JV TASK 108 – CIRCULATING FLUIDIZED-BED COMBUSTION AND COMBUSTION TESTING OF TURKISH TUFANBEYLI COAL

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ABSTRACT

Two combustion tests were performed at the Energy & Environmental Research Center (EERC) using Tufanbeyli coal from Turkey. The tests were performed in a circulating fluidized-bed combustor (CFBC) and a pulverized coal-fired furnace, referred to as the combustion test facility (CTF). One of the goals of the project was to determine the type of furnace best suited to this coal. The coal is high in moisture, ash, and sulfur and has a low heating value.

Both the moisture and the sulfur proved problematic for the CTF tests. The fuel had to be dried to less than 37% moisture before it could be pulverized and further dried to about 25% moisture to allow more uniform feeding into the combustor. During some tests, water was injected into the furnace to simulate the level of flue gas moisture had the fuel been fed without drying. A spray dryer was used downstream of the baghouse to remove sufficient sulfur to meet the EERC emission standards permitted by the North Dakota Department of Health.

In addition to a test matrix varying excess air, burner swirl, and load, two longer-term tests were performed to evaluate the fouling potential of the coal at two different temperatures. At the lower temperature (1051°C), very little ash was deposited on the probes, but deposition did occur on the walls upstream of the probe bank, forcing an early end to the test after 2 hours and 40 minutes of testing. At the higher temperature (1116°C), ash deposition on the probes was significant, resulting in termination of the test after only 40 minutes.

The same coal was burned in the CFBC, but because the CFBC uses a larger size of material, it was able to feed this coal at a higher moisture content (average of 40.1%) compared to the CTF (ranging from 24.2% to 26.9%). Sulfur control was achieved with the addition of limestone to the bed, although the high calcium-to-sulfur rate required to reduce SO_2 emissions resulted in heat loss (through limestone calcination) and additional ash handling. A more efficient downstream sulfur scrubber capable of operation at a much lower Ca/S ratio would result in significantly higher boiler efficiency for this coal. At the operating temperature of a typical CFBC, bed agglomeration and convective pass fouling are not likely to be significant problems with this fuel.

Compared to pulverized coal-firing, CFBC technology is clearly the better choice for this fuel. It provides more efficient sulfur capture, lower NO_x emissions, better solids-handling capability, and can utilize a wetter feedstock, requiring less crushing and sizing. The lower operating temperature of CFBC boilers (820°C) reduces the risk of fouling and agglomeration. Care must be taken to minimize heat loss in the system to accommodate the low heating value of the coal.

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LIST OF ACRONYMS

EERC – Energy & Environmental Research Center
CFBC – Circulating fluidized-bed combustor
CTF – Combustion test facility
FEGT – Furnace exit gas temperature
SEMPC – Scanning electron microscope point count
FACT – Facility for the Analysis of Chemical Thermodynamics
SCMO – Solid calcium magnesium oxide
EHX – External heat exchanger
XRFA – X-ray fluorescence analysis
LOI – Loss on ignition
ASME – American Society of Mechanical Engineers
PTC – Performance test code
EPRI – Electric Power Research Institute
FD – Forced draft
ID – Induced draft
CHX – Combustor heat exchangers
FGHX – Flue gas heat exchangers

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EXECUTIVE SUMMARY

Introduction

A new coal-fired electrical generation plant has been proposed for construction in Turkey. The Energy & Environmental Research Center (EERC) was initially contacted by Tuten, Ltd., of Istanbul, Turkey, to perform combustion testing for ENERJISA with a Turkish coal to provide data for the evaluation of the construction of this proposed power plant in Turkey. Pilot-scale testing was planned to allow evaluation of circulating fluidized-bed combustion (CFBC) technology versus pulverized coal (pc)-fired combustion for this power plant; pc-fired tests were performed in the EERC combustion test facility (CTF).

Fuel Properties

The Tufanbeyli coal was received by the EERC the second week in January 2007. The fuel was very wet and difficult to handle without drying. Samples of the fuel were air-dried in the EERC's coal preparation facility to between 32.5% and 36.5% moisture (from up to 58% moisture in the as-received samples). The portion of the fuel used for the CTF testing was remotely pulverized using a hammer mill to a size consist of approximately 67% less than 200 mesh (45 µm). The fuel was then allowed to further air-dry to about 25% moisture. At these moisture levels, the fuel was more easily handled, although some difficulty was encountered in transferring the fuel from the combustion system charge hopper to the feed hopper. Bridging of the fuel in the charge hopper was the primary difficulty encountered during testing.

For the CFBC testing, the remaining coal was air-dried enough to facilitate processing to −3/8 inches (9.5 mm). The sized coal was transferred to storage hoppers with a net capacity of about 3500 lb (1590 kg). Limestone, also shipped from Turkey, was crushed to −20 mesh and stored in drums. Table ES-1 shows the as-fired fuel analysis for the Tufanbeyli coal for both the CTF and CFBC testing.

PULVERIZED COAL-FIRED TESTING

Testing with pulverized coal was performed in the EERC CTF. The furnace capacity is approximately 45 kg/hr (791,000 kJ/hr) of a fuel with a moderately high Btu content.

The furnace exit gas temperature (FEGT) varied throughout the tests, ranging typically from 815° (half load) to 1115°C (full load). During flame stability tests, the FEGT was allowed to fluctuate while the firing rate remained constant. Firing the Tufanbeyli lignite at the lower end of the temperature range proved difficult because of problems encountered in feeding the fuel while maintaining an expected flue gas moisture content. For most of the flame stability tests,

Table ES-1. Analysis of Tufanbeyli Coal

	CTF Tests			CFBC Tests	
Fuel Description	1–4	5	6	Average	Range
Proximate Analysis, wt%					
Moisture	25.5	26.9	24.2	40.1	35.5 - 42.3
Volatile Matter	29.4	28.1	27.4	21.8	21.0 – 22.9
Fixed Carbon	13.6	13.1	14.2	11.5	10.2 – 13.6
Ash	31.5	31.9	34.2	26.7	23.9 – 29.9
Ultimate Analysis, wt%					
Carbon	23.8	23.2	24.3	20.3	19.0 – 21.5
Hydrogen	2.2	2.0	2.1	1.8	1.7 – 1.9
Nitrogen	0.8	0.8	0.8	0.7	0.7 – 0.7
Sulfur	3.2	3.2	3.2	2.2	2.1 – 2.5
Oxygen	13.0	12.0	11.1	8.2	6.7 – 9.2
Ash	31.5	31.9	34.2	26.7	23.9 – 29.9
Moisture	25.5	26.9	24.2	40.1	35.5 – 42.3
HHV[1], Btu/lb	3951	3846	4021	5431	5001–5916
HHV, kJ/kg	9191	8947	9354	7560	6920–8390

[1] Higher heating value.

moisture was added to the primary air line. This was abandoned when it was determined that there was not enough heat input in the primary airstream to sustain this level of moisture. To keep the total flue gas moisture content in the expected range, water was added above the flame at a rate consistent with the total expected flue gas moisture concentration.

Ash deposition tests were conducted at two temperatures, 1050° and 1116°C. Although there was significant deposit formation at the lower temperature, the deposit strength was low to moderate. For this reason, the FEGT was increased for the second ash-fouling test. Table ES-2 shows the Tufanbeyli coal operating parameters for each test period.

Test Results

The high concentration of sulfur in the coal resulted in SO_2 at the furnace exit in excess of 7000 ppm. The high sulfur concentrations may contribute to significant ash-fouling problems and are certainly a cause for concern with environmental regulations. Flue gas analysis of NO_x shows concentrations ranging from approximately 260 to 475 ppm at the furnace exit. For Test 2, NO_x levels appear to be reduced with reduced burner stoichiometry, even though temperature increased during Test 2C.

The swirl number is a dimensionless quantity related to the amount of “spin” imparted to the secondary airstream. It is the ratio of the tangential momentum to the radial momentum of the airstream. Preliminary observations of flame stability indicate that stable combustion could be attained under full-load conditions at a secondary air swirl setting similar to most low-rank fuels fired in the CTF. A swirl setting of 3.2 was used for all full-load tests. This corresponds to a swirl number of 0.418. Initially, at half-load conditions, the maximum swirl setting of 0 was used

Table ES-2. Tufanbeyli CTF Test Results

Test:	1A	2A	3A-1	4C	5	6
Date:	1/23/2007	1/24/2007	1/25/2007	1/25/2007	1/26/2007	2/1/2007
Start Time	16:10	17:07	16:40	19:06	13:52	9:30
Stop Time	18:16	18:43	17:17	19:17	17:34	10:09
Swirl Setting	3.2	3.2	0	3.2	3.2	3.2
Swirl Number	0.418	0.418	1.9	0.418	0.418	0.418
Fuel Feed Rate, lb/hr	149.6	151.7	79.8	83.6	142	147
Fuel Feed Rate, kg/hr	67.9	68.8	36.2	37.9	64.4	66.7
Fly Ash Carbon Content, %	0.44	0.44	0.57	NM[2]	NM	NM
Temperature, °F (°C)						
Primary Air	700 (371)	702 (372)	695 (368)	700 (371)	692 (367)	625 (330)
Secondary Air	760 (404)	767 (409)	717 (381)	630 (332)	752 (400)	690 (366)
Furnace Exit	1906 (1041)	1834 (1001)	1533 (834)	1607 (875)	1923 (1051)	2040 (1116)
Combustion Efficiency, %	99.8	99.7	99.7	NM	NM	NM
Dry Gas Analysis						
Furnace Exit						
O_2, %	3.18	3.47	4.31	4.54	3.01	2.32
CO_2, %	15.8	15.6	14.5	14.3	16.2	16.4
CO, ppm	89.5	9.8	41.2	31.9	52.0	20.1
SO_2, ppm	6057	6392	5939	6072	6852	7217
NO_x, ppm	433.3	413.8	307.8	431.3	405	N/A
Baghouse Exit						
O_2, %	3.21	3.72	4.51	4.52	3.26	3.07
CO_2, %	15.5	15.2	14.8	14.2	15.4	15.7
CO, ppm	139.7	11.6	50.2	40.0	45.5	23.5
SO_2, ppm	>5000[1]	6570[3]	>5000[1]	>5000[1]	6493[3]	>5000[1]
NO_x, ppm	365.9	379.4	281.6	394.5	404	554
Spray Dryer Exit						
O_2, %	3.81	4.78	NM	NM	3.90	NM
CO_2, %	14.9	13.9	NM	NM	15.0	NM
CO, ppm	160.9	11.2	NM	NM	62.0	NM
SO_2, ppm	3100	4674	NM	NM	3975	NM
NO_x, ppm	361.1	319.2	NM	NM	409	NM

[1] The data logger for analyzer Mobile 1 does not record values over 5000 ppm for SO_2.

[2] Not measured.

[3] Data taken from logbook.

to help maintain a stable flame. This corresponds to a swirl number of 1.9. However, the setting was quickly changed back to 3.2 as this was found to provide an adequately stable flame.

The 0% stoichiometry setting results in the greatest amount of unburned carbon for the first three test periods. Aside from that, the variability in the data makes direct interpretation difficult. Therefore, the data were statistically analyzed to determine if any of the main test variables had a statistically significant impact on unburned carbon. The analysis was again performed using MINITAB statistical software. Although none of the variables had a statistically significant effect on baghouse ash carbon content, it did appear that excess air had a practically significant impact. On average, increasing excess air from 15% to 20% reduced the carbon in the baghouse ash by 0.14%.

Two ash-fouling tests were run to determine the level of ash buildup using fouling probes to simulate ash buildup on boiler tubes. One lower-temperature test (Test 5) and one higher-temperature test (Test 6) were completed. The fouling probes were made of 304 stainless steel and a Cr–Mo composite consisting of 0.07% C, 0.25% Si, 0.48% Mn, 0.017% P, 0.002% S, 2.09% Cr, 0.92% Mo, 0.004% Sn, and 96.167% Fe.

Test 5 was stopped after 2 hr 40 min because of ash deposition in the duct. The system was shut down to evaluate the fouling probes. Upon removal of the probes, very little ash deposit was found, and it was friable. The majority of the ash deposit formed just ahead of the probe bank.

Test 6, a higher-temperature test (1116°C), was intended to evaluate the upper limit of furnace exit temperature. Almost immediately, signs of duct plugging were noted in the system. Ten minutes into the test, the coal feed was reduced by 3 lb/hr, then slowly reduced by another 12 lb/hr over the next 20 minutes. After 39 minutes, the test was ended because of severe plugging in the system. Removal of the fouling probes indicated severe ash buildup. The deposit was initiated at the tube metal surface, quickly bridging between tubes and blocking the duct. The deposit was highly sintered and appeared to present a significantly more difficult problem for sootblowers. Based on these results, it is recommended that the FEGT be controlled to between 1050° and 1070°C, limiting any excursions above these temperatures to maintain manageable sootblower operations.

CIRCULATING FLUIDIZED-BED TESTING

The EERC pilot-scale CFBC operates at 1 MW_{th} input, typically feeding 100 to 200 kg/hr coal. Several coal feed system modifications were completed prior to testing to enable the system to handle a much larger coal and limestone feed rate than is typically fed through this system.

The test matrix called for three test periods without limestone feed to evaluate inherent sulfur capture. SO_2 emissions were as high as 6000 ppm, which exceeded the EERC-permitted emission limits, so the second no-limestone test (full load at a lower temperature) was omitted. The third test without limestone was a half-load test, so the SO_2 emissions were within permitted levels, in spite of very high concentrations in the flue gas. The next two tests were at a medium level of sulfur capture at two different temperatures, 820° and 850°; Tests 6–8 were meant to be high-sulfur-capture tests at the same temperatures as 4 and 5, plus a third test at a higher temperature. However, the highest temperature could not be achieved, so Test 8 was omitted. Tests 9 and 10 were half-load tests at moderate and high sulfur capture, respectively. Low load was achieved by reducing both the coal feed and airflow rates.

TEST RESULTS

Because of the high moisture content of the coal, it was difficult to maintain consistent coal feed, which resulted in fluctuations of temperature and SO_2 emissions. The high ash content, combined with the high limestone feed rate required for the desired level of sulfur removal, resulted in a buildup of solids in the bed, which could be removed either by draining material from the bottom of the bed or draining the ash captured by the secondary cyclone (normally that ash is recycled into the system). Both methods have an impact on the performance of the EERC CFBC. Draining bed material decreases bed temperature, and draining secondary cyclone ash increases bed temperature and decreases sulfur retention. In a full-scale system specifically designed for the high moisture, high ash, and high sulfur content of this coal, the primary cyclone would be sized to optimize solids recirculation, so that a higher percentage of solids removal would be through the baghouse.

The ratios of bottom ash to total ash were calculated for each test period. For full-load tests with limestone addition, the bottom-to-total ash split ranged from 38% to 57%. Bottom ash was drained periodically to try to maintain a consistent amount of bed inventory (as measured by pressure drop across the bed). The secondary cyclone drain rate was kept at a minimum to try to retain as much unreacted limestone in the bed as possible, although cyclone ash was drained (rather than recycled) occasionally to trim the combustor temperature, as described earlier. The bottom-to-total ash split was much lower for the tests without limestone and significantly higher for Test 10 (at low load), where no material was taken from the secondary cyclone. No cyclone ash was drained during Test 9 either, but there was an increase in bed dP during that test period, indicating that more bed material could have been drained during that test.

This fuel requires relatively high limestone and coal feed rates compared to operation with other fuels. At full-load conditions, the bed was turning over approximately every 4.5 hours, assuming an inventory of about 365 kg of bed material. Based upon the feed rates, drain rates, and overall solids compositions, there was no indication of any significant bed material attrition or agglomeration of the limestone or coal ash. In a full-scale system, the best location for draining bed material would be from the loop seal. The solids are somewhat cooler here, so less energy would be lost as opposed to draining large amounts of bed material from the combustor. Also, the size distribution of the solids in the loop seal were very similar to that of the bed material, so it did not appear that there were a lot of larger particles accumulating in the bed that would have to be drained on a frequent basis.

This test matrix was better suited to evaluating operability with coal as opposed to obtaining sufficient data to evaluate the emission trends. While the narrow range of operating temperatures selected for this testing are likely typical of a-full scale operational plant, a wider range would have made it easier to establish the emission trends for this coal. Additionally, the difficulty with achieving the high-temperature Test No. 8 condition resulted in not having any high-temperature data for comparison purposes. The behavior of the system with respect to the flue gas emissions at half-load conditions was significantly different compared to full-load operation.

The temperature range over which these tests were conducted was not sufficiently large to determine the optimum temperature for sulfur capture. Additionally, the two very high sulfur capture rates specified (requiring sulfur retention rates of 96.5% and 98.3%) are not sufficiently different to establish a trend on how the calcium-to-sulfur ratio will affect sulfur retention at a given operating temperature. For full-load operation, it appears that better sulfur capture was possible at approximately 816°C compared to operation at approximately 843°C. Sulfur capture was significantly affected by both the limestone feed rate and by the drain rate of the solids collected by the secondary cyclone. The most effective limestone in the system for good sulfur capture is typically collected by the secondary cyclone. Normally, all of the solids collected by the secondary cyclone would be fed back into the system to be available for sulfur capture. For this coal, the amount of solids to the secondary cyclone tended to increase more rapidly than is typical for most coals. Each time these cooler solids were returned to the downcomer, they would progressively force the combustor operating temperature to slowly decrease. Draining a consistent portion of these solids permitted operation at a more steady and consistent temperature as long as steady coal feed rates, limestone feed rates, and combustion air addition flows were maintained. It would be projected from this testing that with calcium-to-sulfur addition rates approaching 3.5, it should be possible to obtain sulfur dioxide emissions down to 200 mg/Nm^3. This will ultimately be dependent upon such factors as primary cyclone efficiency and limestone properties.

At low-load conditions, high sulfur retention rates were possible at lower Ca/S ratios compared to full-load conditions. It is likely that lower velocities in the combustor allowed the limestone to stay in the system longer for better sulfur capture instead of being blown out at the higher velocities at full-load conditions. Because of the short duration of these tests, no secondary cyclone ash was removed until the test period was over, also contributing to the improved sulfur retention.

High sulfur capture was achieved during this testing at high added Ca/S ratios often greater than 3. In a full-scale system, it would be expected that similar ratios of Ca/S would be required for the same degree of sulfur capture. It will again depend significantly on the full-scale system geometry and, ultimately, the efficiency of the primary cyclone in that system. In the EERC CFBC, it is likely that we are capable of returning more of the finer-sized reactive lime compared to a full-scale system. A full-scale system is more likely to have longer solids residence times for the lime to react for good sulfur capture, mainly because of its increased height. These two factors should somewhat offset each other. It would likely be much more efficient both in terms of the reduction of overall limestone consumption and higher boiler efficiency to only capture a portion of the coal sulfur in-bed and rely on a posttreatment sulfur capture system to control sulfur dioxide emissions to the desired level. An economic analysis will be necessary to determine to what degree that only in-bed sulfur capture be relied upon compared to incorporating a posttreatment sulfur capture system.

We were able to successfully operate at the low-load test parameters with this coal, with only a moderate drop in the operational combustor average bed temperature down to approximately 788°C.

Because CFBC boilers operate at a lower temperature than pc-fired boilers, NO_x emissions tend to be significantly lower. N_2O was measured only at the baghouse outlet and was not measured at all during the tests without limestone addition, since high SO_2 interferes with the N_2O measurement in that analyzer. Unlike NO_x, N_2O emissions increase with decreasing temperature and tend to be higher in CFBC boilers than in pc-fired units. The emissions for each test period are shown in Table ES-3.

Table ES-3. CFB Flue Gas Emission Data

	Test 1	Test 3	Test 6 b	Test 5	Test 4	Test 7	Test 6	Test 9	Test 10
Combustor Outlet									
O_2, %	4.10	3.70	3.70	3.60	3.30	3.40	3.50	3.10	3.60
CO Content, ppm	22	57	82	61	38	38	96	39	79
CO Content[1], ppm	20	49	71	53	32	32	82	33	68
CO Emission, mg/Nm^3 @ 6% O_2	24.4	61.8	88.9	65.7	40.3	40.5	102.9	40.9	85.1
CO_2 Content, %	16.0	15.6	17.1	17.1	17.6	17.6	16.9	16.8	16.4
CO_2 Content[1], %	13.5	13.5	14.8	14.7	14.9	15.0	14.5	14.1	14.1
NO_x Content, ppm	89	34	134	131	139	148	128	34	57
NO_x Content[1], ppm	79	29	116	113	118	126	110	28	49
NO_x Emission, mg/Nm^3 @ 6% O_2	162	61	239	232	242	259	225	59	101
SO_2 Content, ppm	>5000	>5000	129	198	192	76	176	242	39
SO_2 Content[1], ppm			112	171	163	65	151	203	34
SO_2 Emission, mg/Nm^3 @ 6% O_2			320	488	465	185	431	579	96
SO_2 Retention[3], %			98.1	97.1	97.4	98.9	97.5	96.5	99.4
EA	24.1	21.0	21.5	20.8	18.8	19.5	20.0	17.2	20.5
Baghouse Outlet									
O_2, %	4.60	5.60	5.50	4.70	4.20	4.30	4.90	6.10	5.20
CO Content, ppm	51	63	56	48	41	39	61	34	79
CO Content[1], ppm	47	61	54	44	37	35	57	34	75
CO Emission, mg/Nm^3 @ 6% O_2	58.3	76.7	67.7	55.2	45.8	43.8	71.0	42.8	93.8
CO_2 Content, %	15.8	15.4	16.0	15.6	16.1	16.2	16.0	15.7	15.9
CO_2 Content[1], %	14.5	15.0	15.5	14.4	14.4	14.6	14.9	15.8	15.1
NO_x Content, ppm	95	33	130	130	141	151	116	32	50
NO_x Content[1], ppm	86	32	125	119	126	136	108	32	48
NO_x Emission, mg/Nm^3 @ 6% O_2	178	66	258	245	258	279	221	65	98
N_2O Content, ppm	NA	NA	67	55	43	38	59	15	24
N_2O Content[1], ppm			65	51	38	34	55	15	23
N_2O Emission, mg/Nm^3 @ 6% O_2			127.4	99.8	74.9	66.9	107.4	29.1	44.9
SO_2 Content, ppm	5407	4855	110	185	169	66	153	3	176
SO_2 Content[1], ppm	4945	4729	106	170	151	59	143	3	167
SO_2 Emission, mg/Nm^3 @ 6% O_2	14,130	13,511	304	486	431	169	407	9	477
SO_2 Retention[3], %	16.3	24.1	98.2	97.0	97.5	99.0	97.7	100.0	97.4
Ca/S ratio (ls[2] only)	0.00	0.00	2.69	2.78	3.33	3.44	3.01	1.41	1.66
Ca/S ratio (total)	0.61	0.68	3.42	3.59	4.19	4.24	3.71	2.10	2.35
Ca Utilization (ls[2] only)	NA	NA	36.4	34.9	29.3	28.8	32.4	70.9	58.6
Ca Utilization (total)	26.7	35.4	28.7	27.1	23.3	23.3	26.3	47.5	41.4
Alkali-to-Sulfur	0.63	0.69	3.44	3.60	4.20	4.26	3.72	2.12	2.37
Alkali Utilization	26.1	34.7	28.6	26.9	23.2	23.2	26.2	47.2	41.2
Av Comb. Temp., °C	852	841	815	826	847	846	811	835	801
Moisture in Flue Gas, vol%	21.9	22.7	22.6	20.8	21.3	22.0	22.3	22.3	20.6
Fuel Carbon, %	21.5	21.2	20.7	20.1	19.0	20.2	19.5	19.5	21.3
Fuel Sulfur, %	2.4	2.4	2.1	2.1	2.1	2.2	2.1	2.1	2.5

[1] Corrected to 6% O_2.

