom combustion zone, was targeted to not exceed 1,700°F (927°C). The changes in FBG's performance from 1993 to 1996 is displayed in Figures 2 and 3.

The most consistent improvement was the increase achieved in carbon conversion from 74 to 90% wt. since 1993 (Figure 2A). On the other hand, the HHV only increased from 110 to 129 Btu/scf (4.1 to 4.81 MJ/$m^3$ std) from 1993 to 1995. In 1996, the HHV has dropped back to 122 Btu/scf because the combustion zone was operated at a much higher temperature (approx. 1,660ºF, 905ºC) than in the previous years. These heating value figures were not corrected for dilution nitrogen and varying steam feed rates; this can account for much of the variability found over these tests. In addition, parametric tests were conducted varying the air/coal ratios contributed to this variability in product gas heating value. The gas make, i.e. the amount of gas produced, was also increased from a little less than 4,000 scfh (113 $m^3$/hr std) in 1993 to 5,400 scfh (153 $m^3$/hr std) in 1996 as shown in Figure 2B. This was caused by both the increase of coal feed rate from 70 to 76 lb/hr (31.8 to 34.6 kg/hr) and the increased carbon conversion.

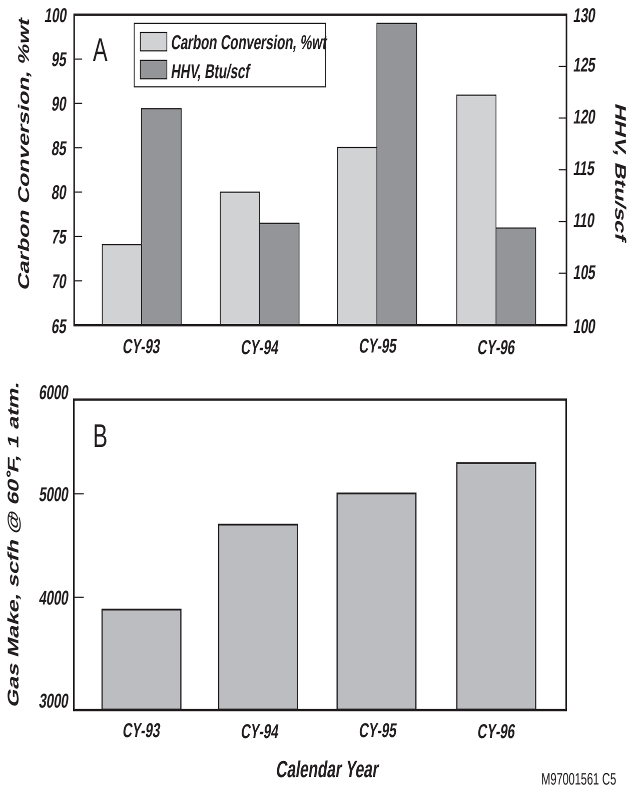


**Figure 2**. Performance measures of the FBG including A) the average carbon conversion and gas quality as measured by the higher heating value, and B) the average gas-make.

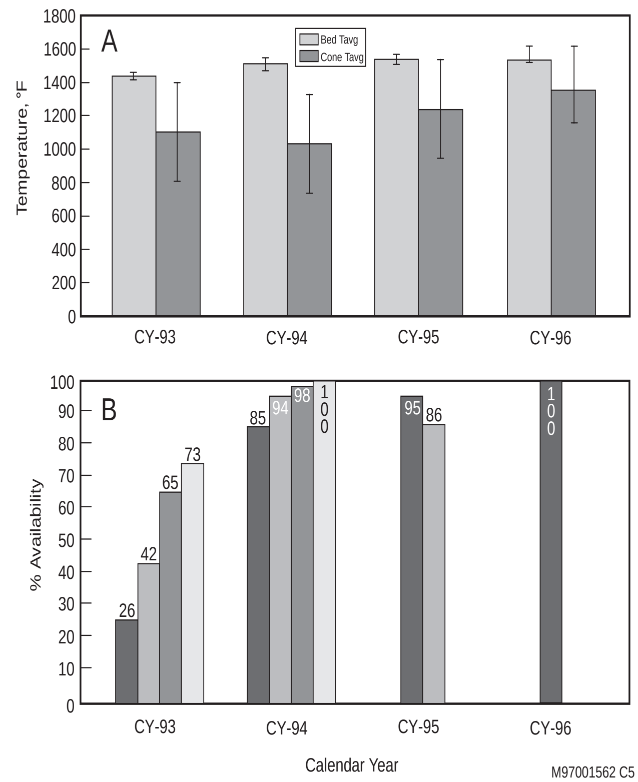


**Figure 3**. Performance measures of the FBG including A) the average and range of temperatures observed in the cone and upper bed, and B) the availability of fuel gas.