[2] Moisture-free coal carbon and sulfur values in the sulfur retention calculation.

[3] Limestone.

No agglomeration tendencies were noted with this coal during this test. There was a very minor layer of ash observed on the fouling tubes located in the simulated convective pass region for this system. It would not be expected that any significant amount of sootblowing would be required in a full-scale system. However, there will be a significant quantity of solids produced by the combustion of this coal that will have to be handled. If possible, solids should be drained from the loop seal region where they are the coolest to have the least impact on the system to maintain operational temperatures.

CONCLUSIONS

Compared to pulverized coal-firing, CFBC technology is clearly the better choice for this fuel. It provides more efficient sulfur capture, lower NO_x emissions, better solids-handling capability, and can utilize a wetter feedstock, requiring less crushing and sizing. The lower operating temperature reduces the risk of fouling and agglomeration. Care must be taken to minimize heat loss in the system to accommodate the low heating value of the coal.

This fuel should be dried as much as possible in any manner that is practical. Conservative design practices should be used for the fuel-handling system. It is recommended to operate this system at as high a velocity as is possible in a full-scale system to help increase the thermal loading into the system. If low sulfur dioxide emissions will be required, installing a post-sulfur dioxide treatment system should be considered. If possible, solids should be drained from the loop seal region where they are the coolest to have the least impact on the system to maintain operational temperatures. Bed material agglomeration and convective pass fouling are not expected to be a problem with this fuel in a full-scale fluidized-bed boiler system.

JV TASK 108 – CIRCULATING FLUIDIZED-BED COMBUSTION AND COMBUSTION TESTING OF TURKISH TUFANBEYLI COAL

INTRODUCTION

A new coal-fired electrical generation plant has been proposed for construction in Turkey. The Energy & Environmental Research Center (EERC) was initially contacted by Tuten, Ltd., of Istanbul, Turkey, to perform combustion testing for ENERJISA with a Turkish coal to provide data for the evaluation of the construction of this proposed power plant in Turkey. Pilot-scale testing was planned to allow evaluation of circulating fluidized-bed combustion (CFBC) technology versus pulverized coal (pc)-fired combustion for this power plant. The results of the CFBC testing are provided in this report.

The EERC evaluated the combustion performance of Turkish Tufanbeyli coal in two combustion systems to evaluate the potential to use Tufanbeyli coal for future power-generating capacity. This report details the results of the combustion testing that took place in the EERC's combustion test furnace (CTF) on January 25–27 and February 1, 2007. The CTF was used to fire the fuel in a series of tests to determine the effect of boiler-operating parameters on flame stability, combustion efficiency, and gaseous emissions at simulated full load and half load. A second set of tests was dedicated to determining the boiler tube ash-fouling potential of the fuel.

FUEL ANALYSES

The Tufanbeyli coal was received by the EERC the second week of January 2007. The fuel was very wet and difficult to handle without drying. Samples of the fuel were air-dried in the EERC's coal preparation facility to between 32.5% and 36.5% moisture (from up to 58% moisture in the as-received samples). This fuel was remotely pulverized using a hammer mill to a size consist of approximately 67% less than 200 mesh (45 µm). The fuel was then allowed to further air-dry to about 25% moisture. At these moisture levels, the fuel was more easily handled, although some difficulty was encountered in transferring the fuel from the combustion system charge hopper to the feed hopper. Bridging of the fuel in the charge hopper was the primary difficulty encountered during testing.

Table 1 shows the as-fired fuel analysis for the Tufanbeyli coal. The fuel had an average as-fired moisture content of 25.5% and an average heating value of 9164 kJ/kg. The nitrogen concentration was determined to be 1.1% on a dry basis. The sulfur concentration for the coal was relatively high, averaging 4.3% on a dry basis. The theoretical maximum sulfur emission rate for the fuel was calculated at 7.79 kg SO_2/MJ (16.29 lb SO_2/MMBtu), resulting in a flue gas concentration of 8357 ppm at 15% excess air. The ash level of the coal was high, 43.7% on a dry basis, which could potentially cause significant fouling. Fuel calcium, sodium, and potassium were low to moderate, indicating a moderate potential for deposit strength development.

Table 1. Analysis of Tufanbeyli Coal

	Tests 1–4		Test 5		Test 6	
Fuel Description	As-Fired	H_2O-Free	As-Fired	H_2O-Free	As-Fired	H_2O-Free
Proximate Analysis, wt%						
Moisture	25.5	N/A	26.9	N/A	24.2	N/A
Volatile Matter	29.4	39.5	28.1	38.4	27.4	36.2
Fixed Carbon	13.6	18.2	13.1	17.9	14.2	18.7
Ash	31.5	42.3	31.9	43.6	34.2	45.1
Ultimate Analysis, wt%						
Carbon	23.8	31.9	23.2	31.7	24.3	32.1
Hydrogen	5.0	2.8	5.0	2.8	4.8	2.8
Nitrogen	0.8	1.1	0.8	1.1	0.8	1.1
Sulfur	3.2	4.3	3.2	4.4	3.2	4.2
Oxygen	35.7	17.5	35.9	16.4	32.6	14.6
Ash	31.5	42.3	31.9	43.6	34.2	45.1
Moisture	25.5	N/A	26.9	N/A	24.2	N/A
HHV,[1] Btu/lb	3951	5304	3846	5262	4021	5307
HHV, kJ/kg	9191	12339	8947	12241	9354	12346
Ash Analysis, wt%	As-Received	BaO-, SO_3-Free	As-Received	BaO-, SO_3-Free	As-Received	BaO-, SO_3-Free
SiO_2	43.5	48.9	44.3	49.2	42.9	49.07
Al_2O_3	21.5	24.2	22.2	24.7	20.8	23.79
Fe_2O_3	9.02	10.15	8.8	9.8	8.59	9.83
TiO_2	0.95	1.12	0.95	1.05	0.91	1.04
P_2O_5	0.40	0.45	0.41	0.46	0.40	0.46
CaO	9.5	10.7	9.4	10.4	9.9	11.32
MgO	1.73	1.95	1.71	1.9	1.70	1.94
Na_2O	0.25	0.28	0.30	0.33	0.22	0.25
K_2O	1.97	2.2	1.95	2.17	1.95	2.23
SO_3	11.03	N/A	9.84	N/A	12.48	N/A
BaO	0.09	N/A	0.1	N/A	0.1	N/A
SrO	0.02	0.02	0.2	0.02	0.02	0.02

[1] Higher heating value.

Advanced Fuel Analyses

The Tufanbeyli coal was analyzed by computer-controlled scanning electron microscopy (CCSEM). Table 2 shows the results. The unclassified portion is quite high, but based on the unclassified composition, it appears most likely that it is very fine pyrite grains in a clay matrix.

Table 3 shows the chemical fractionation results for the Turkish coal. The sulfur was completely extracted by the solvents. Each solvent removed approximately the same amount of sulfur, indicating that the sulfur was present in many forms. The water removes the soluble inorganics like NaCl. The ammonium acetate removes inorganic ion-exchangeable cations associated with organic acid groups. Hydrochloric acid removes elements held in coordination complexes within the organic structure as well as acid-soluble minerals such as carbonates,

Table 2. Tufanbeyli Coal CCSEM Results

	Particle-Size Distribution, μm						
	1.0–2.2	2.2–4.6	4.6–10.0	10.0–22.0	22.0–46.0	46.0–00.0	Totals
Quartz	1.8	2.7	1.5	1.7	0.5	0.1	8.2
Iron Oxide			0.1		0.2		0.3
Rutile	0.1	0.1	0.3	0.1			0.6
Calcite			0.1	0.1	0.3	0.1	0.7
Montmorillonite	0.1	0.1					0.2
K Al Silicate	1.0	3.7	1.2	0.9	0.3		7.2
Fe Al Silicate	1.9	1.9	0.6	0.6	0.4		5.4
Ca Al Silicate				0.1	0.3	0.5	0.9
Na Al Silicate	0.2	0.9	0.3	0.2			1.6
Mixed Aluminosilicate	3.3	2.2	0.3	0.1	0.1		6.0
Pyrrhotite	2.1	2.5	5.1	8.3	6.2	0.6	24.6
Oxidized Pyrrhotite	0.1		0.2		0.2		0.4
Gypsum	0.1	0.4	0.1	0.5	1.1	1.3	3.4
Apatite						0.1	0.1
Gypsum/Al Silicate	0.8	0.5	0.2	0.1	0.2	0.1	2.0
Si-Rich	1.2	1.4	0.1		0.1		2.7
Ca-Rich							0.1
Unclassified	13.6	13.2	4.2	2.6	1.7	0.4	35.7
Totals	26.2	29.6	14.4	15.4	11.6	3.0	100.0

Table 3. Chemical Fractionation Results for the Turkish Coal

Element	Initial Percentage	Percentage Removed by H_2O	Percentage Removed by NH_4OAc	Percentage Removed by HCl	Total Percent Remaining
Silicon	48.52	0	0	0	100
Aluminum	22.20	0	0	7	93
Iron	8.30	4	0	37	59
Titanium	1.05	4	0	0	96
Phosphorus	0.41	4	0	29	67
Calcium	8.58	8	42	37	13
Magnesium	1.73	3	8	9	80
Sodium	0.27	10	0	0	90
Potassium	2.19	2	0	0	98
Sulfur	6.59	37	32	31	0

sulfates, and oxides. The calcium was primarily removed by the NH_4OAc solvent, 42% remaining, and HCl solvent, 37% remaining. Iron, 59% remaining, and phosphorus, 67% remaining, were primarily removed by HCl.

EXPERIMENTAL

Test Matrix

The CTF was used to fire the fuel in a series of tests to determine the effect of boiler operating parameters on flame stability, combustion efficiency, and gaseous emissions at simulated full load and half load. A second set of tests was dedicated to determine the boiler tube ash-fouling potential of the fuel. Table 4 provides a brief description of the test periods utilized.

Not all test periods were performed. Tests 4A and 4B were abandoned because of anomalous effects noted in Test Periods 3A through 4C related to NO_x emissions at half load. The NO_x emissions at half load trended opposite of normal when looking at the relationship of burner stoichiometry to NO_x. This may have been due to significant flame stability issues encountered when firing at half load. The test furnace does not simulate half load well because it involves reducing the flow through the one and only burner. In an industrial application, half load is achieved by reducing the number of burners firing rather than reducing the flow to any specific burner.

A more thorough examination of the NO_x emissions generated at half load will be required to characterize these operating conditions.

Table 4. Tufanbeyli Test Matrix

Test Number	Load	Excess Air	Burner Stoichiometry	Swirl Setting	H_2O Addition, cm^3/min
1A	Full	15%	+15%	3.2	150
1B	Full	15%	0%	3.2	150
1C-1	Full	15%	−15%	2.0	150
1C-2	Full	15%	−15%	3.2	150
1D	Full	15%	+15%	3.2	0
2A	Full	20%	+15%	3.2	150
2B	Full	20%	0%	3.2	150
2C	Full	20%	−15%	3.2	150
3A-1	Half	20%	+15%	0	0
3A-2	Half	20%	+15%	3.2	0
3B	Half	20%	0%	3.2	0
3C	Half	20%	−15%	3.2	0
4A	Half	15%	+15%	3.2	0
4B	Half	15%	0%	3.2	0
4C	Half	15%	−15%	3.2	0
5	Full	15%	+15%	3.2	0
6	Full	15%	+15%	3.2	0

Operating Parameters

The furnace exit gas temperature (FEGT) varied throughout the tests, ranging typically from 815° (half load) to 1115°C (full load). During flame stability tests, the FEGT was allowed to fluctuate while the firing rate remained constant. Firing the Tufanbeyli lignite at the lower end of the temperature range proved difficult because of problems encountered in feeding the fuel while maintaining an expected flue gas moisture content. For most of the flame stability tests, moisture was added to the primary air line. This was abandoned when it was determined that there was not enough heat input in the primary airstream to sustain this level of moisture. To keep the total flue gas moisture content in the expected range, water was added above the flame at a rate consistent with the total expected flue gas moisture concentration.

Ash deposition tests were conducted at two temperatures, 1050° and 1116°C. Although there was significant deposit formed at the lower temperature, the deposit strength was low to moderate. For this reason, the FEGT was increased for the second ash-fouling test. Table 5 shows the Tufanbeyli coal operating parameters for each test period.

Testing Chronological Summary

Tests 1 and 2 were full-load tests that looked at two levels of excess air and three levels of burner stoichiometry (see Table 4). Test 1A was performed on January 23. The furnace was heated to temperature that night, but most of the morning was used to set up a deionized (DI) water injection system that allows simulation of the full moisture content of the coal. A full-scale system would certainly dry the coal down to near 25% moisture levels before feeding, but the test plan called for understanding the as-mined firing characteristics of the Tufanbeyli coal. After the feeder was set up, Test 1A was started around 16:00 on January 23. The test was shut down around 18:15 because of a plug in the spray dryer and a plug in the coal feed line.

Some continuation testing resumed the next day around 11:00, while much of the morning was spent dialing in the optimum parameters for the spray dryer. Test 1B started at 13:00 on January 24, with the U.S. Environmental Protection Agency (EPA) Method 26 test starting shortly thereafter. No major problems were reported during this test period. Test 1C started at around 16:00. Swirl was varied briefly from 2.0 to 3.2 to help understand flame stability issues. A Method 26 test was again performed. Adequate time existed to jump into Test 2 that evening, so Test 2A was started at 17:07. Upon the start of the test, the spray dryer needed to be cleaned again, so the test was shut down and then restarted again at 18:15. That test ran without major incident, and Test 2B was started at 18:50. That test also ran well, and the system was shut down around 19:50.

Test 2C was started at 8:30 on the morning of January 25. The test ran without incident and was shut down at 9:30. It was decided to run an additional test, Test 1D, to look at no H_2O addition into the system. This test started at around 11:00 and ended without major incident at 12:45. Next, the system was set up for the half-load testing to occur under test Condition 3. These runs were also set up with no water addition. Test 3A started at around 16:15 with the swirl setting to maximum (0). It was thought that this would increase flame stability. About halfway through the test, the swirl setting was returned to 3.2 to better compare with the full-load

Table 5. Tufanbeyli CTF Test Results

Test:	1A	1B	1C-1	1C-2	1D	2A	2B	2C
Date:	1/23/2007	1/24/2007	1/24/2007	1/24/2007	1/25/2007	1/24/2007	1/24/2007	1/25/2007
Start Time	16:10	13:30	15:54	16:06	11:13	17:07	18:55	8:50
Stop Time	18:16	14:09	16:04	17:04	12:43	18:43	19:45	9:30
Swirl Setting	3.2	3.2	2.0	3.2	3.2	3.2	3.2	3.2
Swirl Number	0.418	0.418	0.9	0.418	0.418	0.418	0.418	0.418
Fuel Feed Rate, lb/hr	149.6	150.2	151	150.4	151.4	151.7	145.9	148
Fuel Feed Rate, kg/hr	67.9	68.1	68.5	68.2	68.7	68.8	66.2	67.1
Fly Ash Carbon Content, %	0.44	0.82	NM[2]	0.52	NM	0.44	0.61	0.28
Temperature, °F (°C)								
Primary Air	700 (371)	705 (374)	700 (371)	708 (376)	696 (369)	702 (372)	709 (376)	713 (378)
Secondary Air	760 (404)	759 (404)	742 (394)	746 (397)	766 (408)	767 (409)	756 (395)	742 (395)
Furnace Exit	1906 (1041)	1882 (1028)	1854 (1012)	1851 (1011)	1881 (1027)	1834 (1001)	1820 (993)	1965 (1074)
Combustion Efficiency, %	99.8	99.5	99.7	99.7	NM	99.7	99.8	99.9
Dry Gas Analysis								
Furnace Exit								
O_2, %	3.18	2.76	2.31	2.60	3.20	3.47	3.82	3.62
CO_2, %	15.8	16.6	17.0	16.4	15.2	15.6	15.6	14.9
CO, ppm	89.5	9.6	14.7	11.0	20.7	9.8	31.7	63.1
SO_2, ppm	6057	6501	6856	6692	6763	6392	6545	6588
NO_x, ppm	433.3	476.3	351.7	403.6	370.0	413.8	343.6	277.2
Baghouse Exit								
O_2, %	3.21	3.28	NM	3.54	3.71	3.72	3.64	4.40
CO_2, %	15.5	15.7	NM	15.0	15.5	15.2	NM	14.7
CO, ppm	139.7	11.7	NM	13.7	27.8	11.6	NM	30.4
SO_2, ppm	>5000[1]	>5000[1]	NM	6784[3]	6949[3]	6570[3]	6724[3]	6730[3]
NO_x, ppm	365.9	399.2	NM	302.9	328.1	379.4	NM	255.6
Spray Dryer Exit								
O_2, %	3.81	4.63	3.78	4.14	4.53	4.78	4.92	5.06
CO_2, %	14.9	14.6	15.1	14.5	14.9	13.9	14.0	14.2
CO, ppm	160.9	11.3	27.2	13.3	23.9	11.2	58.9	182.8
SO_2, ppm	3100	2576	4959	4791	2311	4674	4106	3494
NO_x, ppm	361.1	368.6	267.9	291.4	309.8	319.2	240.9	261.0

[1] The data logger for analyzer Mobile 1 does not record values over 5000 ppm for SO_2.

[2] Not measured.

[3] Data taken from logbook.

Continued. . .

Table 5. Tufanbeyli CTF Test Results (continued)

Test:	3A-1	3A-2	3B	3C	4C	5	6
Date	1/25/2007	1/25/2007	1/25/2007	1/25/2007	1/25/2007	1/26/2007	2/1/2007
Start Time	16:40	17:26	17:15	18:31	19:06	13:52	9:30
Stop Time	17:17	17:45	18:18	18:46	19:17	17:34	10:09
Swirl Setting	0	3.2	3.2	3.2	3.2	3.2	3.2
Swirl Number	1.9	0.418	0.418	0.418	0.418	0.418	0.418
Fuel Feed Rate, lb/hr	79.8	79.5	80.4	82.3	83.6	142	147
Fuel Feed Rate, kg/hr	36.2	36.1	36.5	37.3	37.9	64.4	66.7
Fly Ash Carbon Content, %	0.57	0.57	0.71	0.60	NM[3]	NM	NM
Temperature, °F (°C)							
Primary Air	695 (368)	696 (369)	705 (374)	700 (371)	700 (371)	692 (367)	625 (330)
Secondary Air	717 (381)	718 (381)	687 (364)	641 (338)	630 (332)	752 (400)	690 (366)
Furnace Exit	1533 (834)	1551 (844)	1592 (867)	1600 (871)	1607 (875)	1923 (1051)	2040 (1116)
Combustion Efficiency, %	99.7	99.7	99.5	99.6	NM	NM	NM
Dry Gas Analysis							
Furnace Exit							
O_2, %	4.31	4.68	4.28	4.14	4.54	3.01	2.32
CO_2, %	14.5	14.2	14.6	14.6	14.3	16.2	16.4
CO, ppm	41.2	38.4	37.4	34.0	31.9	52.0	20.1
SO_2, ppm	5939	5773	5958	6103	6072	6852	7217
NO_x, ppm	307.8	262.3	308.2	424.8	431.3	405	N/A
Baghouse Exit							
O_2, %	4.51	4.81	4.48	4.00	4.52	3.26	3.07
CO_2, %	14.8	14.5	14.5	14.7	14.2	15.4	15.7
CO, ppm	50.2	44.9	43.4	40.9	40.0	45.5	23.5
SO_2, ppm	>5000[1]	5904[2]	6009[2]	6662[2]	>5000[1]	6493[2]	>5000[1]
NO_x, ppm	281.6	239.8	280.1	394.4	394.5	404	554
Spray Dryer Exit							
O_2, %	NM	NM	NM	NM	NM	3.90	NM
CO_2, %	NM	NM	NM	NM	NM	15.0	NM
CO, ppm	NM	NM	NM	NM	NM	62.0	NM
SO_2, ppm	NM	NM	NM	NM	NM	3975	NM
NO_x, ppm	NM	NM	NM	NM	NM	409	NM

[1] The data logger for analyzer Mobile 1 does not record values over 5000 ppm for SO_2.
[2] Data taken from logbook.
[3] Not measured.

tests. This test ran without major incident, and Test 3B was started 17:50. This test also ran well, and Test 3C was then started at 18:30. No major operational problems were noted during this test. It was noted that NO_x was behaving opposite of what is known to be true at the half-load conditions. Because of this, the operators did not feel the data being taken were valid. Additionally, half load is not simulated well in the CTF because it involves reducing the flow to the one and only burner. In a full-scale system, half load is accomplished by reducing the number of burners firing but still keeping the firing burners at full capacity. Because of these issues, Tests 4A and 4B were abandoned. Test 4C, was run for a brief period (19:00 to 19:20) and then shut down. Upon shutting down the test, it was discovered that the stack was plugged with $CaSO_4$ deposits, so the system was shut down to clean the stack piping.

Test 5 was a longer-term test intended to look at the fouling potential of the coal. It was started at 13:50 on January 26. This lower-temperature test (1051°C) ran well, without major incident reported through the test period. The test ended at 17:30 because of ash deposition in the duct. The system was shut down to evaluate the fouling probes. Upon removal of the probes, very little ash deposit was found, and it was friable. The majority of the ash deposit formed just ahead of the probe bank.