During these 4 years, we were able to increase the bed temperature from 1,400 to 1,660°F (760 to 905°C) as shown in Figure 3A. There have been even a greater increase in the combustion zone temperature from 1,100 to 1,300°F (593 to 705°C). These increases were primarily due improvements in mixing within the bed which also permitted increases in the air/coal and decreases in the steam/coal ratios.

The ultimate measure of performance of the FBG is its availability. It is defined as the number of gasification hours divided by the targeted testing time. With all the modifications described below our FBG availability has been increased from 26% in early 1993 to 100% in 1996 (Figure 3B). In 1995, however, the availability dropped down to 95% then to 86% due to the foreign material from the coal feed that plugged the feeder and the extensive parametric studies on the system, respectively. After instituting strict quality control precautions on the feed material, we regained the 100% availability in 1996.

GASIFIER PROCESS IMPROVEMENT

Prior to 1994, the FBG operations required very low temperatures (1500 F at reactor wall and 1200 F at gas outlet) to minimize ash clinker formation. Although the Montana Rosebud subbituminous coal tested is quite reactive, the reactor volume and char residence times were large and the gas composition was adequate (HHV=120 BTU/scf), but the conversion levels were rather low (70 - 75 %). To improve the gasifier performance in this regard and avoid ash clinker formation, it was decided that it was necessary to evaluate and improve solid mixing in the bed, under the jet inlet, and near the bottom of the cone. This was a prerequisite to operation of the gasifier on caking coal.

START-UP SEQUENCE. Prior to 1993 the fluid bed start-up procedure was to fill the bed with coke breeze and preheat the gasifier with hot nitrogen until the temperature was sufficient to permit bed ignition. This operation was plagued with slow start-up (15-24 hours) due to limited velocities with solids present, as well as high ignition temperature required for coke (800 F at 400 psig). In addition, the onset of ignition was achieved by introducing air into the hot bed causing a thermal excursion which was difficult to control. Often the start-up resulted in the formation of a clinker which would gradually grow large enough to shutdown the reactor.

A new startup sequence was developed in which the reactor was heated empty up using air to 600 F and sub-bituminous coal was injected, ignited, and used to begin building a bed. In this way, the more easily ignited fuel was fed in a controlled fashion to obtain the desired heat-up temperature (Figure 4). Heat up was reduced from 15 to 6 hours. The relatively low boiling characteristics of the volatiles from the Montana Rosebud coal, the 600 F preheat temperature, and the excess air assured that no volatile matter escaped the reactor. The coal feed rate was adjusted to control the bed temperature and achieve the desired vessel heat-up rate.

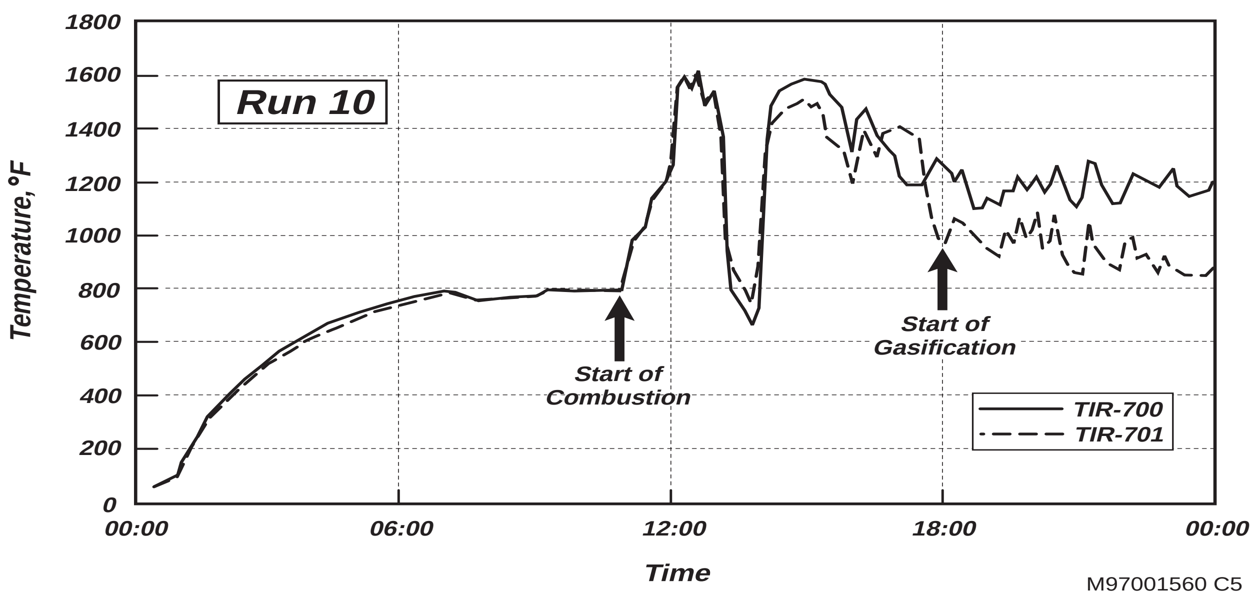


**Figure 4**. Log of temperature at the bottom of the reactor through a typical improved start up sequence from run 10 late in 1994.

Once sufficient bed material was added to the bed, taking about 2 hours, the process was converted from combustion to gasification. This was accomplished by purging the oxygen from the vessel using air and steam, and then introducing the air and coal at the desired ratio for gasification, nominally 3:1 lb/lb. Temperature excursions, and thus clinkers, were completely avoided with this procedure and a char bed could continue to be built using the desired nominal coal feed rate. In this way a consistent coal gas quality can be reached relatively quickly for downstream testing of gas clean-up processes.

EFFECT OF JET MOMENTUM. In 1994 the flow in the central coal convey tube of the nozzle was increased to maximize the momentum in the jet. This successfully improved mixing and jet penetration sufficiently to eliminate the formation of ash clinkers. Prior to these tests, the distribution of convey air was set at some minimum level sufficient to convey the coal and maintain 30 ft/sec velocity while the balance of the air and all the steam was introduced in the annular region of the nozzle (Figure 5). The primary design basis of this redistribution of reactants was to achieve mixing between center and annular inlets. However, little was known about the effect that this would have on the mixing between bed char and coal in the jet. After observing operation in this mode, it was proposed that the high velocity annular nozzle jet flow (170 ft/s, reactor air and steam) created a jet with great vertical penetration, though little radial mixing. A test was conducted during the last test run in 93' with a nozzle providing 30 ft/s flow in both the central and annular inlets in the nozzle. But for one or more reasons, the result of that run was the rapid formation of a clinker.

The effect of redistributing reactor air in the annular jet to convey air in the central jet was illustrated early in 1994. This was done to effect the hydrodynamics mixing in the cone/jet region of the gasifier. The convey (center jet) air was increased by almost 100% and the annular air (reactor air) was reduced by 50%, while the overall amount of air feed to the reactor remained unchanged. In doing this, a shift in the dominant jet was accomplished. Prior operations were conducted with the dominant jet being the outer jet, ie. with a momentum of about 4 times that of the center (coal/air) jet. After making the change, the center jet had a momentum two times that of the outer jet and the overall combined jet momentum increased by about 8% (Figure 5). This produced the desired result at the bottom of the gasifier - a reduction of the temperature at the bottom