Test 6 was started on February 1 at 9:30. This higher-temperature test (1116°C) was intended to evaluate the upper limit of furnace exit temperature. Almost immediate signs of duct plugging were noted in the system. Ten minutes into the test, the coal feed was reduced by 3 lb/hr and then slowly reduced by another 12 lb/hr over the next 20 minutes. At 10:09, the test was ended because of severe plugging in the system. Removal of the fouling probes indicated severe ash buildup.

TEST RESULTS

Gas Analysis Summary

Table 5 also summarizes the dry flue gas concentration for each test in three locations—the furnace exit, the baghouse exit, and the spray dryer exit. The concentrations listed are on an as-measured O_2 basis; the concentrations are reported on a 6% O_2 basis in Appendix B. The furnace exit was continuously monitored with analyzer Mobile 2. Analyzer Mobile 1 was used to monitor both the baghouse exit and spray dryer exit. The analyzer was alternated between flue gas streams as needed. During several of the runs, the flue gas from either the baghouse exit or spray dryer exit was only spot-checked with the analyzer; therefore, there were not enough steady-state data gathered to be able to report meaningful values. These runs are denoted with "NM."

Multiple spikes in the CO readings are the cause of some of the anomalous CO levels in the table. The spikes were most likely caused by fluctuations in the fuel feed. O_2 levels ranged from 2.3% to 4.7% at the furnace exit. O_2 concentration at the baghouse outlet was typically 0%–1% higher than the furnace exit, and O_2 concentration at the spray dryer exit was 0.5%–2% higher than the furnace exit. CO_2 levels were typically steady throughout the test periods, ranging from 13.9% to 17.0%.

The concentration of SO_2 at the furnace exit was extremely high, in many cases approaching or exceeding 7000 ppm. These concentrations are indicative of the high sulfur concentration in the coal. The high sulfur concentrations may contribute to significant ash-fouling problems and are certainly a cause for concern with environmental regulations. Flue gas analysis of NO_x shows concentrations ranging from approximately 260 to 475 ppm at the furnace exit. For Test 2, NO_x levels appear to be reduced with reduced burner stoichiometry, even though temperature increased during Test 2C.

Figures 1 and 2 compare the NO_x levels at the furnace exit to flue gas temperature and burner stoichiometry. Temperature is shown to have some impact on NO_x levels as expected; however, the overriding effect of burner stoichiometry results in an outlier on the chart. There is a definite upward trend in NO_x with increasing burner stoichiometry.

The effect of each of the test variables on the various gas components was determined using MINITAB statistical software. In general, very few of the variables were shown to have a significant impact on the gas concentrations. The most noteworthy finding from the statistical analysis was that the O_2 and CO_2 levels varied significantly with load. This indicates that, on average, the air-to-coal ratio was slightly higher for the half-load tests than the full-load tests.

Swirl Effect and Flame Stability

The swirl number is a dimensionless quantity related to the amount of “spin” imparted to the secondary airstream. It is the ratio of the tangential momentum to the radial momentum of the airstream. Preliminary observations of flame stability indicate that stable combustion could be attained under full-load conditions at a secondary air swirl setting similar to most low-rank fuels fired in the CTF. A swirl setting of 3.2 was used for all full-load tests. This corresponds to a swirl number of 0.418. Initially, at half-load conditions, the maximum swirl setting of 0 was used to help maintain a stable flame. This corresponds to a swirl number of 1.9. However, the setting was quickly changed back to 3.2 as this was found to provide an adequately stable flame.

Unburned Carbon and Combustion Efficiency

Figure 3 shows the carbon concentration in the baghouse ash for each of the runs. The 0% stoichiometry setting results in the greatest amount of unburned carbon for the first three test periods. Aside from that, the variability in the data makes direct interpretation difficult. Therefore, the data were statistically analyzed to determine if any of the main test variables had a statistically significant impact on unburned carbon. The analysis was again performed using MINITAB statistical software. Although none of the variables had a statistically significant effect on baghouse ash carbon content, it did appear that excess air had a practically significant impact. On average, increasing excess air from 15% to 20% reduced the carbon in the baghouse ash by 0.14%.

Carbon concentration in the baghouse ash was also used to determine combustion efficiency for Runs 1, 2, and 3. The combustion efficiencies are reported in Table 5. The combustion efficiency was calculated based on the amount of unburned carbon in the baghouse ash compared to the amount of carbon fed during each test. For all tests, combustion efficiency was 99.5% or greater. A conversion of 99.9% was achieved in Test 2C.

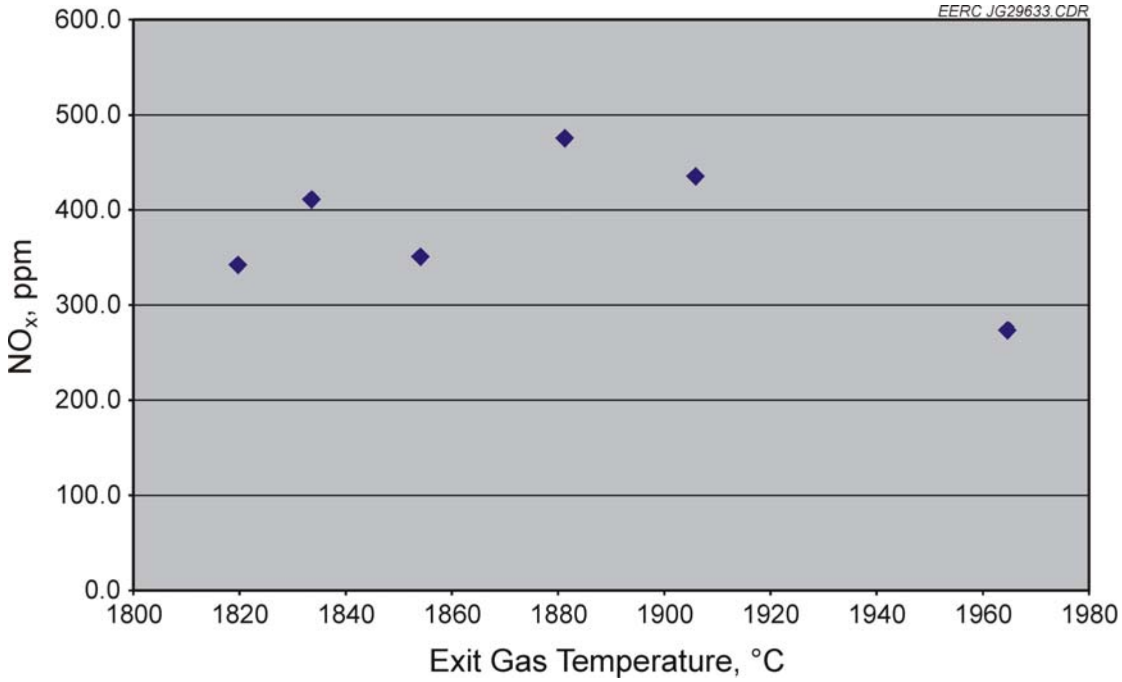

Figure 1. NO_x furnace exit combustion temperature for full-load parametric tests.

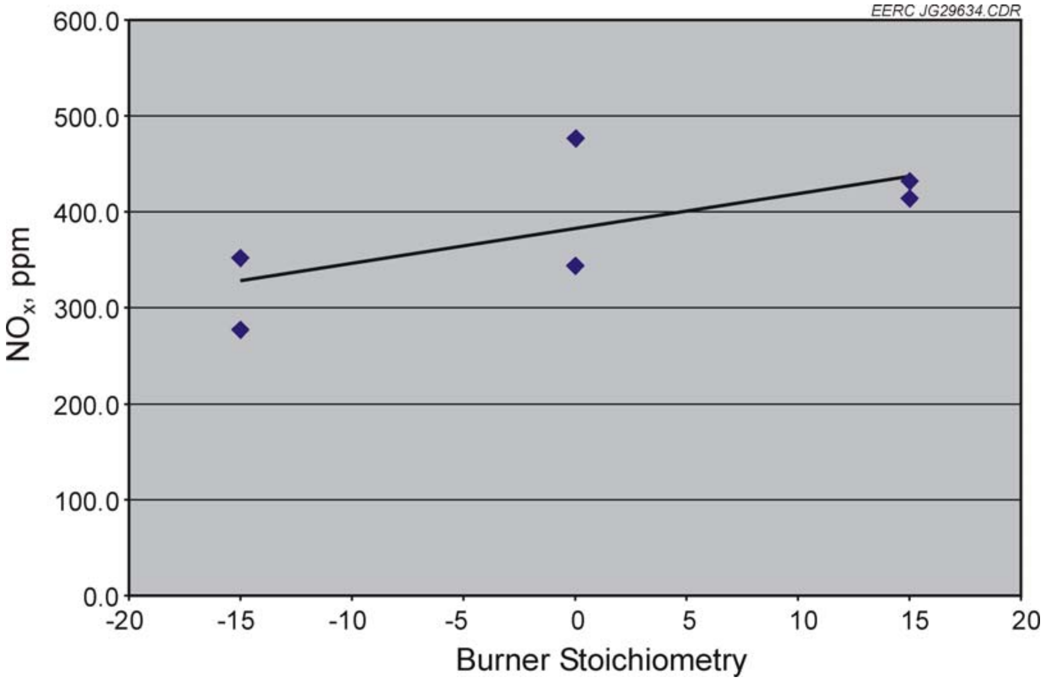

Figure 2. NO_x burner stoichiometry for full-load parametric tests.

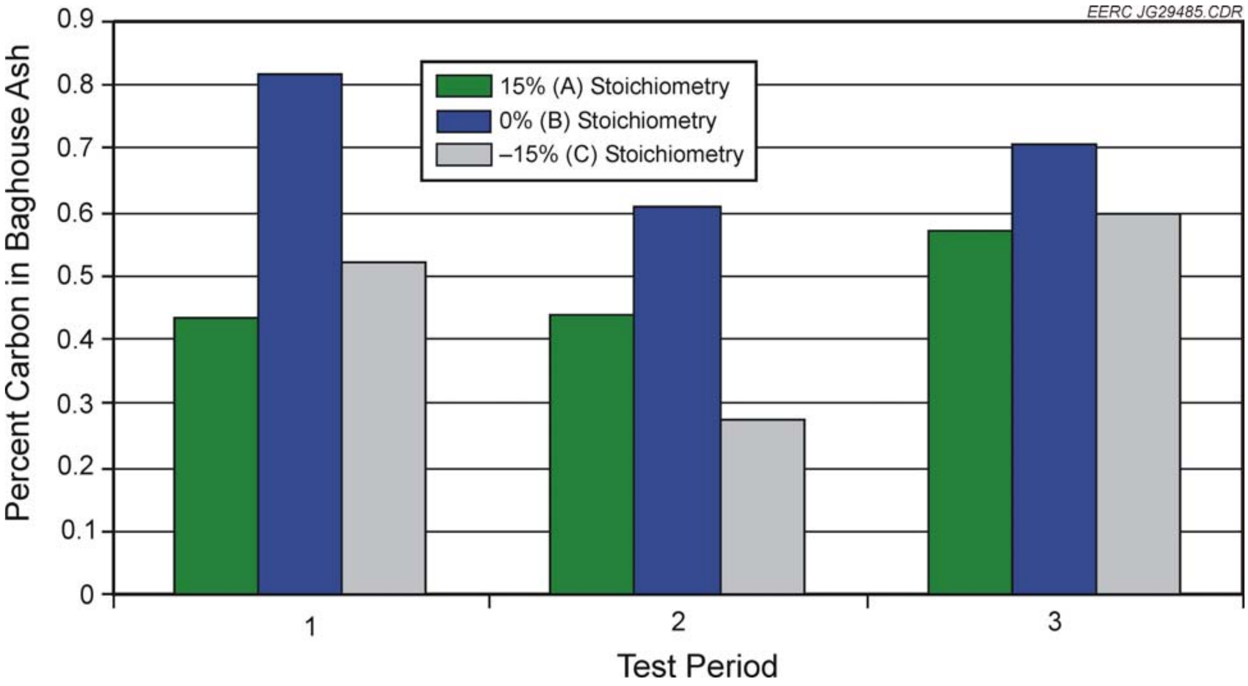

Figure 3. Effect of excess air and burner stoichiometry on unburned carbon.

Method 26

In addition to the NO_x evaluation, a Method 26 test was conducted to look for the halogen concentration of the fuel. A summary of the extractive sampling operating conditions and Method 26 test is presented in Tables 6 and 7, respectively. Table 7 shows that there was a significant amount of fluorine, 152.8 mg/m^3, in the flue gas, which could potentially cause significant corrosion in the cooler regions of the boiler and the ash collection equipment. Chlorine and bromine levels were both relatively low.

Ash-Fouling Tests

Two ash-fouling tests were run to determine the level of ash buildup using fouling probes to simulate ash buildup on boiler tubes. One lower-temperature test (Test 5) and one higher-temperature test (Test 6) were completed. The fouling probes were made of 304 stainless steel and a Cr–Mo composite consisting of 0.07% C, 0.25% Si, 0.48% Mn, 0.017% P, 0.002% S, 2.09% Cr, 0.92% Mo, 0.004% Sn, and 96.167% Fe.

Based upon the results from Test 5, the FEGT was increased from 1050° to about 1120°C. In this case, the deposit was initiated at the tube metal surface, quickly bridging between tubes and blocking the duct after only 40 minutes of operation. The deposit was highly sintered and appeared to present a significantly more difficult problem for sootblowers. The deposit was submitted for lab analysis, which is shown in Tables 8–10. A photo of the deposit formed on the tubes during Test 5 is provided in Figure 4, and the Test 6 deposit is shown in Figure 5.

Table 6. Extractive Sampling Operating Conditions

Test Number:	AT-CTS-824
Fuel Description:	Tufanbeyli Lignite
Date:	1/24/2007
Sampling Method	Method 26
Sampling Location	Baghouse outlet
Start Time	16:32
Test Duration, min	60
Stack Temperature, °F (°C)	336 (169)
Sample Volume, ft^3 (m^3)	34.42 (0.97)
Flue Gas Moisture, %	26.0%
% Isokinetic	118.6
Dust Loading, grains/scf	14.19

Table 7. Method 26 Results

Date: 1/24/2007 Test : AF-CTS-824-M26A-1	Probe Rinse	H_2SO_4	NaOH
Analysis Volume, mL	250	500	500
Concentration, mg/L			
Chlorine	3.7	34	N/A
Fluorine	4.8	178	35.0
Bromine	N/A	6.5	N/A
Total Chlorine, mg		18.43	
Total Fluorine, mg		107.70	
Total Bromine, mg		3.25	
Sample Volume, m^3 at 20°C, 1 atm, dry		0.721	
Sample Volume, m^3 at 20°C, 1 atm, wet		0.975	
Flue Gas Concentration, mg/m^3 (dry)			
Chloride		25.5	
Fluoride		149.3	
Bromide		4.5	
Flue Gas Oxygen, %		3.41	
Flue Gas Concentration Corrected at 6% O_2, mg/m^3 (dry)			
HCl		22.4	
HF		134.0	
HBr		3.9	

A standard fouling test is 5.25 hours long, and both tests plugged off the duct before that time was achieved. Based on these results, it is recommended that the FEGT be controlled to between 1050° and 1070°C, limiting any excursions above these temperatures to maintain manageable sootblower operations.

Table 8. Summary of Ash-Fouling Probe Test Results

	AF-CTS-825 1/26/2007 Test 5		AF-CTS-826 2/1/2007 Test 6	
FEGT, °F (°C)	1923 (1050)		2040 (1115)	
Probe Metal Temp., °F (°C)	1010 (543)		1020 (549)	
Length of Run, hr	3.7		0.65	
Fuel Feed Rate, lb/hr (kg/hr)	142 (64.4)		147 (66.7)	
Ash Input Rate, lb/hr (kg/hr)	56 (25.4)		57 (25.9)	
Deposit Weight, g	966		1025	
Deposition Rate, g/hr	261		1577	
Bulk Analysis	As-Received	BaO-, SO_3-Free	As-Received	BaO-, SO_3-Free
SiO_2	50.0	50.1	50.1	50.2
Al_2O_3	23.3	23.3	22.3	22.3
Fe_2O_3	9.68	9.69	10.92	10.93
TiO_2	1.09	1.09	1.07	1.07
P_2O_5	0.42	0.42	0.17	0.17
CaO	11.2	11.2	11.0	11.0
MgO	1.63	1.63	1.6	1.69
Na_2O	0.39	0.39	0.30	0.30
K_2O	2.13	2.13	2.37	2.37
SO_3	0.00	N/A	0.00	N/A
BaO	0.12	N/A	0.05	N/A
SrO	0.02	N/A	0.03	N/A

The bulk chemical analysis of the deposits is shown in Table 8. It is very similar in composition: high in silicate 50%, aluminum ~22%, iron ~10%, and calcium ~11%. The presence of the aluminum and calcium indicates that there were clays present in the coal. Silicate is more of an inert material and is not as reactive as phosphorus or sulfur; however, it can become problematic if there is a significant amount of silicate. The Tufanbeyli lignite had a very high ash content, resulting in significant fouling during both tests; however, there was significantly more fouling during the Test 6 higher-temperature test.

Table 9 shows the SEMPC (scanning electron microscope point count) analysis for the deposits. The deposits are very similar in composition, primarily silicon-rich, approximately 95.6% in the lower-temperature test and 89.2% in the higher-temperature test. Quartz, anorthite, illite, and mixed-silicon rich were the main constituents of the silicon-rich points. There was a slight variation between the samples; however, it is relatively small. Montmorillonite was present in the higher-temperature deposit (9.2%) and not present in the lower-temperature deposit, which was probably due to the effect of temperature and the formation of crystals. Also present was iron oxide at 1.2% and 2.0% for Tests 5 and 6, respectively. There was also a slight variability of the iron and mixed-metal-rich group from Tests 6 to 5, likely due to variability in the coal. The presence of sulfur in the Test 6 deposit and no sulfur present in the Test 5 deposit point count appear to be significant; however, this lack of sulfur in the Test 5 deposit point count is likely

Table 9. Summary of SEMPC Analyses

SEMPC	AF-CTS-825 1/26/2007 Test 5	AF-CTS-826 2/1/2007 Test 6
Oxide-Rich		
Titanium Oxide	0.4	0.0
Iron Oxide	0.8	2.0
Total for Group	1.2	2.0
Phosphorus-Rich		
Apatite	0.4	0.4
Metal-Rich		
Iron	0.0	1.2
Mixed-Metal-Rich	0.8	3.6
Total for Group	0.8	4.8
Silicon-Rich		
Quartz	2.0	2.8
Albite	0.0	0.4
Anorthite	49.6	33.2
Potassium Feldspar	0.4	0.0
Illite	2.8	4.0
Montmorillonite	0.0	9.2
Wollastonite	0.8	0.4
CaSilicate	0.4	0.0
Mixed-Silicon-Rich	28.0	39.2
Total for Group	95.6	89.2
Sulfur-Rich		
Pyrite	0.0	0.4
Mixed-Sulfur-Rich	0.0	0.4
Total for Group	0.0	0.8
Carbon-Rich		
Mixed-Carbon-Rich	0.0	1.6
Other	2.0	1.2

Table 10. Major Crystalline Phases

AF-CTS-825 1/26/2007 Test 5		AF-CTS-826 2/1/2007 Test 6	
Major Phases	Nominal Composition(s)	Major Phases	Nominal Composition(s)
Quartz	SiO_2	Quartz	SiO_2
Anorthite Intermediate	$(Ca,Na)(Si,Al)_4O_8$	Anorthite intermediate	$(Ca,Na)(Si,Al)_4O_8$
Hematite	Fe_2O_3	Hematite	Fe_2O_3

EERC JG29483.CDR

Figure 4. Probe bank at end of Test 5, 1050°C.

because of variability in the coal and a slight variability in the SEMPC method. Overall, SEMPC indicates that the deposits were silicon primarily composed of anorthite and mixed-silicon-rich crystals, indicating the presence of clays in the fuel.

Table 10 shows the major crystalline phases of the deposits determined by SEM morphology. The major phases present in both sample were quartz, anorthite intermediate, and hematite. This result also indicates that the deposits were formed from a fuel containing clay minerals.

Shown in Table 11 is the bulk chemical analysis of the baghouse ash samples collected during the formation of the deposits in Tests 5 and 6. The baghouse ash samples were almost identical to the deposit compositions and to each other. The baghouse ash had sulfur present in both samples, and the bulk chemical analysis for the deposits indicated no sulfur present; however, SEMPC indicated the presence of sulfur. The amount of sulfur was low, less than 1.5%; thus the relative fouling potential for the sulfur is low; however, sulfur tends to build up in systems over time and could potentially be problematic.

Contrary to results obtained in Tests 5 and 6, the ash fusion analysis (Table 12) of the coal indicated very high temperatures, ranging from an initial deformation temperature of 1302°C to a fluid temperature of 1368°C. This coal sample was taken from Test 6. These temperatures are well above those utilized in Tests 5 and 6. Because the fly ash grains become segregated and undergo transformation in the flame and in the flue gas, it is not surprising that the ash fusion temperatures reported here do not reflect the nature of the deposits formed during testing.

EERC JG29481.CDR

Figure 5. Probe bank deposit formed in Test 6, at 1120°C.

Table 11. Bulk Chemical Analysis of Baghouse Ash Samples

	AF-CTS-825 Test 5 1/26/2007		AF-CTS-826 Test 6 2/1/2007	
Bulk Analysis	As-Received	ClO-, BaO-, SO_3-Free	As-Received	ClO-, BaO-, SO_3-Free
SiO_2	48.8	49.4	49.1	49.6
Al_2O_3	24.1	24.4	23.3	23.6
Fe_2O_3	10.17	10.29	10.36	10.48
TiO_2	1.07	1.08	1.03	1.04
P_2O_5	0.42	0.43	0.43	0.43
CaO	9.8	10.0	10.2	10.3
MgO	1.83	1.85	1.80	1.82
Na_2O	0.42	0.43	0.42	0.43
K_2O	2.16	2.19	2.29	2.32
SO_3	1.24	N/A	1.09	N/A
BaO	0.11	N/A	0.14	N/A
SrO	0.2	0.2	0.02	0.2

Table 12. Ash Fusion Analysis

Oxidizing Atmosphere	°F	°C
Initial Deformation	2376	1302
Softening Temperature	2398	1314
Hemispherical Temperature	2466	1352
Fluid Temperature	2494	1368

During each fouling test, a water-cooled probe was inserted through the wall of the furnace to capture radiant zone deposits. The surface metal was cooled to approximately 350°C during each test. A photo of the slag probe from Test 5 is provided in Figure 6. In both tests, only a dusting formed at the surface of the probe, indicating a low potential for ash deposition on the cooler walls in the radiant zone of the boiler.

FACT Analysis

The FACT (Facility for the Analysis of Chemical Thermodynamics) viscosity of the coal is shown in Figure 7. The viscosity of Tests 5 and 6 are shown on the figure, indicated by circles. As the temperature increases from 1051° to 1116°C, the viscosity decreases dramatically. The T_{250} line is where the material flows as a liquid. During the testing, that temperature was never reached; however, the increase in viscosity with the large ash generated by this coal caused significant fouling.

EERC JG29509.CDR

Figure 6. Slag probe utilized in Tests 5 and 6.

Figure 8 shows the percentage of sulfur in solids as the temperature increases. It is important to note that at the lowest temperature, all of the sulfur has been converted out of the solid phase and is likely in the gas phase. When flue gas becomes cool enough, the sulfur will precipitate out, and it will be very problematic because of its binding properties and the high ash content of the coal. The result was probably one of the most significant reasons for the fouling of the system.

The SCMO (solid calcium magnesium oxide) solution and the Slag C are shown in Figures 9 and 10, respectively. The figures correlate well with interaction going on in the system. As shown in Figure 9, as the temperature increases, there is less and less of the SCMO, meaning that it melted and became a liquid and or gas. The SCMO started to melt around 800°C and was completely melted by approximately 930°C, well below the Test 5 fouling test. Slag C starts to appear at approximately 900°C and steadily increases. When the temperature is increased from approximately 1040° to 1116°C, the percentage of Slag C increases from approximately 4% to 10%. This result was also another factor that contributed to the significant fouling that occurred during Test 6.

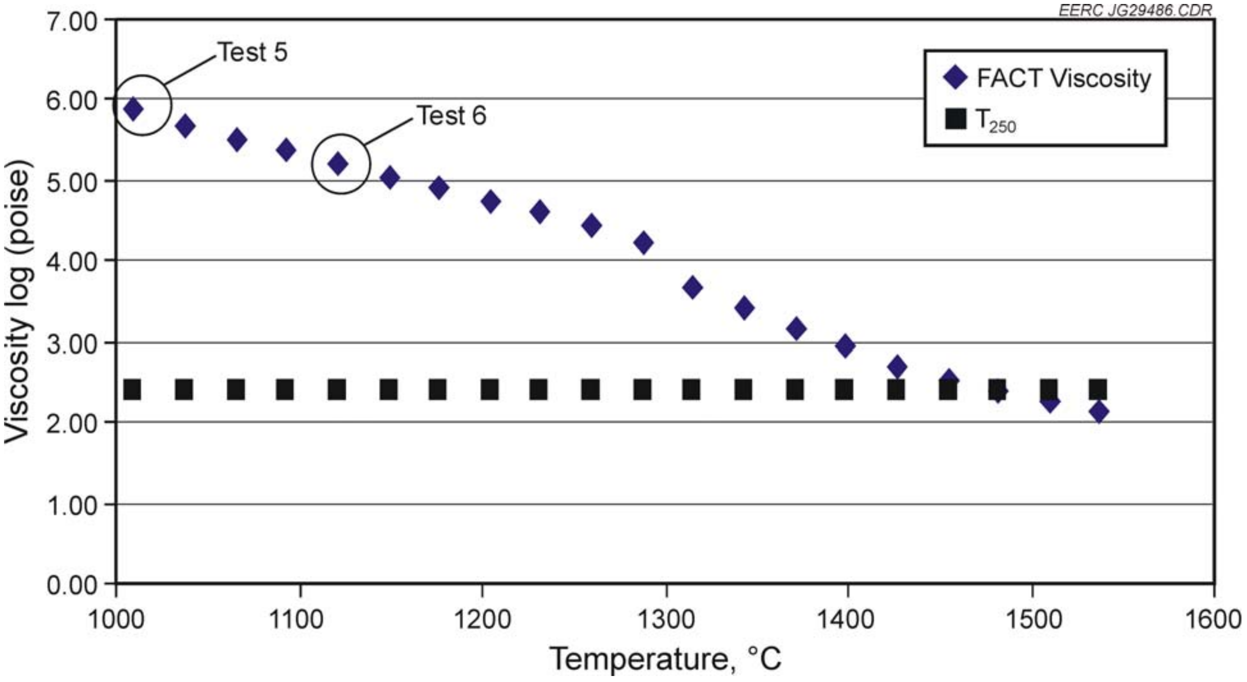

Figure 7. Bulk viscosity of the Tufanbeyli coal.

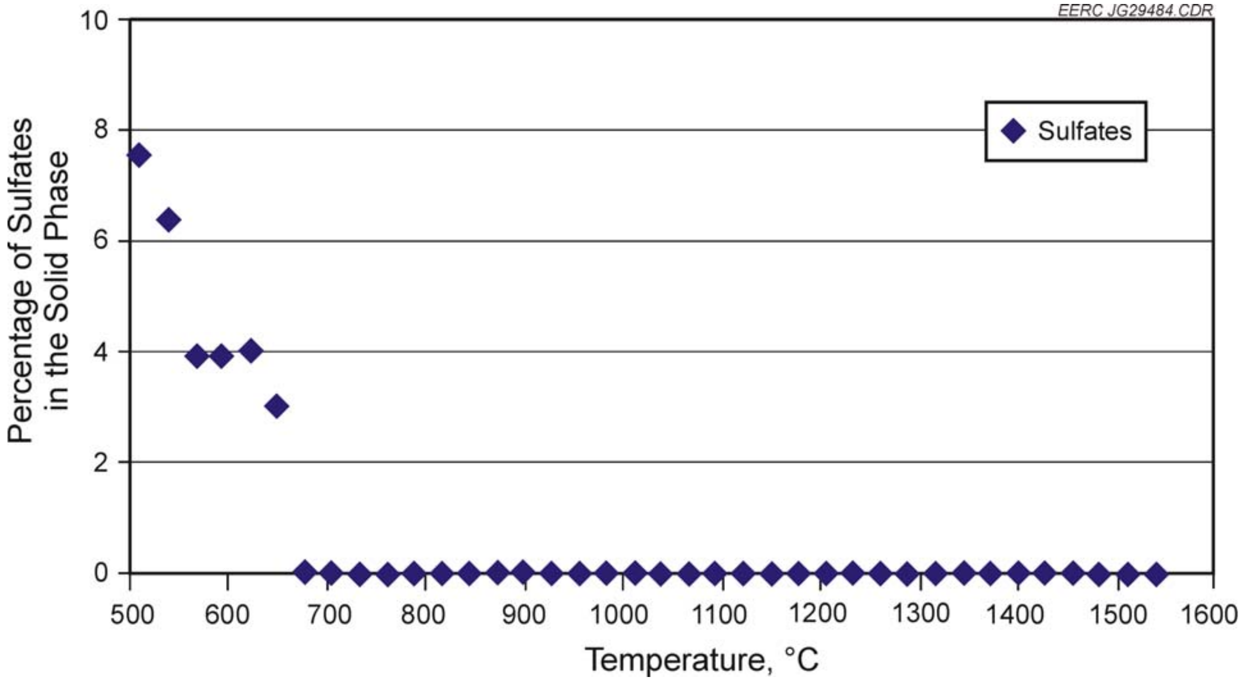

Figure 8. Percentage of sulfur in the solid.

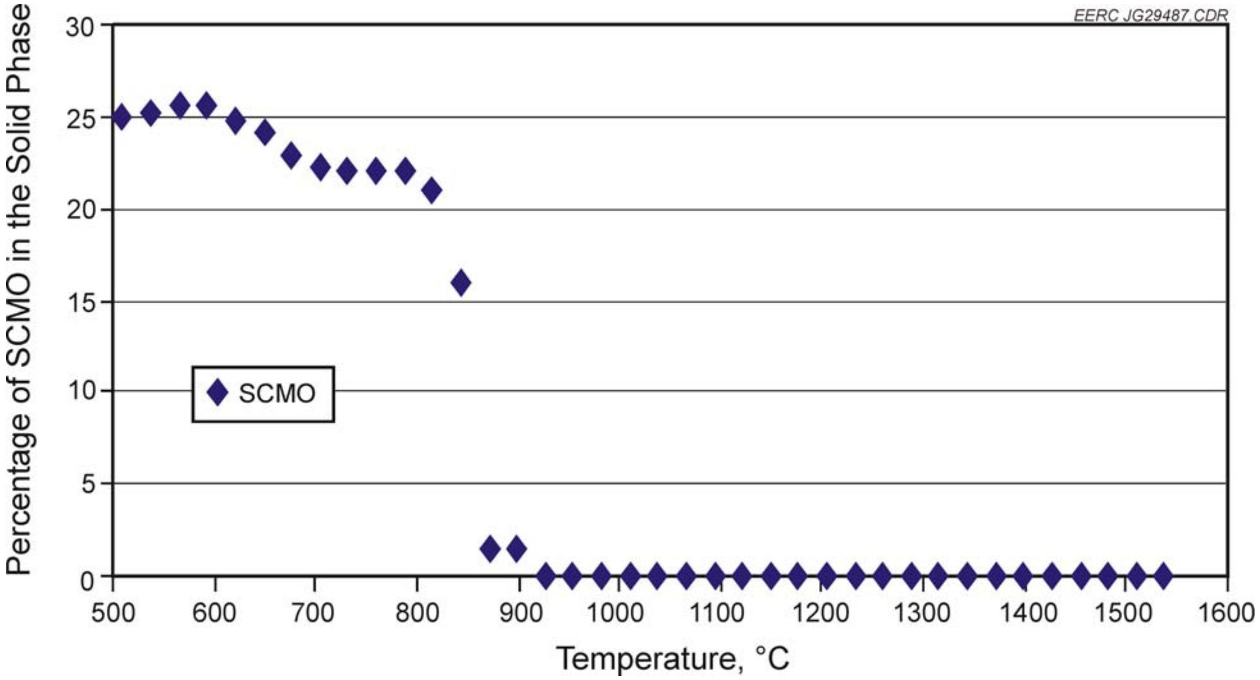

Figure 9. Percentage of SCMO in the solid phase.

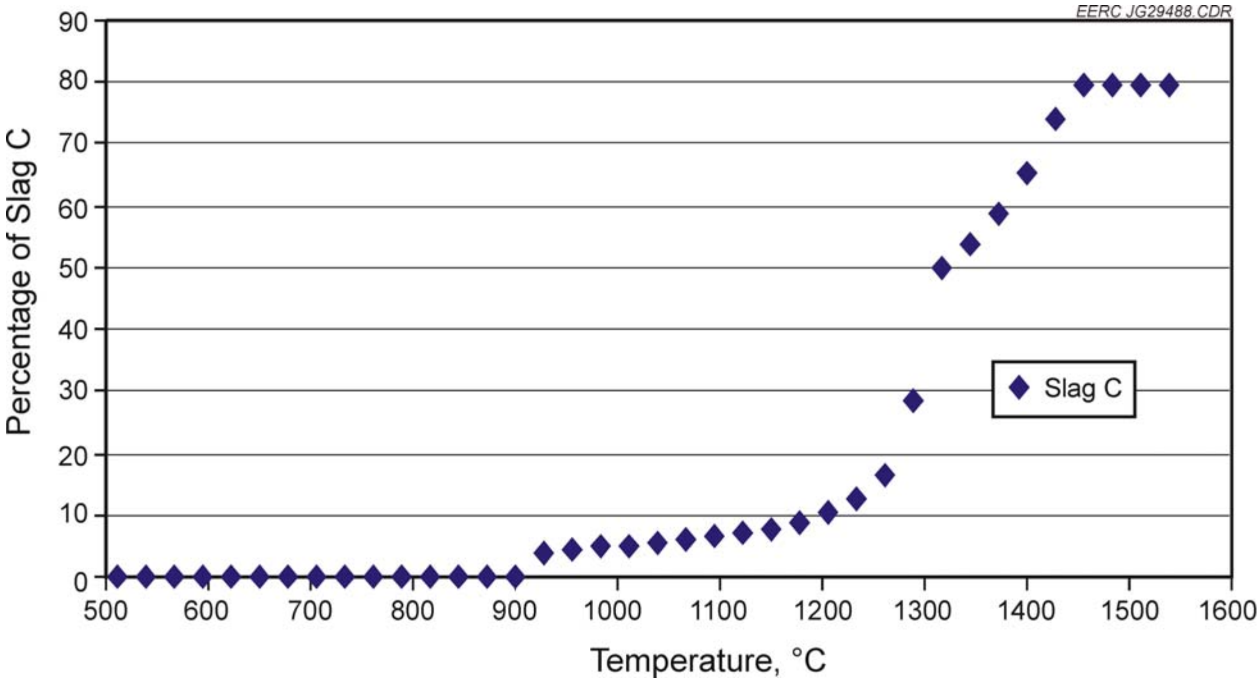

Figure 10. Percentage of Slag C.

The viscosity analyses were done for each deposit; however, because of the similarity between the two deposits in composition, the FACT analyses were identical. The baghouse ash samples were also analyzed using FACT analysis; the results were almost identical and did not differ greatly from the FACT of the coal and deposits.

CIRCULATING FLUIDIZED-BED TESTING

Proposed Test Matrix

A matrix of test parameters was presented to the EERC as a basis for the proposed testing and is shown as Table 13.

Tests 1–3 were proposed to measure inherent sulfur capture. Test 1 was to be a baseline condition (850°C) with no limestone addition. Test 2 was to be a low-temperature condition (820°C) with no limestone addition. Test 3 was a low-load test with no limestone addition.

Tests 4 through 8 were for determining sulfur capture rates required over a range of test conditions. Test 4 was baseline testing (850°C) with limestone addition to decrease the SO_2 concentration down to 148 ppm (equivalent to 400 mg/Nm^3 at 6.0% O_2). Assuming 30% utilization of the coal ash and limestone calcium, it was projected that a limestone feed rate of 42 lb/hr would be required. For lower utilization rates, higher limestone feed rates are required. Method 26 sampling for determination of bromine, chlorine, and fluorine flue gas concentrations at the inlet and outlet of the baghouse was scheduled during Test Period 4. Test 5 was a low-temperature test (820°C) with limestone addition to decrease the SO_2 concentration down to 148 ppm (equivalent to 400 mg/Nm^3 at 6.0% O_2). Test 6 was a low-temperature test (820°C) at a high sulfur capture rate with limestone addition to decrease the SO_2 concentration down to 72 ppm (equivalent to 200 mg/Nm^3 at 6.0% O_2). Test 7 was a baseline temperature test (850°C) with limestone addition to decrease the SO_2 concentration down to 72 ppm (equivalent to 200 mg/Nm^3 at 6.0% O_2). Test 8 was a high-temperature test (880°C) with limestone addition to decrease the SO_2 concentration down to 72 ppm (equivalent to 200 mg/Nm^3 at 6.0% O_2). Performance of Test 8 was dependent upon the success of Test 7, as the project sponsor directed.

Tests 9 and 10 were low-load tests with limestone addition to decrease the SO_2 concentration down to 148 ppm (equivalent to 400 mg/Nm^3 at 6.0% O_2) and 72 ppm (equivalent to 200 mg/Nm^3 at 6.0% O_2), respectively. Performance of Test 10 was dependent upon the success of Test 9, as the project sponsor directed.

The actual test matrix was modified during the course of the week of operation, with approval of the project sponsors. Tests 2 and 8 were eliminated from the matrix, and the order of testing was changed. The following tables of results will show the tests in the order in which they were performed, but test number designations will be the same as in the test matrix in Table 13.

Table 13. Proposed Specified Test Matrix

Test No.	Av Bed Temp., °C	Fluidization Velocity, m/sec	Excess Air, %	Limestone Addition	Target SO_2, mg/Nm^3	Remarks
1	850	5	20	No	TBD[1]	Inherent sulfur capture
2	820	5	20	No	TBD	Inherent sulfur capture
3	TBD	2.5	20	No	TBD	Low load, inherent sulfur capture
4	850	5	20	Yes	400	Moderate sulfur capture
5	820	5	22	Yes	400	Moderate sulfur capture
6	820	5	20	Yes	200	High sulfur capture
7	850	5	20	Yes	200	High sulfur capture
8	880	5	20	Yes	200	High sulfur capture
9	TBD	2.5	20	Yes	400	Low load, moderate sulfur capture
10	TBD	2.5	20	Yes	200	Low load, high sulfur capture

[1] To be determined.

Notes: If Test 7 fails then no need to perform Test 8.
If Test 9 fails then no need to perform Test 10.
All tests will be carried out at constant primary air/secondary air (PA/SA) ratio unless otherwise found necessary.

COAL AND LIMESTONE PROPERTIES

The coal and limestone used for this combustion test was supplied from near the city of Adana in Turkey. The coal arrived in seventy-four 1-ton supersacks, which required crushing to −⅜ inches to achieve the desired size distribution. The EERC was told that all of the coal was from the same mine, but it was apparent that there were two distinctly different coal samples in the supersacks. The difference is possibly because some of the coal was extracted from a lower location in the mine where the water table was still high, resulting in extremely high moisture content for the coal in these sacks. Since this was a new unopened portion of the mine, it had not yet been drained, as is the normal procedure upon opening, resulting in a lower-moisture coal for feed to the proposed power plant. In order to be able to handle this coal through the EERC crusher and classifier and eventually feed the coal to the circulating fluidized bed, it was laid out on the floor to air-dry prior to processing. It was attempted to minimally dry the coal so it would be more representative of what would actually be fed at the proposed new power plant. The sized coal was transferred to storage hoppers having a net capacity of about 3500 lb (1590 kg).

The limestone arrived in six of the 1-ton supersacks shipped with the coal. It was initially crushed to−1 inch by a local contractor. It was then crushed at the EERC to −20 mesh and placed in 55-gallon drums.

The size distributions of the coal, limestone, and sand used as start-up bed material are shown in Figure 11. The size distribution for the lignite shown in Figure 11 indicated a wider variance in the lignite than was likely the case. Overall, there were relatively few larger coal particles present in the processed lignite, so it was difficult to obtain a representative sample. The top size was less than 9.5 mm, with a d_{50} of approximately 1.5 mm. The limestone had a maximum size less than 850 µm, with d_{50}s ranging from 300 to 575 µm for the three samples shown in Figure 11. Based upon the preliminary size analysis performed for this limestone, it is more likely that the size distribution for the limestone was closer to the d_{50} of 300 µm shown for Limestone 3 for this testing than the larger d_{50}s determined for limestone Samples 1 and 2.

Proximate and ultimate analyses of the coal and x-ray fluorescence analysis (XRFA) of the coal ash were performed, and the results are shown in Table 14. The limestone analyses are shown in Table 15.

Coal and Limestone Feed System Modifications

A schematic of the EERC CFBC system showing the configuration used for this test is shown in Figure 12. Appendix C describes the EERC CFBC system in more detail.

The coal and limestone feed system consists of several components. Removable coal storage hoppers are used to transport the prepared coal to the test facility. These are hoisted up in place to the sixth floor of the facility. A pneumatically actuated 12.25-cm gate valve is used to control the amount of coal transferred from the coal storage hopper to the coal feed hopper. The coal feed hopper is suspended from a load cell to measure the total amount of coal in it. A 6-inch coal rotary feed control valve is located at the bottom of the coal feed hopper, and a manual speed controller is used to control the rotation speed of this valve. The coal then drops into another 6-inch rotary valve identified as the coal/limestone rotary seal valve. This valve's function is to help isolate the upper portion of the coal feed system from the pressure in the lower portion of the coal feed system where the coal is actually fed into the combustor. Beneath the coal/limestone rotary seal valve, the coal drops into a 10-cm auger that conveys the coal approximately 2.75 meters over to where the coal finally drops into a 7.6-cm line that enters into the combustor. A pneumatic assist is used at this location to help the coal feed into the combustor. The coal enters at a 60 degree downward angle into the combustor 2 meters above the distributor plate.

The limestone system used a screw feeder to control the limestone feed rate. A scale was used to weigh all limestone added to the feeder to determine the actual feed rate. The screw feeder had a manual speed control, and weights were taken periodically to determine the actual feed rate. The screw feeder dropped into a 10-cm rotary valve that helps to seal the auger from any back pressure in the system. The rotational speed of the rotary valve used here had to be increased to handle a higher limestone feed rate than is typically used for the EERC CFBC system. At the exit of this rotary valve, the limestone drops down an inclined 10-cm pipe just above the coal/limestone rotary seal valve where it mixes in with the coal feed. As described above, the coal–limestone mixture drops through the coal/limestone rotary seal valve and continues through the 4-inch auger and into the combustor.

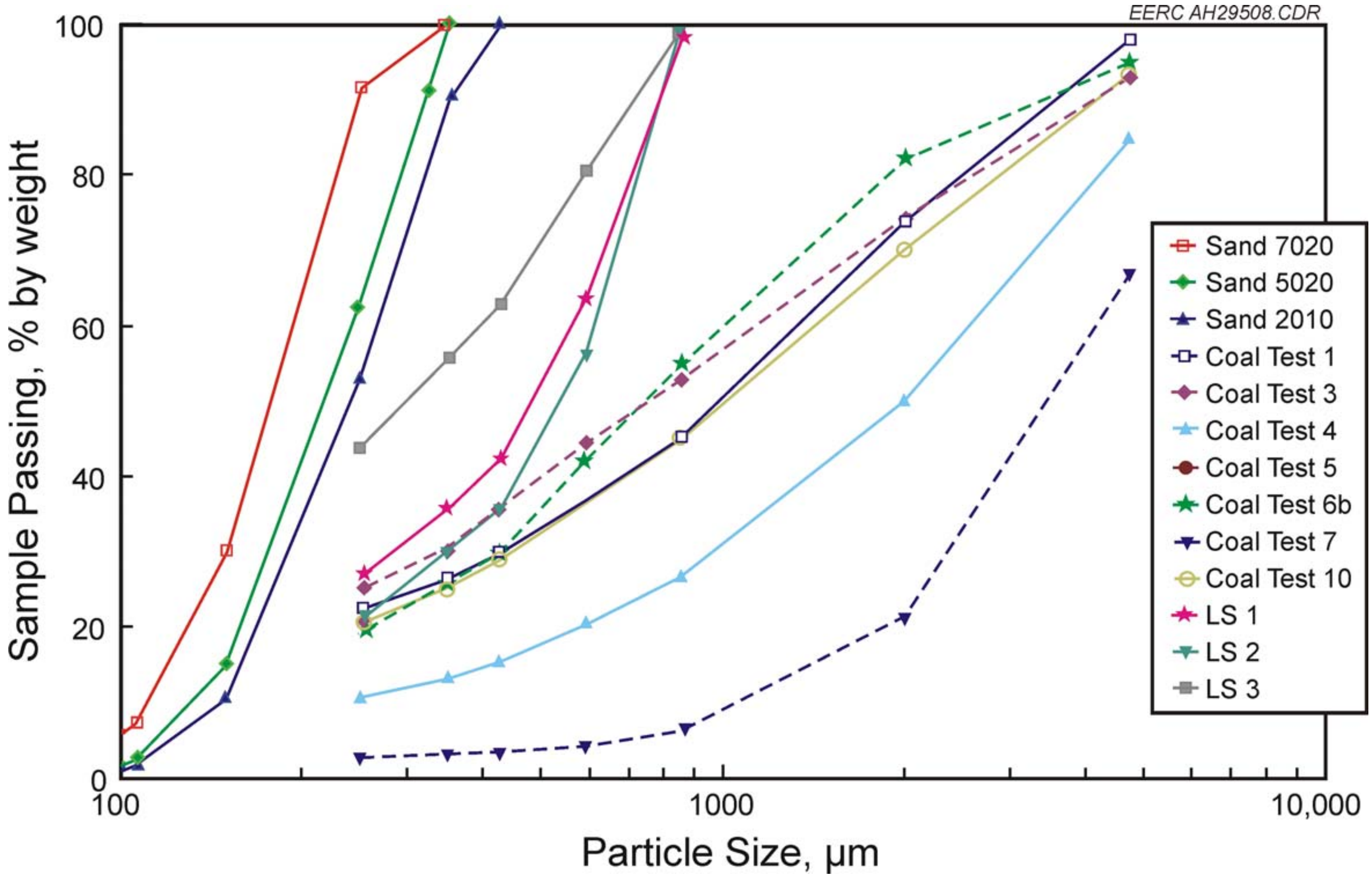

Figure 11. Size distributions for bed material, coal, and limestone.