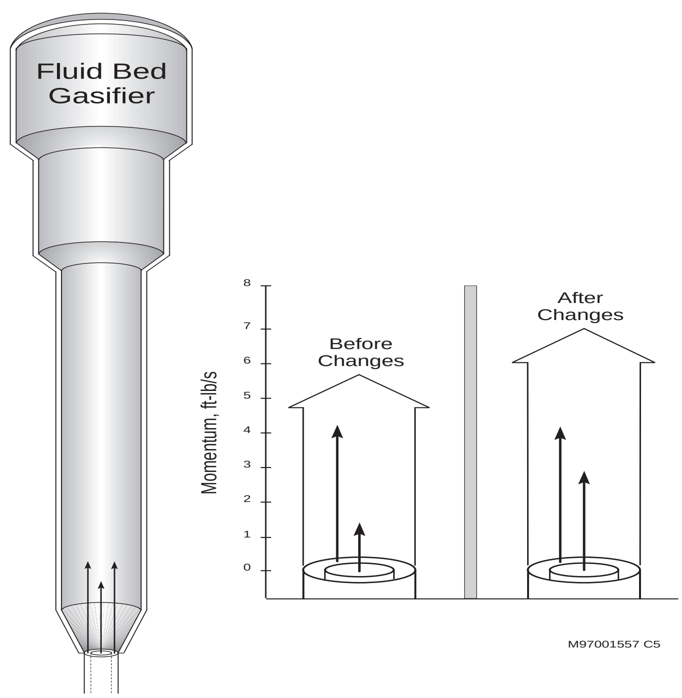


**Figure 5**. The effect of changing gas flows to the central nozzle on the distribution of momentum in the jet.

of the bed where clinker formation was problematic and reduced the temperature of 18" above the jet inlet (TIR-701) from 1405°F to 1029°F, while slightly increasing the mean bed temperature by 14 °F.

The decrease in temperature at the bottom of the gasifier is accredited to greater jet penetration (forcing the combustion zone higher in the bed), cold underflow nitrogen fluidizing gas below the jet, and better solid mixing in the bed. The result was less leaking of oxygen into the stagnant, downward, moving bed, and thus lower temperatures near the reactor wall. It is believed that this change in air/jet distribution and the addition of fluidizing nitrogen to the underflow region were the two biggest reasons that clinker formation was eliminated in 1994. As a point of reference, these conditions simulate a turndown of 10:1 for comparable industrial gasifiers such as the Foster Wheeler carbonizer and the KRW process.

EFFECT OF UNDERFLOW AND CONICAL JETS. The amount of underflow nitrogen is a primary parameter used to control mixing of solids below the central jet. Without taking underflow samples, it was found in 1993 that the bed would form a clinker in the cone when no underflow nitrogen was flowing. This gas is introduced below the outer annular region under the tip of the nozzle (Figure 6) in which a bed of solids forms. The flow rate was chosen to be at least equal to the minimum fluidization velocity of the bed material.