Table 14. Coal Analyses

Fuel Analyses	Test 1	Test 3	Test 6 b	Test 5	Test 4	Test 7	Test 6	Test 9	Test 10
Proximate Analysis, as received, wt%									
Moisture	39.20	41.80	41.70	37.50	40.10	40.60	42.30	42.30	35.50
Volatile Matter	22.06	22.53	21.58	22.25	21.45	21.02	21.08	21.08	22.86
Fixed Carbon	13.57	11.76	12.10	11.49	10.16	10.53	10.88	10.88	11.72
Ash	25.17	23.91	24.62	28.76	28.29	27.85	25.74	25.74	29.92
Ultimate Analysis, as received, wt%									
Carbon	21.53	21.21	20.69	20.12	18.98	20.16	19.46	19.46	21.34
Hydrogen	1.87	1.84	1.83	1.78	1.68	1.79	1.73	1.73	1.91
Nitrogen	0.66	0.70	0.69	0.70	0.66	0.68	0.65	0.65	0.70
Sulfur	2.35	2.35	2.14	2.12	2.10	2.20	2.14	2.14	2.46
Oxygen	9.22	8.19	8.33	9.02	8.18	6.73	7.98	7.98	8.17
Ash	25.17	23.91	24.62	28.76	28.29	27.85	25.74	25.74	29.92
Moisture	39.20	41.80	41.70	37.50	40.10	40.60	42.30	42.30	35.50
Ash Composition, % as oxides									
Calcium, CaO	10.00	11.70	11.10	10.40	11.10	11.10	10.10	10.10	9.90
Magnesium, MgO	1.57	1.53	1.64	1.58	1.53	1.57	1.62	1.62	1.59
Sodium, Na_2O	0.25	0.22	0.21	0.19	0.18	0.13	0.21	0.21	0.18
Silica, SiO_2	42.10	40.80	40.30	43.20	41.50	40.90	42.80	42.80	41.60
Aluminum, Al_2O_3	19.80	18.90	19.10	19.80	18.90	18.90	20.20	20.20	19.80
Ferric, Fe_2O_3	9.74	10.04	9.38	9.29	9.70	9.15	9.42	9.42	9.25
Titanium, TiO_2	0.92	0.93	0.93	0.95	0.96	0.98	0.97	0.97	0.94
Phosphorus, P_2O_5	0.18	0.19	0.18	0.16	0.17	0.17	0.17	0.17	0.17
Potassium, K_2O	1.93	1.94	1.97	2.13	1.98	1.97	2.04	2.04	1.95
Sulfur, SO_3	13.41	13.66	15.16	12.34	13.97	15.11	12.47	12.47	14.53
High Heating Value									
Moisture-free, MJ/kg	13.76	13.76	13.42	11.88	11.63	11.64	12.76	12.76	12.10
As-received, MJ/kg	8.36	8.01	7.82	7.42	6.96	6.92	7.37	7.37	7.81

Table 15. Average Limestone Analysis

Limestone Analyses, % of oxides	
Component	Moisture free wt%
SiO_2	2.57
Al_2O_3	1.54
Fe_2O_3	0.00
TiO	0.05
CaO	52.64
MgO	1.55
Na_2O	0.11
K_2O	0.08
SO_3	0.08
CO_2 (by difference)	41.36
Moisture, %	0.35

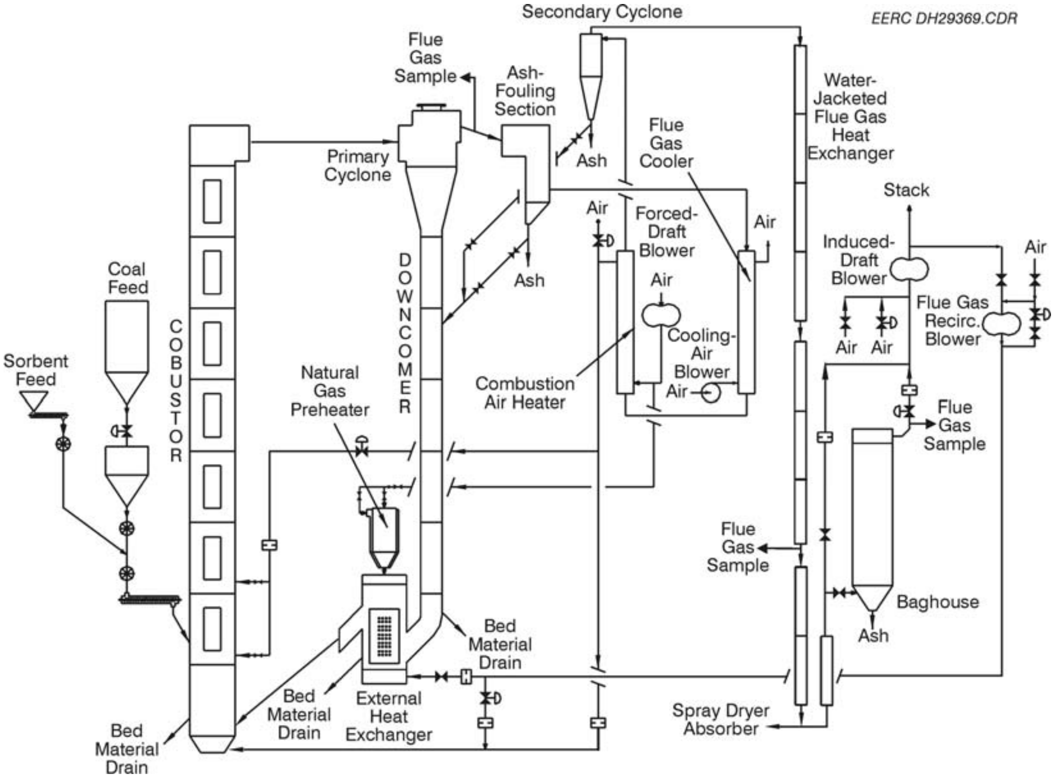

Figure 12. Schematic of EERC CFBC system.

Several coal feed system modifications were completed prior to testing to enable the system to handle a much larger coal and limestone feed rate than is typically fed through this system. First, the coal rotary feed control valve was modified to turn at a high rotational speed. Feed tests were conducted, and the maximum feed rate was approximately 291 kg/hr. Additionally, the coal/limestone rotary seal valve was unable to keep up to this feed rate. Instead of speeding up this valve, the valve was modified by removing the restrictor plates installed in the vane pockets to obtain a more consistent coal feed when feeding coal at lower rates. The coal rotary feed control valve was again modified to turn at an even greater rotational speed. Another set of feed tests was conducted, and the maximum feed rate now exceeded what the 10-cm auger could handle. The auger was then also modified to turn at a higher rotational speed. A final set of coal feed tests was conducted, and it was determined that the coal/limestone feed system was capable of handling at least 386 kg/hr of coal and limestone. At that time, the rate was projected as the maximum feed rate of coal and limestone that would be fed to the system. Additionally, the air purge system was modified before and during testing. These purges are at the bottoms of all of the rotary valves to prevent solid material from accumulating in the pockets of the rotary valves.

OPERATIONAL PERFORMANCE

Heatup on natural gas was started Sunday, January 29, at 0230. Approximately 800 pounds of silica sand was used for start-up. The system was allowed to heat up to greater than 427°C in the combustor before the coal feed was started. Coal feed was then started at 1530 with the Tufanbeyli coal that had been used for the coal feed testing. The average bed temperature was brought up gradually to greater than 593°C before the natural gas was totally switched off at 1754. The average bed temperature was then steadily increased by periodically increasing the coal feed rate so that by 2200 it was approaching Test 1 conditions of 850°C. Table 16 shows the periods identified for the specified test conditions.

For Tests 1 and 2 at full load, it was intended to operate without limestone addition to establish the amount of inherent sulfur capture possible as well as measure other emissions and determine combustion characteristics. The EERC is limited to 8.3 kg/hr of SO_2 based on an agreement with the North Dakota Department of Health. With sufficient inherent sulfur capture by the alkaline components in the coal ash, it was projected that it could be possible to meet this limit without limestone addition. After a short period of operation at Test 1 conditions, as the system stabilized, it was apparent that the SO_2 limit was going to be exceeded, so this test was concluded, and limestone was added to bring the SO_2 emission level below this limit. It was then decided that Test 2 was not possible without exceeding the emission limit, and testing proceeded with the completion of Test 3, a low-load test without limestone addition. By shutting off the flow of the secondary combustion air and reducing the primary combustion air while reducing the coal feed rate, the EERC was able to successfully achieve half-load operation with the permitted levels of SO_2.

Upon completion of Test 3, the system was brought back up to full-load operation, and limestone feed was initiated to turn the bed over as quickly as possible to establish a bed consisting of primarily sulfated lime and coal ash, with limited silica sand present. It had been

Table 16. Summary of Process Data

Test No.:	Test 1	Test 3	Test 6b	Test 5	Test 4	Test 7	Test 6	Test 9	Test 10
Start Time:	23:30	14:00	14:00	2:00	12:00	15:00	1:00	5:45	11:00
Stop Time:	1:08	16:00	14:30	6:00	15:00	21:00	4:20	7:30	13:50
Date:	1/29-30/07	1/30/2007	2/1/2007	2/2/2007	2/2/2007	2/2/2007	2/3/2007	2/3/2007	2/3/2007
Fuel Feed Rate, kg/hr	242	122	279	266	265	255	253	124	127
Sorbent Feed Rate, kg/hr	0	0	54	52	62	64	54	12	17
Combustor Pressure Drop, kPa	11.29	12.26	12.73	14.05	13.5	14.15	13.78	15.05	14.15
Combustion Air:									
EHX[1] Flow, Nm^3/min	1.43	1.62	1.34	1.29	1.27	1.28	1.29	1.42	1.39
Primary Air, Nm^3/min	4.15	2.52	4.87	5.9	5.84	5.82	5.74	3.21	4.09
Secondary Air, Nm^3/min	4.62	0	5.23	4.09	4.01	4	3.99	0	0
Feed Assist Air, Nm^3/min	0.19	0.2	0.21	0.21	0.22	0.22	0.2	0.21	0.21
DC[2] Aeration Air, Nm^3/min	0.43	0.46	0.49	0.45	0.46	0.46	0.5	0.5	0.5
Purge Air, Nm^3/min	0.44	0.44	0.44	0.44	0.44	0.44	0.44	0.44	0.44
Total Air, Nm^3/min	11.25	5.24	12.56	12.39	12.24	12.22	12.16	5.76	6.62
PA/SA, %	49.5	79.1	49.4	58.1	58.1	58.1	57.9	80.2	82.8
Excess Air, %	24.1	21	21.5	20.8	18.8	19.5	20	17.2	20.5
FG[3] SGV[4], m/sec	4.56	2.13	5	4.95	5.1	4.98	4.9	2.39	2.53
EHX SGV, m/sec	0.55	0.49	0.56	0.55	0.55	0.55	0.54	0.53	0.53
BH[5] Air/Cloth	2.07	1.03	2.31	2.08	1.98	2.06	1.97	0.94	1.06
Flue Gas Flow Rate, Nm^3/min	12.15	7.33	15.07	13.86	13.72	13.53	13.5	6.96	7.64

[1] External heat exchange.

[2] Downcomer.

[3] Flue gas.

[4] Superficial gas velocity.

[5] Baghouse.

planned to proceed with limestone feed for at least a period of 24 hours to allow the bed to turn over before proceeding with the remaining tests. Upon completing this 24-hour period trying to stabilize to Test 4 conditions, there were some issues with maintaining a steady coal feed rate and achieving the desired operational average bed temperature while controlling the sulfur dioxide to the required concentration, resulting in another 24 hours before proceeding with the test matrix.

It became apparent during this period that when the coal exceeded approximately 42% moisture content, the coal feed system was unable to feed at a consistent rate, resulting in wide variations in average bed temperature. In fact, when surges in the coal feed rate occurred, in some instances, they were sufficiently high to plug the final coal auger. At one point, the coal auger jammed and burned out some of the electrical components. The components were replaced in sufficient time so that a total restart on natural gas was not required. A couple of hours later, in what was believed to be an unrelated incident, the electric motor for the limestone feed rotary valve burned out and had to be replaced.

Because of problems with maintaining bed temperature, it was decided to attempt the lower-temperature Test 6 conditions for the EPA Method 26 sampling for measurement of bromine, chlorine, and fluorine concentrations in and out of the CFBC baghouse. As the test proceeded, the average bed temperature did tend to drop off and, thus, was designated as Test 6b, with the thought of completing Test 6 conditions later in the week.

It was determined that some additional limited floor drying of the coal would be required to enable better stable coal-feeding characteristics in order to achieve the desired test conditions. It was attempted to limit drying so that the moisture content would be approximately 40% or greater. The higher-than-expected limestone feed rates for sulfur capture resulted in some problems with maintaining the specified operational temperatures. Additionally, as testing proceeded, it was determined that by a combination of very careful control of the amount of secondary cyclone ash drained and with a sufficient limestone addition rate, it was possible to achieve all but Test 8 conditions.

Testing was then completed for full-load Tests 5, 4, 7, and 6 in that order. Then a short transition period was initiated for the low-load conditions of Tests 9 and 10. During Test 9 conditions, the flue gas stream temperature was below the dew point going into the baghouse so the sulfur dioxide analyzer after the baghouse was not getting a good reading. This was not recognized until after the test, when the flue gas heat exchangers were shut off to bring the flue gas temperature going into the baghouse back up. There was then concern with the sulfur dioxide reading at the combustor outlet for Test 10. Several corrective actions were taken during Test 10 to resolve this issue. Testing was completed on February 2, 2007, at 1500.

The biggest issue from an operational standpoint was coal feed. In spite of coal feed system modifications and air-drying of the coal, some hoppers had a high enough moisture level to make consistent coal feed difficult. Figure 13 shows the coal feed rate and average combustor temperature during a period of initially consistent and later poor coal feed. Control of the bed temperature with inconsistent coal feed was extremely difficult.

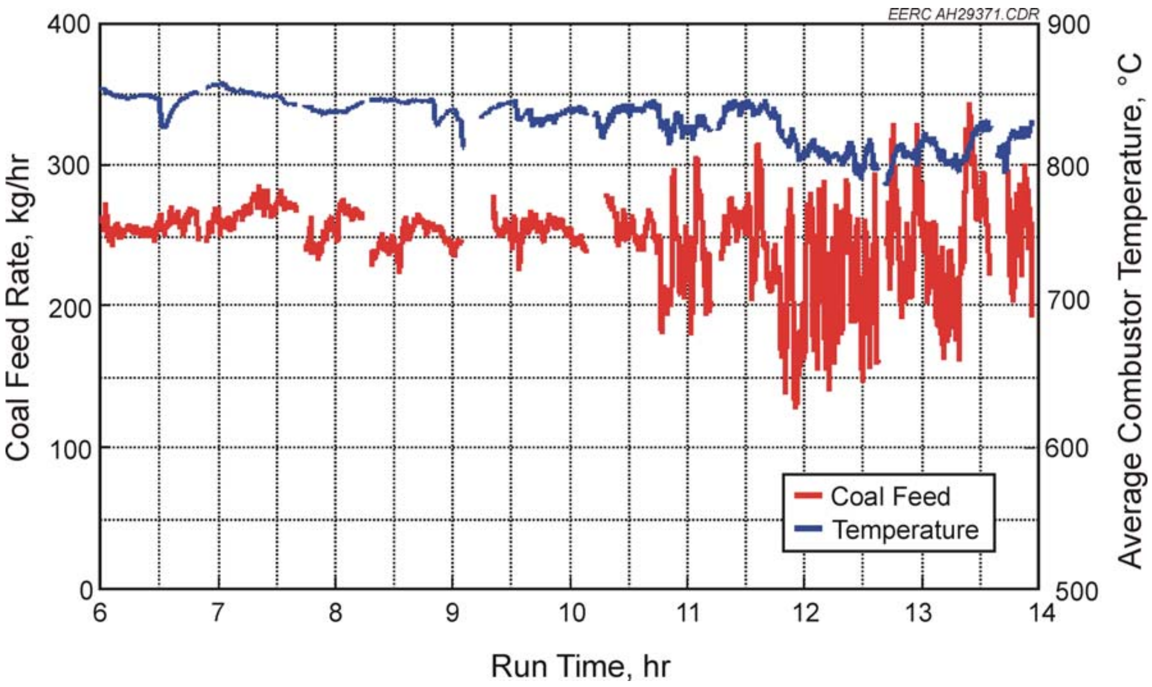

Figure 13. Coal feed and average combustor temperature.

We were able to readily feed this coal after some floor drying to get the moisture content down, approaching 40%. As previously stated, the existing feed systems for both coal and limestone were modified so that all of the rotating mechanical valves and the auger would turn at higher rates of speed; additionally, purges were modified with the addition of flowmeters so that there was a positive indication of purge flows to the valves. At coal moistures of 42% or higher, it was not possible to maintain a steady feed rate. The coal was likely sticking to the pockets of the coal rotary feed valve. A high-pressure air pulse, duration, and frequency set on the control computer was used in this location to help prevent this from occurring. Going to a high rate, long-duration pulse was still not sufficient to maintain a steady coal feed rate. In some instances when coal feed rate fluctuations were significantly high, a surge in the coal feed rate would be sufficient to plug the coal/limestone feed auger, resulting in a total loss of coal feed until the auger plug could be cleared and coal feed restarted.

In a larger full-scale system, it should be possible to handle this coal at higher moisture contents, greater than 40%, with careful attention to the coal feed system design. Live-bottom bins should be used for feeding the coal where possible. Equipment should be oversized so that when coal adheres to surfaces, there will still be sufficient feed capability. It would be highly recommended that the coal be partially dried down to some minimal level, possibly 42% or lower, before final feed into the combustor. This will result in a more stable feed rate, which is essential for good operation of a full-scale system. Additionally, if the moisture can be removed from the coal, the coal will have a higher heating value, which will be more beneficial for improved system efficiency.

Sustainable combustion was achieved with this coal. Before limestone feed was begun, some cooling coils in the external heat exchanger were required to control the desired operating temperature. After limestone feed was started, the loss of heat through the refractory walls was sufficient, with no additional water-cooled heat transfer required to control the desired operating temperature. Therefore, limited heat-transfer data will be presented here.

As previously stated, after limestone feed was started, temperature control in the bed was more of an issue. High limestone feed rates were required to obtain the low sulfur dioxide target emissions specified. Heating up the limestone to the operational temperature of the combustor required some additional energy. It was also likely that the heat of calcination, an endothermic reaction, at these high limestone feed rates was significantly greater than the heat released by the partial sulfation of the lime, an exothermic reaction. While not directly measured, it is believed that high solids recirculation rates were occurring during this test, because of the large amounts of fine coal ash present and the high limestone feed rates. The large amounts of solids that were captured by the secondary cyclone were significantly cooler, and when introduced back to the downcomer, this resulted in a larger amount of these cool solids being introduced back into the external heat exchanger and then into the combustor. These three factors all contributed to the progressive cooldown of the system as operation progressed. When draining all or most of the solids collected by the secondary cyclone, this resulted in the combustor operating and sustaining higher operational temperatures; however, it removed a large portion of the available lime in the system for capturing sulfur dioxide. So to simultaneously control sulfur dioxide emissions and operate at a consistently high operating temperature without excessively high limestone feed rate, it was necessary to balance the amount of solids circulated back to the combustor with the amount of secondary cyclone solids drained out of the system. This was accomplished for Test 4 conditions.

Temperature control will be less of an issue with a full-scale system, as solids will likely be only circulated from the primary cyclone. It can be concluded from this that it will be critical to have as efficient a primary cyclone as possible. This will result in a high recirculation rate of hot solids to sustain high operating temperatures. It will keep the small calcined lime particles in the system available for efficient sulfur capture. It appears that minimal or no external heat exchange will be required for operation with this coal.

SUMMARY OF RESULTS

Upon completion of the run, data for each of the steady-state periods were averaged. A summary of the process data for each test period is presented in Table 16; operating temperatures are presented in Table 17. The test periods correspond to those presented in the test matrix listed in Table 13.

Table 17. Operating Temperatures, °C

	Test 1	Test 3	Test 6b	Test 5	Test 4	Test 7	Test 6	Test 9	Test 10
Combustor Temperatures									
Plenum	186	111	221	219	220	220	208	94	119
Section 1	848	844	820	833	855	854	818	846	810
Section 2	858	848	822	835	857	855	820	848	811
Section 3	837	826	798	806	827	826	793	816	784
Section 4	856	842	811	820	842	840	806	829	796
Section 5	827	818	789	796	816	816	783	803	773
Section 6	848	837	809	817	838	837	803	823	792
Section 7	846	834	808	816	837	836	802	821	791
Section 8	905	874	853	864	888	886	847	866	833
Section 9	907	866	854	865	889	886	847	859	831
Average	852	841	815	826	847	846	811	835	801
EHX Temperatures									
Plenum	53	53	53	53	53	53	53	53	53
1.5' Above Dist.	731	488	809	824	871	846	798	687	715
2.7' Above Dist.	652	487	781	792	816	813	796	685	714
3.8' Above Dist.	712	496	798	802	824	821	783	673	696
5.3' Above Dist.	677	496	758	763	787	782	747	644	624
Average	693	492	787	796	824	815	781	672	687
Downcomer Temperatures									
Section 3	724	612	762	777	802	798	750	666	661
Section 4	757	704	784	793	817	812	772	730	722
Section 6	814	734	801	809	832	829	789	754	752
Section 8	822	764	NA	NA	NA	NA	NA	769	754

RECIRCULATION RATES AND SIZE DISTRIBUTIONS

Normally, recirculation rate is calculated using a heat balance around the external heat exchanger. This calculation uses the heat removed by the water-cooled EHX coils as one of the inputs; since EHX coils were not used for most of the testing, no online recirculation calculation was available. The other component of the calculation is the temperature of the solids entering the EHX, which is strongly influenced by the temperature and mass of solids recycling back from the secondary cyclone.

Ash samples were collected at the end of most test periods. The size distributions of the bed material, downcomer, and secondary cyclone samples are shown in Figure 14. The d_{50} of the bed material samples ranged from 340 µm for Test 6 to 477 µm for Test 3. The downcomer samples for full-load tests had d_{50}s only slightly smaller than the bed material, ranging from 287 µm for Test 7 to 337 µm for Test 4. The low-load/low-velocity tests, however, produced a much smaller-sized downcomer ash, as one would expect. The cyclone ash samples had d_{50}s between 32 (low-load Test 10) and 74 µm (Test 7). The baghouse ash samples were fairly consistent, with d_{50}s ranging from 6 to 10 µm. These are shown, along with downcomer and cyclone sizes, in Figure 15.

Bottom Ash–Total Ash Split

The ash balance for each test period is presented in Table 18. Ash input to the system was composed of calculated quantities of coal and limestone ash, based on their respective analyses and feed rates. The limestone-derived ash was further broken down into estimates of sorbent which was either calcined or had undergone sulfation. The output was composed of measured quantities of bottom ash (drained from the combustor bed), fly ash removed from the secondary cyclone, and fly ash removed from the baghouse.

The ratios of bottom ash to total ash, as well as the percent closure, are included in Table 16. For full-load tests with limestone addition, the bottom-to-total ash split ranged from 38% to 57%. Bottom ash was drained periodically to try to maintain a consistent amount of bed inventory (as measured by pressure drop across the bed). The secondary cyclone drain rate was kept at a minimum to try to retain as much unreacted limestone in the bed as possible, although cyclone ash was drained (rather than recycled) occasionally to trim the combustor temperature, as described earlier. The bottom-to-total ash split was much lower for the tests without limestone and significantly higher for Test 10 (at low load), where no material was taken from the secondary cyclone. No cyclone ash was drained during Test 9 either, but there was an increase in bed dP during that test period, indicating that more bed material could have been drained during that test.

Figures 16–19 show, respectively, the compositions of the bed material, downcomer solids, secondary cyclone material, and baghouse ash for test periods as determined by XRFA. Additionally, for comparison purposes, the composition of the coal ash and limestone are also shown. Initially, start-up was with a silica sand bed. Solid samples were not obtained for Test Period 1 because of the initial confusion identifying this as a valid test period. The bed material,

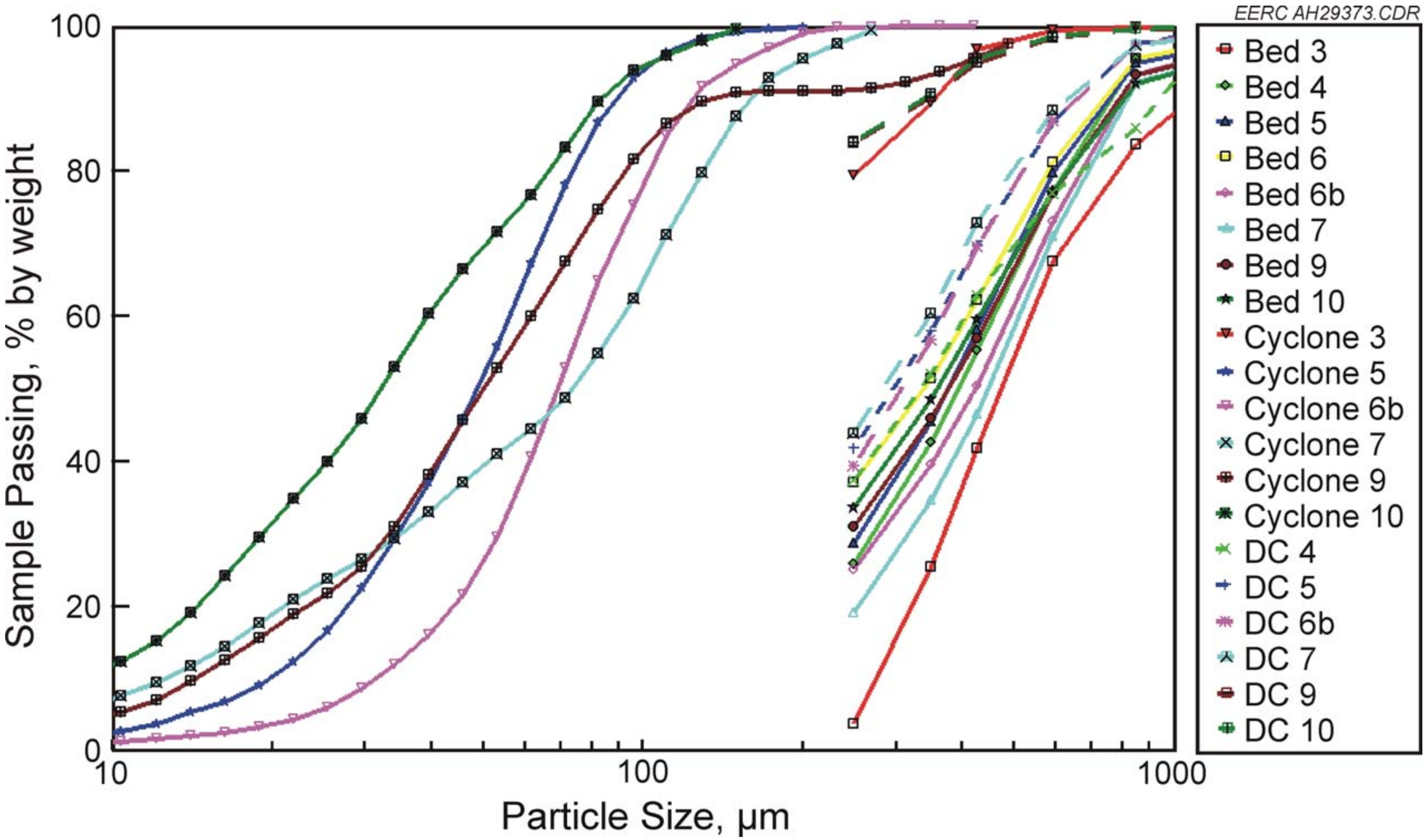

Figure 14. Size distributions of the bed material, downcomer, and secondary cyclone samples.

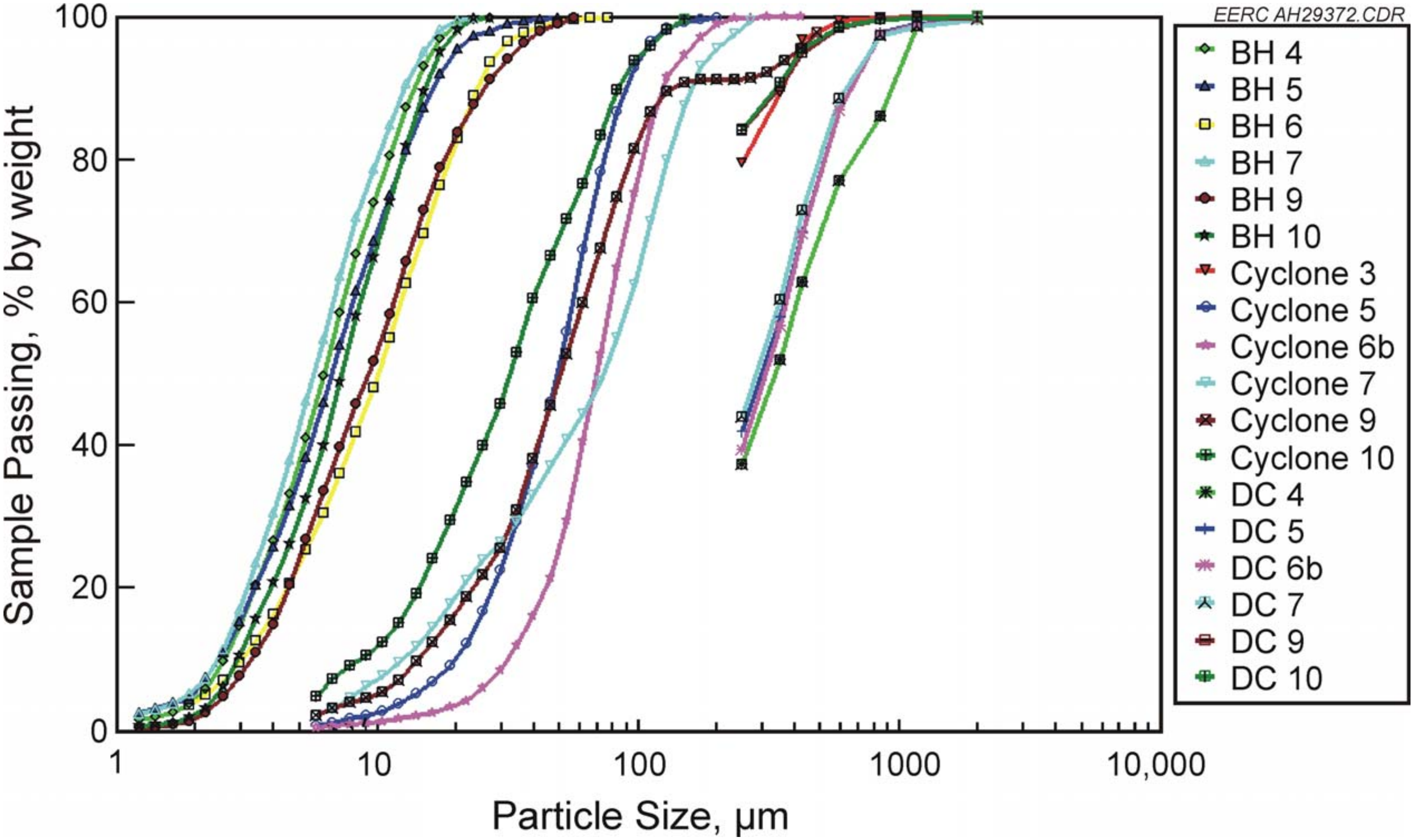

Figure 15. Size distributions of the downcomer, secondary cyclone, and baghouse ash samples.

Table 18. Ash Balance

	Test 1	Test 3	Test 6b	Test 5	Test 4	Test 7	Test 6	Test 9	Test 10
Input, kg/hr									
Ash	60.9	29.2	68.7	76.6	75.1	71.0	65.1	31.8	38.0
Sorbent: [1]									
CaO	0.0	0	20.1	20.0	24.9	25.9	21.0	3.4	5.3
$CaSO_4$	0.0	0	19.6	18.0	18.4	19.1	18.3	7.5	9.2
Inerts	0.0	0	3.2	3.1	3.7	3.8	3.2	0.7	1.0
Total Solids In	60.9	29.2	111.7	117.8	122.0	119.9	107.6	43.5	53.5
Output, kg/hr									
Bed Material Drained	0.0	1.4	68.0	49.4	52.6	41.7	57.2	17.2	25.9
Cyclone Ash	48.1	21.8	20.9	51.7	60.8	60.8	17.7	0.0	0.0
Baghouse Ash	20.0	15.4	27.7	19.1	13.6	13.6	25.4	21.3	7.3
Total Solids Out	68.0	38.6	116.6	120.2	127.0	116.1	100.2	38.6	33.1
Closure, %	111.7	132.3	104.4	102.0	104.1	96.8	93.2	88.7	61.8
Bottom Ash/Total Ash, %	0.0	3.5	58.4	41.1	41.4	35.9	57.0	44.7	78.1

[1] The CaO and $CaSO_4$ mass inputs are included to express sorbent equivalent mass inputs.

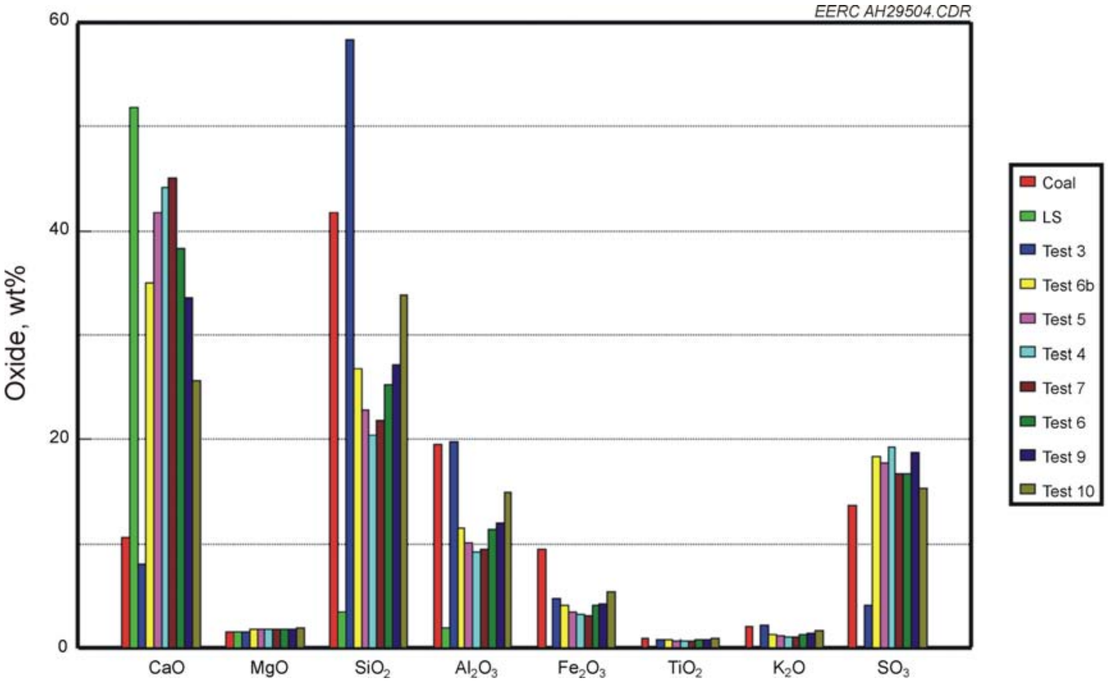

Figure 16. Bed material composition.

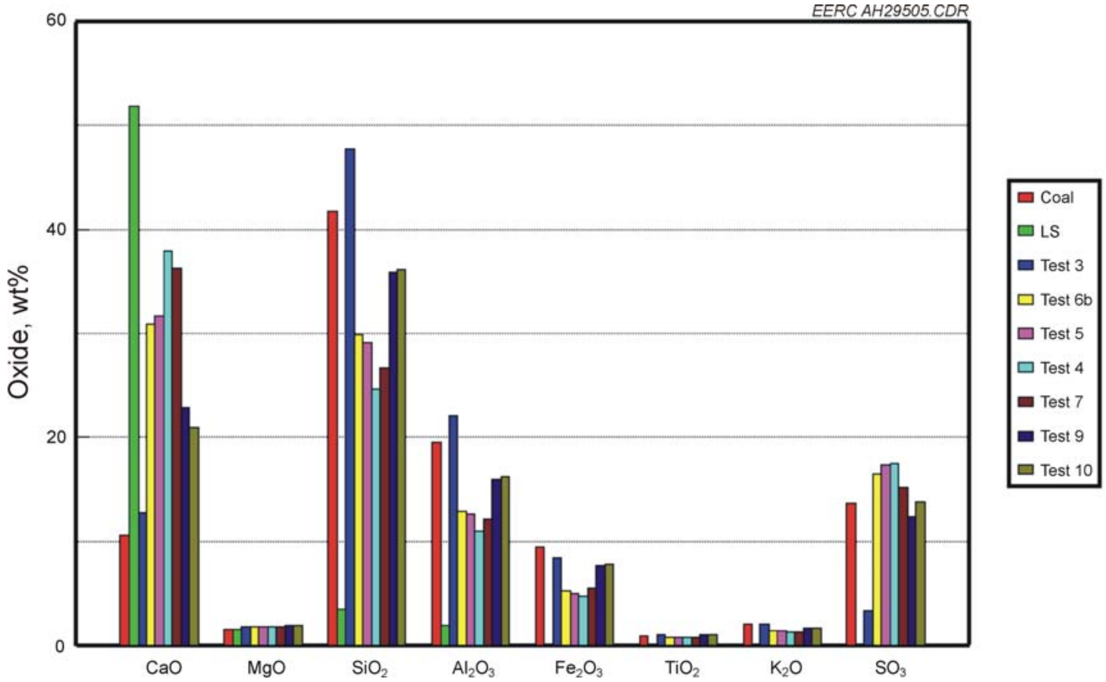

Figure 17. Downcomer solids composition.

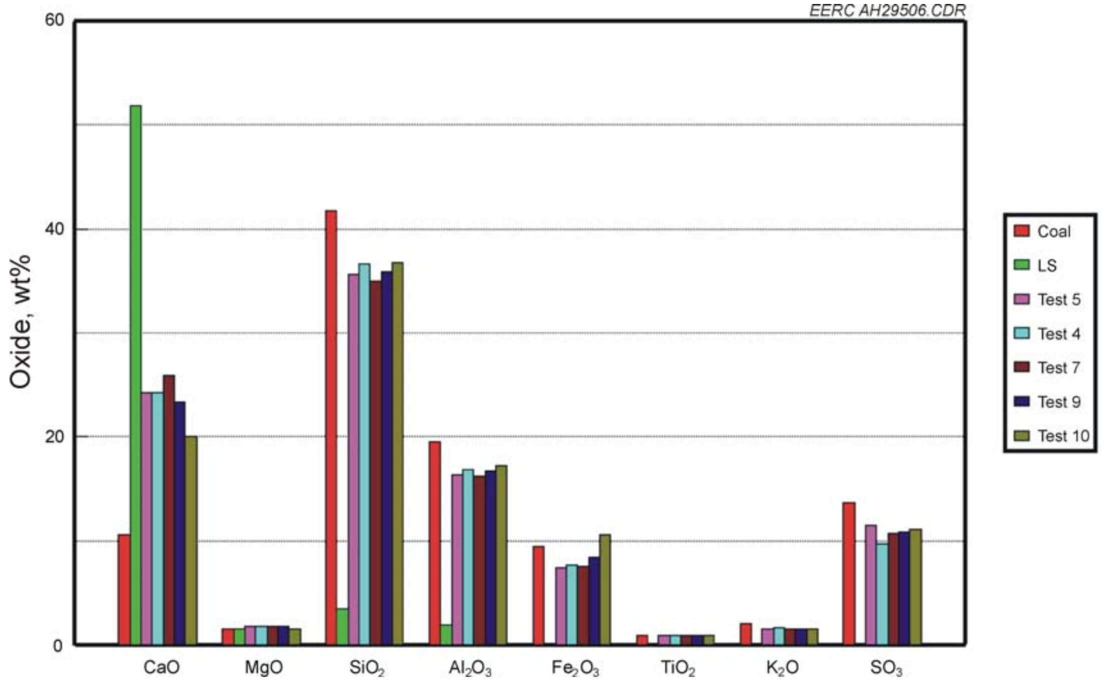

Figure 18. Secondary cyclone material composition.

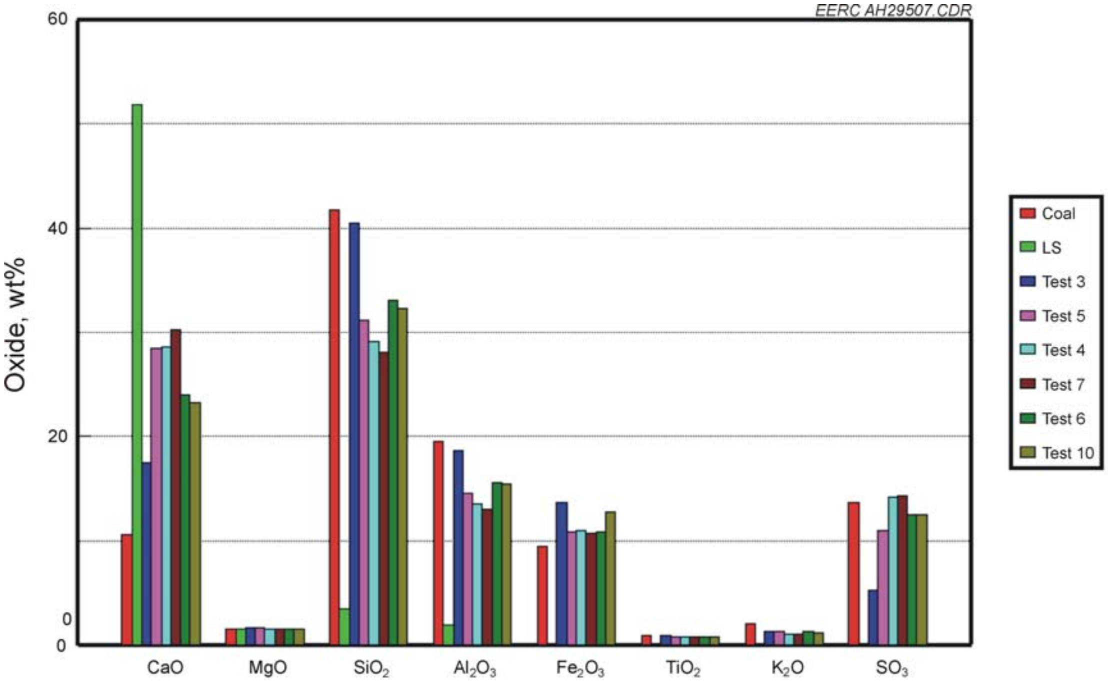

Figure 19. Baghouse ash composition.

downcomer material, and secondary cyclone material would have been close to 100% silica sand. The baghouse ash would likely have been mostly coal ash. After Test 3, when limestone feed was initiated, there was sufficient operational time to turn the bed over so that operation was with primarily a bed of limestone and coal ash. Operation at low-load conditions for Tests 9 and 10 did result in changes in the overall bed material composition. A smaller ratio of limestone addition to coal feed allowed the bed to approach the composition of the coal ash.