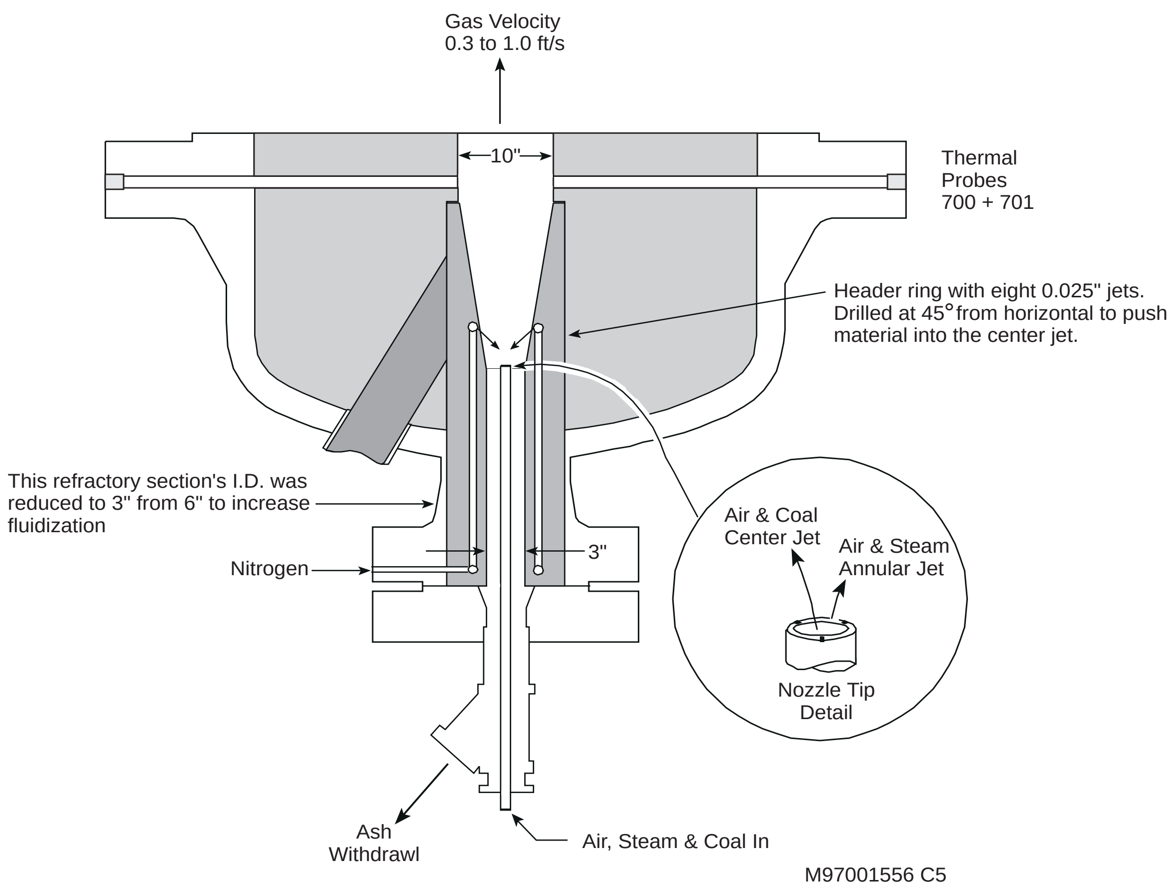


**Figure 6**. Configuration at the bottom of the FETC fluid bed gasifier.

In the early test in 1994, we found that clinkering can be avoided even without solids removal as long as fluidization gas is maintained below the nozzle inlet. In set point 07-2 a clinker was produced while no clinker was formed in 07-3. Both cases had no solids removal, but the successful test case (07-3) used 40 ft/sec convey velocity, while the earlier cases used a coal convey inlet velocity of only 20 ft/sec. Thus, as discussed above, we found that it was most critical to maintain the jet momentum with the coal convey velocity. However, in tests conducted later in 94', the solids removal below the jet was operational. In these tests solids were removed in lock-hopper batches, ie. in 5-10 lb batches every 30 minutes, and the underflow nitrogen was preheated, thereby increasing the velocity from about 1.4 times the minimum fluidization velocity ($U_{mf}$) to about 2.5 time the $U_{mf}$. In these runs both subbituminous and bituminous coals were used for nearly 300 continuous hours without producing any significant clinkers.

Based on these results, an understanding of the effect of changing the distribution of the reactants in the nozzles was obtained. The series of modifications made in the FBG over 1994 culminated in the successful operation on Illinois #6 bituminous coal over a test period lasting 7 hours. This first time success on caking coal was made possible due to several operational and mechanical improvements, the most notable of which include: 1) introducing heated fluidization gas in the char solids below the jet inlet, 2) increasing the coal inlet jet momentum, and 3) establishing a high temperature inert carbon-rich bed prior to introducing the caking coal. The gasifier ran at 425 psig while being fed 70 pounds per hour coal and reaching a peak bed temperature of 1735 °F. The gasifier was shutdown normally and post test inspections found no evidence for the formation of clinkers or agglomerates in the bed.

During the last test in 1994, a parametric test of the cone nitrogen flow showed that carbon conversion was significantly increased due to increasing hot char re-circulation into the air/coal jet. The cone jets were added around the slanted conical region of the bottom of the gasifier (Figure 6). These jets were configured with the jets blowing char down and back into the central jet. The carbon content in the overflow and underflow char was lower and the ash content in these streams higher after the adding nitrogen jets (Table 2).

Table 2. Effects of Side Jets on Gasifier Char

| Char Analysis | Side Jets installed (Oct. 1994) | | | |
|---|---|---|---|---|
| | Before | | After | |
| (%wt) | Underflow | Overflow | Underflow | Overflow |
| Fixed Carbon | 34.1 | 67.6 | 28.3 | 44.1 |
| Volatile Matter | 13.8 | 6.5 | 6.0 | 4.8 |
| Ash | 51.8 | 24.8 | 64.1 | 50.3 |

During these tests it was found that a reduction in dilution nitrogen flow from 140 to 50 scfh increased the fuel gas to non-combustibles ratio from 40 to 42 while also increasing carbon monoxide production by 16%. Though the fuel gas increase was expected, the increase in CO production may indicates that nitrogen interferes with the reactions that produce CO. The rest of the gas components did not fluctuate. This indicated that at underflow nitrogen rates, the jets still function as desired to fluidize the bottom of the bed. This results in a more stable operating fluid bed.