This fuel requires relatively high limestone and coal feed rates compared to operation with other fuels. At full-load conditions, the bed was turning over approximately every 4.5 hours, assuming an inventory of about 365 kg of bed material. Based upon the feed rates, drain rates, and overall solids compositions, there was no indication of any significant bed material attrition or agglomeration of the limestone or coal ash. In a full-scale system, the best location for draining bed material would be from the loop seal. The solids are somewhat cooler here, so less energy would be lost as opposed to draining large amounts of bed material from the combustor. Also, the size distribution of the solids in the loop seal were very similar to that of the bed material, so it did not appear that there were a lot of larger particles accumulating in the bed that would have to be drained on a frequent basis.

Combustion Efficiency

Loss on ignition (LOI) was performed on several bed material, cyclone ash, and baghouse ash samples. They were extremely low for all tests – less than 1% in all cases and less than 0.1% in many cases. Since LOI includes both carbon and carbonate, such low LOIs for the tests, including limestone addition, indicate extremely good combustion efficiency (greater than 98%). The combustion efficiencies are presented in Table 19.

Boiler Efficiency

Table 20 shows the calculated boiler efficiency for all of the test periods for this test. Boiler efficiency was calculated using a modified American Society of Mechanical Engineers (ASME) PTC 4.1 as recommended by the Electric Power Research Institute (EPRI) and assuming a 149°C (300°F) stack gas exit temperature and a 0.4% loss due to radiative and convective losses. This fuel had the lowest boiler efficiencies measured when compared to operation of all other fuels tested on the EERC CFBC system at similar test conditions. Test 7 had a boiler efficiency of 65.9% compared previously to a low of 77.6% measured for an Asian lignite to a high of 89.1% for a low-sulfur subbituminous coal tested on the EERC CFBC. The low boiler efficiencies obtained for the Tufanbeyli lignite were due primarily to the large amount of energy required to vaporize the moisture in the lignite plus the high limestone feed rates required for high sulfur dioxide capture rates. Test 1 without limestone addition had the highest full-load boiler efficiency of 77.0%.

Temperature Profiles

The temperature profiles (as a function of height above the combustor distributor plate) are shown in Figure 20. The temperature drop at 3.7 m (6 feet) is the result of cold solids (coal and

Table 19. Combustion Efficiency

	Test 1	Test 3	Test 6b	Test 5	Test 4	Test 7	Test 6	Test 9	Test 10
Input:									
Fuel Feed Rate, kg/hr	163	82	188	179	178	171	170	83	85
Fuel Carbon, %	21.5	21.2	20.7	20.1	19.0	20.2	19.5	19.5	21.3
Carbon Feed Rate, kg/hr	35.0	17.4	38.8	36.0	33.8	34.5	33.0	16.2	18.2
Total, kg/hr	35.0	17.4	38.8	36.0	33.8	34.5	33.0	16.2	18.2
Output:									
Bottom Ash Discharge Rate, kg/hr	0	1	46	33	35	28	38	12	17
Unburned Carbon, %	NA	0.0	0.0	0.0	0.0	0.7	0.0	0.0	0.2
Bottom Ash Carbon Discharge Rate, kg/hr	NA	0.0	0.0	0.0	0.0	0.2	0.0	0.0	0.0
Cyclone Ash Discharge Rate, kg/hr	32	15	14	35	41	41	12	0	0
Unburned Carbon, %	0.1	0.2	0.4	0.8	0.8	0.8	0.7	0.7	0.3
Cyclone Ash Carbon Discharge Rate, kg/hr	0.0	0.0	0.1	0.3	0.3	0.3	0.1	0.0	0.0
Baghouse Discharge Rate, kg/hr	13	10	19	13	9	9	17	14	5
Unburned Carbon, %	NA	0.7	0.3	1.2	1.2	1.3	1.4	4.8	0.7
Baghouse Carbon Discharge Rate, kg/hr	NA	0.1	0.1	0.2	0.1	0.1	0.2	0.7	0.0
Total, kg/hr	0.02	0.10	0.12	0.42	0.44	0.64	0.32	0.68	0.08
Combustion Efficiency, %	NA	99.43	99.70	98.84	98.70	98.15	99.03	95.77	99.58

Table 20. Boiler Efficiency

	Test 1	Test 3	Test 6	Test 5	Test 4	Test 7	Test 6	Test 9	Test 10
Assume Flue Gas Exit Temp., °C	149	149	149	149	149	149	149	149	149
Losses, kJ/hr									
Dry Gas	118,148	57,988	130,815	120,516	114,131	117,569	111,929	53,630	61,315
Water in Fuel	262,903	141,248	322,656	276,770	294,850	286,941	296,297	144,798	125,047
Comb. of Fuel Hydrogen	12,571	6203	14,134	13,162	12,386	12,643	12,118	5922	6712
Unburned Carbon	788	4814	5706	20,328	21,464	31,144	15,572	33,328	3770
Sorbent Calcination	0	0	95,359	92,935	109,906	114,350	96,572	22,062	30,790
Radiation and Convection[1]	8096	3904	8738	7908	7,388	7056	7447	3639	3968
Solids	63,240	35,345	103,495	108,158	117,350	107,153	88,552	35,079	28,886
Sorbent Sulfation	0	0	(76,437)	(70,220)	(71,544)	(74,578)	(71,089)	(29,209)	(35,582)
Total	465,746	249,501	604,466	569,557	605,930	602,279	557,398	269,250	224,906
Losses, %									
Dry Gas	5.8	5.9	6.0	6.1	6.2	6.7	6.0	5.9	6.2
Water in Fuel	13.0	14.5	14.8	14.0	16.0	16.3	15.9	15.9	12.6
Comb. of Fuel Hydrogen	0.6	0.6	0.6	0.7	0.7	0.7	0.7	0.7	0.7
Unburned Carbon	0.0	0.5	0.3	1.0	1.2	1.8	0.8	3.7	0.4
Sorbent Calcination	0.0	0.0	4.4	4.7	6.0	6.5	5.2	2.4	3.1
Radiation and Convection[1]	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4	0.4
Solids	3.1	3.6	4.7	5.5	6.4	6.1	4.8	3.9	2.9
Sorbent Sulfation	0.0	0.0	-3.5	-3.6	-3.9	-4.2	-3.8	-3.2	-3.6
Total, %	23.0	25.6	27.7	28.8	32.8	34.1	29.9	29.6	22.7
Boiler Efficiency, %	77.0	74.4	72.3	71.2	67.2	65.9	70.1	70.4	77.3

[1] Assumes 0.4% radiative and convective losses.

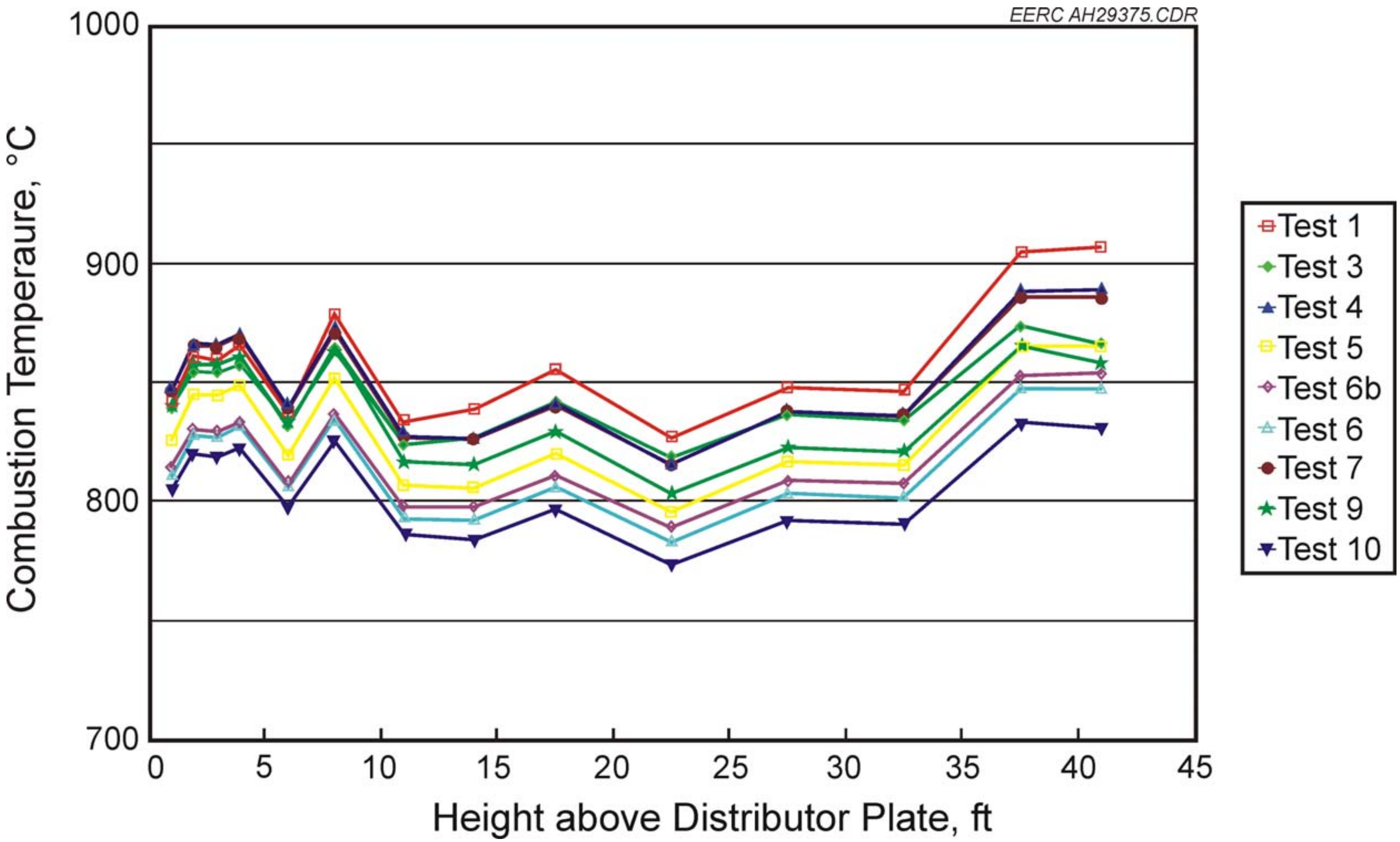

Figure 20. Combustor temperatures as a function of height above distributor plate.

limestone) entering the combustor at this location. In previous runs on this unit, we see a temperature drop at 1.8 m (3 feet) above the distributor plate, where solids return to the combustor from the EHX. However, since very little heat was removed from the EHX during these tests, no change in temperature was observed at that height. The lack of EHX heat removal resulted in a somewhat higher temperature in the dense bed region than in the upper portions of the combustor, where typically the reverse is true.

ENVIRONMENTAL PERFORMANCE

Average flue gas emissions for each steady-state test period are presented in Table 21 and discussed in the following sections. Flue gas was sampled at the primary cyclone exit and the baghouse exit. The difference in oxygen content between the two sampling locations is the result of air leakage into the system between the primary cyclone and the baghouse. All emissions have, therefore, been corrected to 6% oxygen.

Overall this test matrix was better suited to evaluating operability with coal as opposed to obtaining sufficient data to evaluate the emission trends. While the narrow range of operating temperatures selected for this testing are likely typical of a-full scale operational plant, a wider range would have made it easier to establish the emission trends for this coal. Additionally, the difficulty with achieving the high-temperature Test No. 8 condition resulted in not having any high-temperature data for comparison purposes. The behavior of the system with respect to the flue gas emissions at half-load conditions was significantly different compared to full-load operation.

Sulfur Dioxide Emissions

The temperature range over which these tests were conducted was not sufficiently large to determine the optimum temperature for sulfur capture. Additionally, the two very high sulfur capture rates specified (requiring sulfur retention rates of 96.5% and 98.3%) are not sufficiently different to establish a trend on how the calcium-to-sulfur ratio will affect sulfur retention at a given operating temperature. For full-load operation, it appears that better sulfur capture was possible at approximately 816°C compared to operation at approximately 843°C. Sulfur capture was significantly affected by both the limestone feed rate and by the drain rate of the solids collected by the secondary cyclone. The most effective limestone in the system for good sulfur capture is typically collected by the secondary cyclone. Normally, all of the solids collected by the secondary cyclone would be fed back into the system to be available for sulfur capture. For this coal, the amount of solids to the secondary cyclone tended to increase more rapidly than is typical for most coals. Each time these cooler solids were returned to the downcomer, they would progressively force the combustor operating temperature to slowly decrease. Draining a consistent portion of these solids permitted operation at a more steady and consistent temperature as long as steady coal feed rates, limestone feed rates, and combustion air addition flows were maintained. It would be projected from this testing that with calcium-to-sulfur addition rates approaching 3.5, it should be possible to obtain sulfur dioxide emissions down to 200 mg/Nm^3. This will ultimately be dependent upon such factors as the primary cyclone efficiency and the limestone properties.

Table 21. Flue Gas Emission Data

	Test 1	Test 3	Test 6 b	Test 5	Test 4	Test 7	Test 6	Test 9	Test 10
Combustor Outlet									
O_2, %	4.10	3.70	3.70	3.60	3.30	3.40	3.50	3.10	3.60
CO Content, ppm	22	57	82	61	38	38	96	39	79
CO Content[1], ppm	20	49	71	53	32	32	82	33	68
CO Emission, mg/Nm^3 @ 6% O_2	24.4	61.8	88.9	65.7	40.3	40.5	102.9	40.9	85.1
CO_2 Content, %	16.0	15.6	17.1	17.1	17.6	17.6	16.9	16.8	16.4
CO_2 Content[1], %	13.5	13.5	14.8	14.7	14.9	15.0	14.5	14.1	14.1
NO_x Content, ppm	89	34	134	131	139	148	128	34	57
NO_x Content[1], ppm	79	29	116	113	118	126	110	28	49
NO_x Emission, mg/Nm^3 @ 6% O_2	162	61	239	232	242	259	225	59	101
SO_2 Content, ppm	>5000	>5000	129	198	192	76	176	242	39
SO_2 Content[1], ppm			112	171	163	65	151	203	34
SO_2 Emission, mg/Nm^3 @ 6% O_2			320	488	465	185	431	579	96
SO_2 Retention[2], %			98.1	97.1	97.4	98.9	97.5	96.5	99.4
EA	24.1	21.0	21.5	20.8	18.8	19.5	20.0	17.2	20.5
Baghouse Outlet									
O_2, %	4.60	5.60	5.50	4.70	4.20	4.30	4.90	6.10	5.20
CO Content, ppm	51	63	56	48	41	39	61	34	79
CO Content[1], ppm	47	61	54	44	37	35	57	34	75
CO Emission, mg/Nm^3 @ 6% O_2	58.3	76.7	67.7	55.2	45.8	43.8	71.0	42.8	93.8
CO_2 Content, %	15.8	15.4	16.0	15.6	16.1	16.2	16.0	15.7	15.9
CO_2 Content[1], %	14.5	15.0	15.5	14.4	14.4	14.6	14.9	15.8	15.1
NO_x Content, ppm	95	33	130	130	141	151	116	32	50
NO_x Content[1], ppm	86	32	125	119	126	136	108	32	48
NO_x Emission, mg/Nm^3 @ 6% O_2	178	66	258	245	258	279	221	65	98
N_2O Content, ppm	NA	NA	67	55	43	38	59	15	24
N_2O Content[1], ppm			65	51	38	34	55	15	23
N_2O Emission, mg/Nm^3 @ 6% O_2			127.4	99.8	74.9	66.9	107.4	29.1	44.9
SO_2 Content, ppm	5407	4855	110	185	169	66	153	3	176
SO_2 Content[1], ppm	4945	4729	106	170	151	59	143	3	167
SO_2 Emission, mg/Nm^3 @ 6% O_2	14,130	13,511	304	486	431	169	407	9	477
SO_2 Retention[2], %	16.3	24.1	98.2	97.0	97.5	99.0	97.7	100.0	97.4
Ca/S ratio (ls[3] only)	0.00	0.00	2.69	2.78	3.33	3.44	3.01	1.41	1.66
Ca/S ratio (total)	0.61	0.68	3.42	3.59	4.19	4.24	3.71	2.10	2.35
Ca Utilization (ls[3] only)	NA	NA	36.4	34.9	29.3	28.8	32.4	70.9	58.6
Ca Utilization (total)	26.7	35.4	28.7	27.1	23.3	23.3	26.3	47.5	41.4
Alkali-to-Sulfur	0.63	0.69	3.44	3.60	4.20	4.26	3.72	2.12	2.37
Alkali Utilization	26.1	34.7	28.6	26.9	23.2	23.2	26.2	47.2	41.2
Avg. Comb. Temp., °C	852	841	815	826	847	846	811	835	801
Moisture in FG, vol%	21.9	22.7	22.6	20.8	21.3	22.0	22.3	22.3	20.6
Fuel Carbon, %	21.5	21.2	20.7	20.1	19.0	20.2	19.5	19.5	21.3
Fuel Sulfur, %	2.4	2.4	2.1	2.1	2.1	2.2	2.1	2.1	2.5

[1] Corrected to 6% O_2.

[2] Moisture-free coal carbon and sulfur values used in the sulfur retention calculation.

[3] Limestone.

Better sulfur retention was achieved at the low-load conditions since high sulfur retention rates were possible at lower calcium-to-sulfur ratios compared to the high-load conditions. It is likely that lower velocities in the combustor allowed the limestone to stay in the system longer for better sulfur capture instead of being blown out at the higher velocities at full-load conditions.

Tests 1 and 3 used no limestone addition for sulfur capture. The inherent sulfur capture for Test 1 was about 27%, with SO_2 emissions (corrected to 6% O_2) greater than 5000 ppm. With limestone addition, SO_2 emissions could be controlled to less than 200 ppm or sulfur retention greater than 97%. An added calcium-to-sulfur ratio of 2.7 to 3.4 was required to achieve this level of sulfur removal at full load; at partial load, the desired level of sulfur capture was achieved with Ca/S ratios around 1.5, because of the lower velocity and, therefore, longer residence time of the calcium in the bed. Sulfur dioxide levels measured at the combustor outlet and the baghouse outlet matched reasonably well for most tests. During low-load Test 9, the baghouse temperature dropped below the dew point, causing the sulfur dioxide in the flue gas to be scrubbed out in the baghouse; there was a discrepancy between the two analyzers for Test 10, with the baghouse outlet showing a much higher SO_2 level than the combustor outlet. Because the limestone feed rate was higher for Test 10 than Test 9, it is assumed that the combustor outlet SO_2 reading is the correct one.

High sulfur capture was achieved during this testing at high added Ca/S ratios often greater than 3. In a full-scale system, it would be expected that similar ratios of Ca/S would be required for the same degree of sulfur capture. It will again depend significantly on the full-scale system geometry and, ultimately, the efficiency of the primary cyclone in that system. In the EERC CFBC, it is likely that we are capable of returning more of the finer-sized reactive lime compared to a full-scale system. A full-scale system is more likely have longer solids residence times for the lime to react for good sulfur capture, mainly because of its increased height. These two factors should somewhat offset each other. It would likely be much more efficient both in terms of the reduction of overall limestone consumption and higher boiler efficiency to only capture a portion of the coal sulfur in-bed and rely on a posttreatment sulfur capture system to control sulfur dioxide emissions to the desired level. An economic analysis will be necessary to determine to what degree that only in-bed sulfur capture be relied upon compared to incorporating a posttreatment sulfur capture system.

We were able to successfully operate at the low-load test parameters with this coal, with only a moderate drop in the operational combustor average bed temperature down to approximately 788°C.

NO_x Emissions

Because CFBC boilers operate at a lower temperature than pc-fired boilers, NO_x emissions tend to be significantly lower. For the full-load tests with the Tufanbeyli coal, NO_x (measured as combined NO and NO_2) ranged from 86 to 136 ppm (corrected to 6% O_2) at the baghouse outlet. The corrected NO_x emissions for the low-load tests ranged from 32 to 48 ppm at the baghouse outlet. Although not specifically determined, these measured NO_x emissions are believed to be primarily NO (nitric oxide) NO_x emissions. As with the SO_2, the NO_x emissions measured at the two locations were reasonably close. These NO_x emissions were consistent, about average, when compared to the NO_x emission results for previous coals tested on the EERC CFBC.

N_2O was measured only at the baghouse outlet and was not measured at all during the tests without limestone addition, since high SO_2 interferes with the N_2O measurement in that analyzer. Unlike NO_x, N_2O emissions increase with decreasing temperature and tend to be higher in CFBC boilers than in pc-fired units. The corrected N_2O emissions measured during these tests ranged from 15 ppm for low-load Test 9 to 65 ppm for Test 6b. There was a slight decrease in N_2O emissions at higher combustor temperatures. N_2O emissions for this coal tended to be lower than for most of the other coals tested on the EERC CFBC. The combined totals of the NO_x and N_2O emissions were typical of the results obtained for other lignites tested on the EERC CFBC, tending to be approximately two-thirds of the amount measured for subbituminous or bituminous coals tested on the EERC CFBC.

Method 26 Sampling

EPA Method 26 sampling at the baghouse inlet and outlet was performed on January 31 and February 1; in both cases, the inlet and outlet were sampled simultaneously. The first set of samples was discarded because a coal plug in the feed auger caused the sampling to be stopped after just a few minutes. The results of the second set of samples are shown in Table 22. No bromide or fluoride was detected in any of the samples; chloride was only detected in the baghouse outlet sample. The baghouse outlet sampling period was 60 minutes; the inlet sample was stopped after 35 minutes because the high ash content in the flue gas into the baghouse plugged the sample filter. This difference in sampling duration may explain the presence of chloride in the outlet, while none was detected at the inlet. The actual amount of chloride detected in the outlet sample was only 0.5% of the amount fed in with the coal.

Sintering, Agglomeration, and Deposit Evaluation

No agglomeration tendencies were noted with this coal during this test. There was a very minor layer of ash observed on the fouling tubes located in the simulated convective pass region for this system. Figure 21 shows a picture of the back side of the fouling tubes, with Figure 22 showing the front of the tubes. A very thin layer of ash was present on the tubes. It would not be expected that any significant amount of soot blowing would be required in a full-scale system. There will be, however, a significant quantity of solids produced by the combustion of this coal that will have to be handled. If possible, solids should be drained from the loop seal region where they are the coolest to have the least impact on the system to maintain operational temperatures.