One interesting thing to note is that a higher fuel gas to non-combustibles ratio does not always mean more Btu's per scf. When the cone nitrogen was reduced from 138 to 48 scfh the fuel gas to non-combustibles ratio increased from 40 to 42. However, the gas heating value decreased from 114.2 to 112.2 Btu/scf. This reduction was due to a loss in methane production. This suggests that lowering the nitrogen dilution increased the combustion zone temperature as supported by increased carbon conversion, from 86.7 to 89.5%. Methane is thought to decompose in the presence of oxygen at higher temperatures.

INVESTIGATION OF CLINKER FORMATION IN RUN 10. After two consecutive successful runs, the fluid bed gasifier clinkered during run the final test run in 1994, (run 10). The success of the previous runs was credited to the change in the introduction of reactants into the gasifier. These changes increased the momentum of the jet, thus not allowing the coal and char particles to agglomerate in the cone of the gasifier. The improvement of the inlet flows prevented clinkers from forming, but for run 10, the outlet flow caused a clinker to form.

Part of run 10 was dedicated to a parametric study of the gasifier by performing small step changes to all of the inlet flows in both the positive and negative directions. No problems arose until the adjustment of the gasifier pressure from 425 psig to 400 psig occurred. This action caused the back pressure control valves to react first to decrease the pressure slightly, then to increase the pressure slightly, and finally oscillated until a new steady state was found.

During the back pressure control valve's travel, data indicate that the valve had on one occasion stopped the outlet flow of gases completely. This is when it is believed that the gasifier slumped then clinkered. The stoppage of outlet flow caused the bed to slump as indicated by an out-of-range indication of the lowest change in pressure indicator, a stoppage of the underflow nitrogen flow, temperature increases in the lower bed, and a large, continual decrease in inlet steam rate; Even the inlet flows showed abnormal glitches at the time of the step change.

The response of the gasifier and the controllers indicated that the back pressure control was tuned for a quick, under damped response to disturbances. Previous to this run, the back pressure control set point had not been changed. Even though an under damped response allows for tight control of the gasifier pressure, a price is paid in outlet flow. The control of a reacting process should be of deliberate, over damped responses to prevent any uncontrollable reactions from occurring.

INSTRUMENTATION PROBE. A real time indication of bed status is important to the operations of any type of fluidized bed. They could indicate stability and mixing in an otherwise sealed process. FETC researchers will place two differential pressure probes at the same radial but at different vertical positions in the FBG at a location above the jetting region. The objective of this probe is to identify dynamic activities within the bed. The principal identified activity is bubble movement, slugging, and mixing. The measurement method is by high speed differential pressure measurements. The frequencies of interest within a fluidized bed are below 20 Hz (Clark and Atkinson, 1987). Future probes will be designed for good response up to 20 Hz and the sampling system designed to sample at 100 Hz.

Clark and Atkinson explained the design of these types of probes. Based on their design criteria, probes being built for the FBG will comprise of 1/4 inch tubing that is less than 3.5 ft long, two check valves to protect the sensitive pressure transmitter, and two critical flow orifices to facilitate a continuous purge of 1 m/s of nitrogen at the probe's tip. Frequency and modeled results of the designed probe indicate accurate results for frequencies less then 20 Hz. A schematic of the final design is shown in Figure 7.

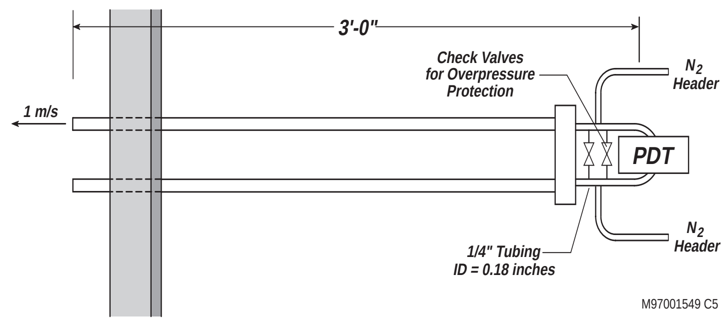


**Figure 7.** Schematic of the instrumentation probe.

Published reports and data previously collected at FETC indicated the Figure 8 represents what a typical data set should look like. Significant pressure fluctuations can be seen in this figure. If two such time series placed in the same radial plan but shifted in the z-plane were taken simultaneously, a cross-correlation would pinpoint the bubble movement and estimate the bubble velocity in the FBG.

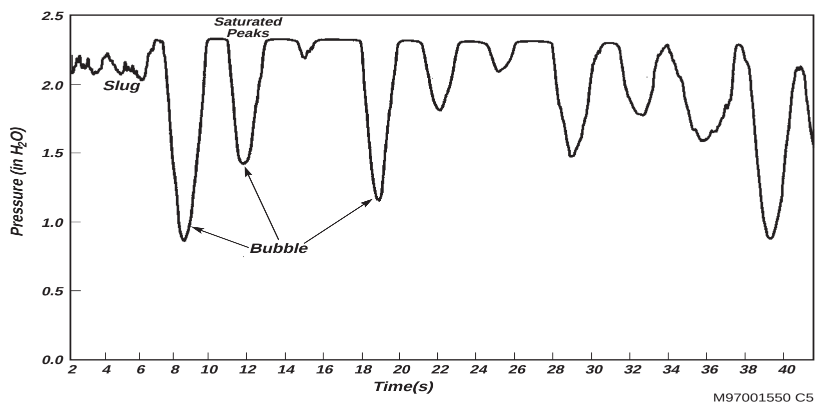


**Figure 8**. High speed differential pressure measurement of the FBG taken during the test run in 1996.

## CONCLUSIONS

Jet momentum was found to be a primary parameter used to control mixing and increasing jet momentum resulted in marked improvements in operational reliability. Improved mixing was measured indirectly by observing a combination of decreased peak temperatures, increased average bed temperatures, and/or additional gas-make and carbon conversion. Operational reliability was measured by the occurrence of clinkers, thermal excursions, and coal agglomerates. FBG tests have demonstrated that, under these conditions, increased momentum in the center region of jet with the feed coal markedly enhanced jet penetration and vertical mixing within the fluid bed. By contrast, jet momentum obtained by increasing the velocity in the annular region of the nozzle was less effective in promoting solids recirculation, solids mixing, and jet penetration. Adjustments in which the center jet momentum was increased resulted in more uniform temperature profiles across the bed and increased carbon conversion. This, along with fluidization under the central feed jet were necessary to eliminate clinkering. An underflow fluidization of 1.5-3 times the minimum fluidization velocity were found to be successful.

Conical jets were also found to promote mixing and recirculation of hot char into the air/coal jet. These jets resulted in an increase in carbon conversion and improved operating stability in the reactor. This was accomplished using both underflow fluidizing gas, jets in the bottom cone, and by removing larger more dense solids which accumulate in the bottom of the bed periodically.

The improvements to the FBG over the past 4 years helped to make the FBG a world class project. FETC researchers continually increased gas-make and carbon conversion while successfully operating over all planned hours and equalizing the gasifier temperatures. Each year of operation added an incremental step toward the FBG's objective. In 1993 the modified start-up procedures helped operators overcome barriers that made start-up unreliable. In 1994 two-major improvements, that increased the momentum of the reactants in the jet, lowered the possibility of developing localized hot spots. In 1995 the addition of a series of eight jets configured in a ring around the combustion area of the gasifier further reduced localized hot spots. With these changes in place, the reliability of the gasifier increased to where automation and small research projects could operate on the FBG.

In 1996 a new distributed control system added flexibility to gasifier operations and assisted in the implementation of a high speed, differential probe and the production of more coal gas. The DCS affords essential flexibility to the integration and controllability needed for a research project.

Other improvements that add to the FBG's operation includes a modern Distributed Control System, high pressure purge system on the pressure transmitters, improved vent and stack operation, and improved process documentation problems and upgrades. Because the FBG's reliability of producing actual coal gas, it is poised to be used for experimental testing and research of new feedstocks and configurations.

## CRADA OPPORTUNITIES

FETC has successfully operated the FBG for several years. This paper records great strides towards a commercial technology. Learning to operate this process has built capabilities into FETC personnel that is second to none. As well as operating the gasifier to produce dirty coal gas, other experiments could be tried.  For instance, FETC can gasify alternative feedstocks; add in-vessel, particulate clean-up devices; instrument the present process for different insight; integrate and control additional processes through the DCS; and collect data.

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