CONCLUSIONS

CFBC appears to be better suited for use with this coal compared to pc combustion. This fuel should be dried as much as possible in any manner that is practical. Conservative design practices should be used for the fuel-handling system. It is recommended to operate this system at as high a velocity as is possible in a full- scale system to help increase the thermal loading into the system. If low sulfur dioxide emissions will be required, installing a post-sulfur dioxide

Table 22. Results of Method 26 Halide Sampling

	Concentration in Coal, µg/g	Theoretical Concentration in Flue Gas, mg/Nm^3	Measured Concentration, mg/L		Concentration in Sampled Flue Gas, mg/Nm^3	Removal Efficiency, %
			Baghouse Inlet	Baghouse Outlet		
Bromide	18.4	223	<1[1]	<1	NA[2]	NA
Chloride	67.8	818	<1	6.1	5	99.5
Fluoride	450	5406	<1	<1	NA	NA

[1] Below detection limit.
[2] Not applicable.

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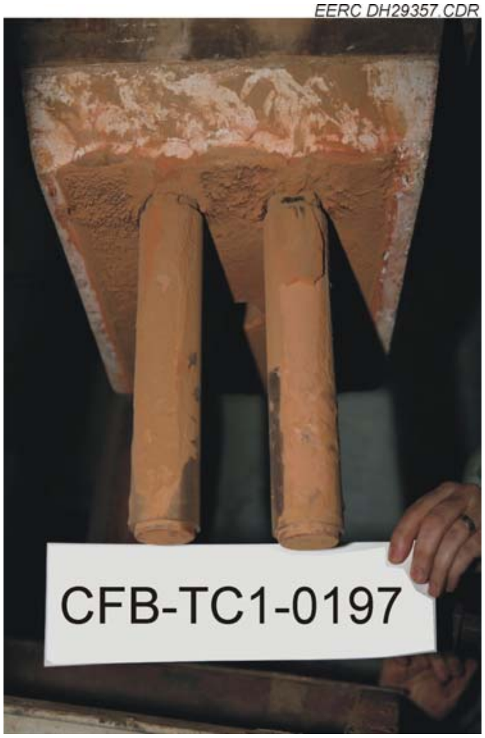

Figure 21. Back side of the ash-fouling probes removed from EERC CFBC convective pass.

Figure 22. Front side of the ash-fouling probes removed from EERC CFBC convective pass.

treatment system should be considered. If possible, solids should be drained from the loop seal region where they are the coolest to have the least impact on the system to maintain operational temperatures. Bed material agglomeration and convective pass fouling are not be expected to a problem with this fuel in a full-scale fluidized-bed boiler system.

The limestone tested was a good candidate for use with this coal. While the size tested was larger than specified for this project, the size tested provided for effective sulfur capture. It is recommended that some additional testing be performed for the full-scale system to determine the optimum size range for that system. Larger-sized limestone will stay in the system longer, thus having a longer residence time for potentially more sulfur capture, but will have less surface area available so that all of the particles can be used for sulfur capture. The same mass of smaller-sized limestone particles will have more surface area available for sulfur capture as long as they can be kept in the system for sufficient time to react, so there will be an optimum limestone size range based upon its properties and the CFBC operating conditions and system geometry.

REFERENCES

Not applicable.

APPENDIX A

COMBUSTION TEST FURNACE TEST EQUIPMENT

COMBUSTION TEST FURNACE TEST EQUIPMENT

COMBUSTION TEST FACILITY (CTF)

An isometric drawing of the Energy & Environmental Research Center (EERC) CTF is shown in Figure A-1. The furnace capacity is approximately 45 kg/hr (791,000 kJ/hr) of a fuel with a moderately high Btu content. The combustion chamber is 0.76 meters in diameter, 2.44 meters high, refractory-lined, and has been used for combustion testing of fuels of all ranks. The furnace can be configured in many different arrangements, and the graphic shows a second probe bank used for low-temperature ash-fouling evaluations. This section was replaced by a series of water-cooled, refractory-lined heat exchangers for the tests reported here.

The furnace diameter may be reduced to 0.66 meters to elevate the temperature entering the convective pass. Furnace exit gas temperatures (FEGTs) as high as 1093°C have been achieved during combustion testing in this mode. Most tests are performed using the standard configuration (0.76 meters inside diameter), with the FEGT maintained between 1093° and 1204°C during combustion tests utilizing low-rank coals. The FEGT was operated between 815° and 1120°C for combustion testing of the Tufanbeyli lignite. Two Type S thermocouples, located at the top of the combustion chamber, are used to monitor the FEGT. They are situated 180° apart at the midpoint of the transition from vertical to horizontal flow. Excess air levels are controlled manually by adjusting valves on the primary and secondary airstreams. The typical distribution is 15% primary and 85% secondary to achieve the specified excess air level. Tertiary air (or overfire air) may be added through two ports located just above the burner quarl.

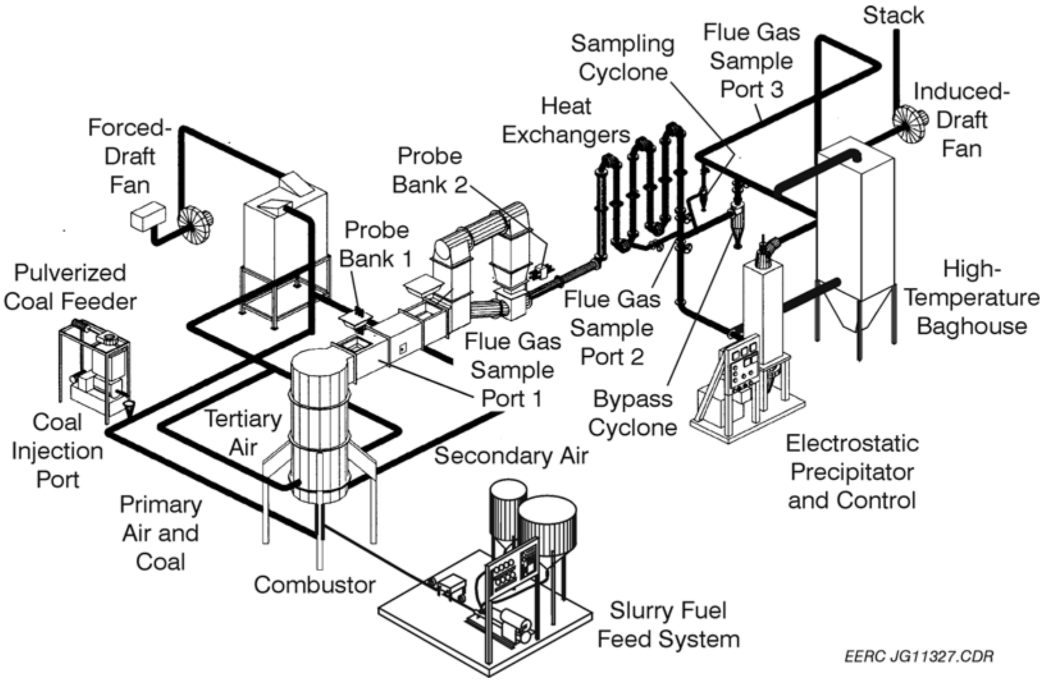

Figure A-1. CTF and auxiliary systems.

When solid fuels are fired, the fuel is normally pulverized remotely in a hammer mill pulverizer, targeted to a size of 70% less than 200 mesh. It is then charged to a microprocessor-controlled weight loss feeder from a transport hopper. Combustion air is preheated by an electric air heater. The pulverized fuel is screw-fed by the gravimetric feeder into the throat of a venturi section in the primary air line to the burner. Heated secondary air is introduced through an adjustable swirl burner, which uses only primary and secondary air. Flue gas passes out of the furnace into a 25-cm (10-inch)-square duct that is also refractory-lined. Located in the duct is a vertical probe bank designed to simulate superheater surfaces in a commercial boiler. After leaving the probe duct, the flue gas passes through a series of water-cooled, refractory-lined heat exchangers and a series of air-cooled heat exchangers before being discharged through either an electrostatic precipitator (ESP) or a baghouse. A baghouse was used for the tests performed here.

Flame Stability Testing

Flame stability is assessed by observation of the flame and its relation to the burner quarl as a function of secondary air swirl and operating conditions at full load and under turndown conditions. An International Flame Research Foundation (IFRF)-type adjustable secondary air swirl generator (shown in Figure A-2) uses primary and secondary air at approximately 15% and 85% of the total air, respectively, to adjust swirl between 0 and a maximum of 1.9. Swirl is defined as the ratio of the radial (tangential) momentum to axial momentum imparted to the secondary air by movable blocks internal to the burner and is used to set up an internal recirculation zone (IRZ) within the flame that allows greater mixing of combustion air and coal. Swirl is imparted by moving blocks to set up alternate paths of radial flow and tangential flow, creating a spin on the secondary airstream that increases the turbulence in the near-burner zone.

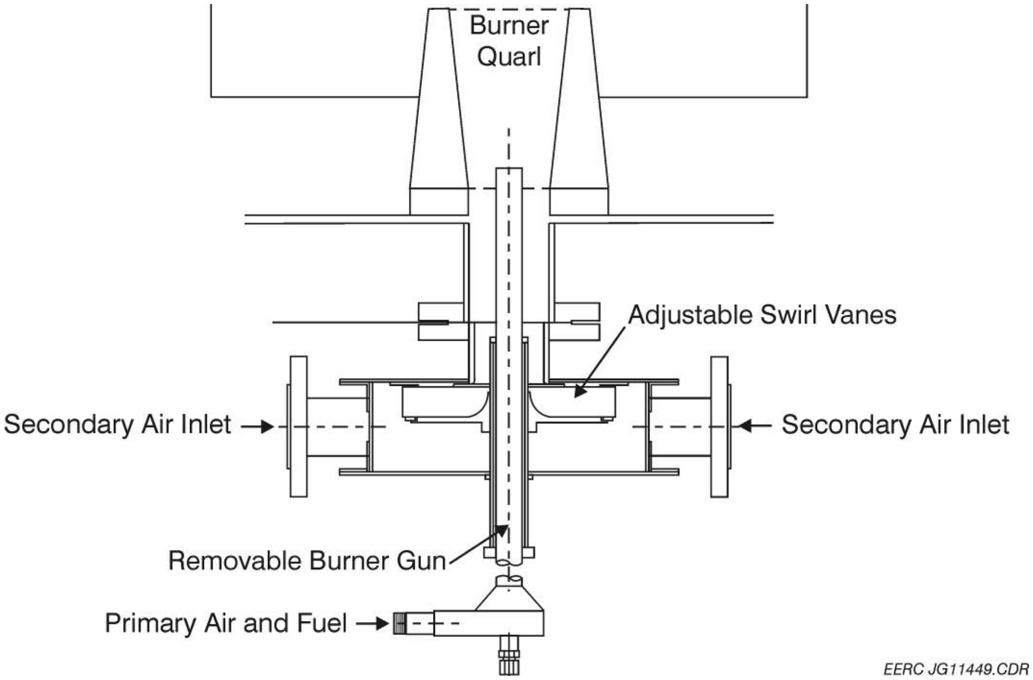

Figure A-2. IFRF adjustable swirl burner.

At the fully open position of the swirl block, the secondary air passes through the swirl burner unaffected, and the momentum of this stream has only an axial component when the air enters the burner region. It is the ratio of this radial component of the momentum to the axial component that establishes the quantity defined as swirl.

The adjustable swirl burner used by the EERC during flame stability testing consists of two annular plates and two series of interlocking wedge-shaped blocks, each attached to one of the plates. The two sets of blocks can form alternate radial and tangential flow channels, such that the airflow splits into an equal number of radial and tangential streams that combine further downstream into one swirling flow, as shown in Figure A-3. By a simple rotation of the movable plate, radial channels are progressively closed and tangential channels opened so that the resulting flux of angular momentum increases continuously between zero and a maximum value. This maximum swirl depends on the total airflow rate and the geometry of the swirl generator. Swirl can be calculated from the dimensions of the movable blocks (the ratio of the tangential and radial openings of the blocks) or from the measurement of the velocity of the airstream (obtaining both radial and axial components).

Secondary air swirl is used to stabilize the flame. In the absence of swirl, loss of flame may result, increasing the risk of dust explosion. As swirl is applied to the combustion air, fuel particles are entrained in the IRZ, increasing the heating rate of the particles, leading to the increased release of volatiles and char combustion. The flame becomes more compact and intense as swirl is increased to an optimum level, which is characterized in the EERC test facility. Swirl settings are made manually by moving a lever arm attached to the movable block

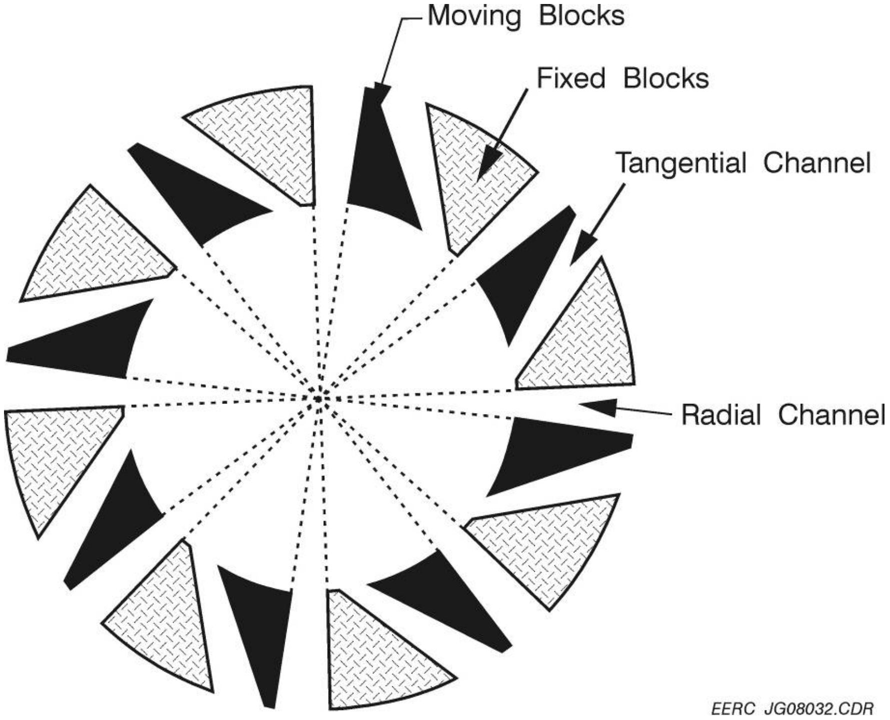

Figure A-3. Cross section of movable block assembly.

assembly. The position of the blocks is noted on a linear scale. The relationship between the swirl setting and the swirl number can be seen in Figure A-4.

Increasing swirl to provide flame stability and increased carbon conversion can also affect the formation of NO_x. The high flame temperatures and increased fuel–air mixing associated with increased swirl create an ideal situation under which NO_x may form. In full-scale burners with adjustable vanes, swirl is often increased to reach the optimum condition and then decreased slightly to reduce the production of NO_x.

Gas Analyses

The CTF utilizes two banks of Rosemount NGA gas analyzers to monitor O_2, CO, CO_2, and NO_x. Sulfur dioxide (SO_2) is monitored by analyzers manufactured by Ametek. The analyzers are typically located at the furnace exit and the particulate control device exit. The gas analyses are reported on a dry basis. Baldwin Environmental manufactures the flue gas conditioners used to remove water vapor from each gas sample. The flue gas constituents are constantly monitored and recorded by the CTF's data acquisition system.

Spray Dryer Adsorber

Because of the relatively high level of sulfur found in the Tufanbeyli coal, a spray dryer adsorber was added to the back end of the process to control sulfur emissions. The spray dryer was added to ensure the EERC complies with local environmental regulations throughout the testing of the coal. A solution of $CaCO_3$ was used for sulfur control. Gas concentration at the outlet of the spray dryer was monitored periodically using one of the gas analyzers and a three-way valve to switch input streams into the analyzer.

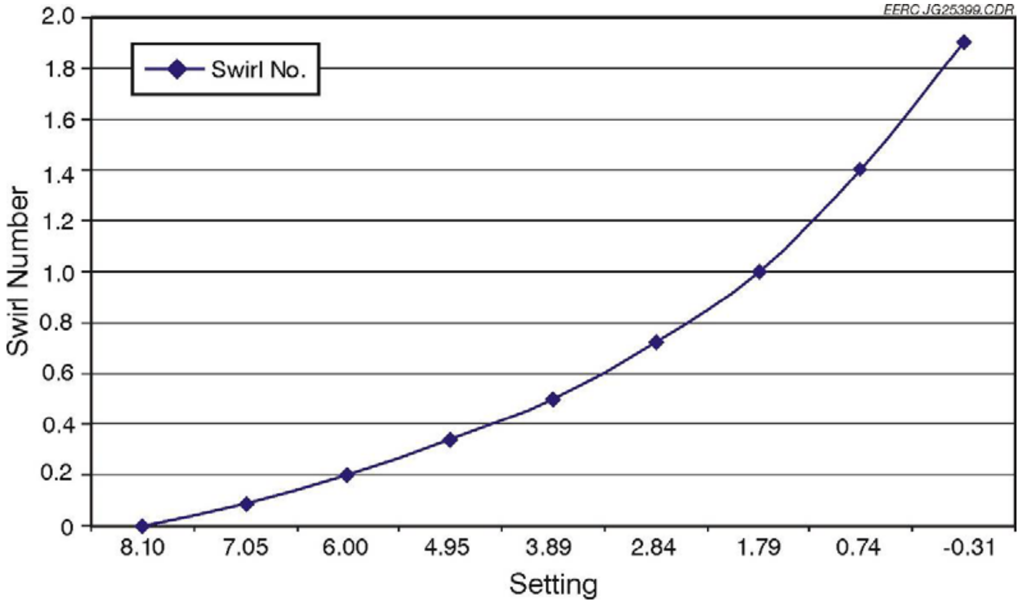

Figure A-4. Swirl number as a function of block setting.

APPENDIX B

ADDITIONAL DATA

Table B-1. Flue Gas Concentrations on a 6% O_2 Basis

Test: Date:	1A 1/23/2007	1B 1/24/2007	1C-1 1/24/2007	1C-2 1/24/2007	1D 1/25/2007	2A 1/24/2007	2B 1/24/2007	2C 1/25/2007
Start Time	16:10	13:30	15:54	16:06	11:13	17:07	18:55	8:50
Stop Time	18:16	14:09	16:04	17:04	12:43	18:43	19:45	9:30
Swirl Setting	3.2	3.2	2.0	3.2	3.2	3.2	3.2	3.2
Swirl Number	0.418	0.418	0.9	0.418	0.418	0.418	0.418	0.418
Fuel Feed Rate, lb/hr	149.6	150.2	151	150.4	151.4	151.7	145.9	148
Fuel Feed Rate, kg/hr	67.9	68.1	68.5	68.2	68.7	68.8	66.2	67.1
Fly Ash Carbon Content, %	0.44	0.82	NM**	0.52	NM	0.44	0.61	0.28
Temperature, °F (°C)								
Primary Air	700 (371)	705 (374)	700 (371)	708 (376)	696 (369)	702 (372)	709 (376)	713 (378)
Secondary Air	760 (404)	759 (404)	742 (394)	746 (397)	766 (408)	767 (409)	756 (395)	742 (395)
Furnace Exit	1906 (1041)	1882 (1028)	1854 (1012)	1851 (1011)	1881 (1027)	1834 (1001)	1820 (993)	1965 (1074)
Dry Gas Analysis—6% O_2 Basis								
Furnace Exit								
O_2, %	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
CO_2, %	13.3	13.7	13.6	13.4	12.8	13.3	13.6	12.9
CO, ppm	75.3	7.9	11.8	9.0	17.4	8.4	27.7	54.5
SO_2, ppm	5098	5346	5502	5455	5699	5469	5714	5686
NO_x, ppm	364.7	391.7	282.3	329.0	311.8	354.1	300.0	239.2
Baghouse Exit								
O_2, %	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
CO_2, %	13.1	13.3	NM	12.9	13.4	13.2	NM	13.3
CO, ppm	117.8	9.9	NM	11.8	24.1	10.1	NM	27.5
SO_2, ppm	>5000*	>5000*	NM	5828	6029	5703	5809.9	6081.3
NO_x, ppm	308.5	337.9	NM	260.2	284.6	329.3	NM	231.0
Spray Dryer Exit								
O_2, %	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A
CO_2, %	13.0	13.4	13.2	12.9	13.6	12.9	13.1	13.4
CO, ppm	140.4	10.4	23.7	11.8	21.8	10.4	54.9	172.0
SO_2, ppm	2705.1	2360.4	4319.7	4262.5	2104.7	4322.4	3830.2	3288.0
NO_x, ppm	315.1	337.8	233.4	259.3	282.1	295.2	224.7	245.6

* Over the detection limit of 5000 ppm for analyzer Mobile 1.

** Not measured.

Continued. . .

Table B-1. Flue Gas Concentrations on a 6% O_2 Basis (continued)

Test:	3A-1	3A-2	3B	3C	4C	5	6
Date:	1/25/2007	1/25/2007	1/25/2007	1/25/2007	1/25/2007	1/26/2007	2/1/2007
Start Time	16:40	17:26	17:15	18:31	19:06	13:52	9:30
Stop Time	17:17	17:45	18:18	18:46	19:17	17:34	10:09
Swirl Setting	0	3.2	3.2	3.2	3.2	3.2	3.2
Swirl Number	1.9	0.418	0.418	0.418	0.418	0.418	0.418
Fuel Feed Rate, lb/hr	79.8	79.5	80.4	82.3	83.6	142	147
Fuel Feed Rate, kg/hr	36.2	36.1	36.5	37.3	37.9	64.4	66.7
Fly Ash Carbon Content, %	0.57	0.57	0.71	0.60	NM**	NM	NM
Temperature, °F (°C)							
Primary Air	695 (368)	696 (369)	705 (374)	700 (371)	700 (371)	692 (367)	625 (330)
Secondary Air	717 (381)	718 (381)	687 (364)	641 (338)	630 (332)	752 (400)	690 (366)
Furnace Exit	1533 (834)	1551 (844)	1592 (867)	1600 (871)	1607 (875)	1923 (1051)	2040 (1116)
Dry Gas Analysis—6% O_2 Basis							
Furnace Exit							
O_2, %	N/A	N/A	N/A	N/A	N/A	N/A	N/A
CO_2, %	13.0	13.1	13.1	13.0	13.0	13.5	13.2
CO, ppm	37.0	35.3	33.6	30.2	29.1	43.4	16.1
SO_2, ppm	5338	5306	5345	5430	5533	5713	5795
NO_x, ppm	276.6	241.1	276.5	377.9	393.0	337.7	N/A
Baghouse Exit							
O_2, %	N/A	N/A	N/A	N/A	N/A	N/A	N/A
CO_2, %	13.5	13.4	13.2	13.0	12.9	13.0	13.1
CO, ppm	45.7	41.6	39.4	36.1	36.4	38.5	19.7
SO_2, ppm	>5000*	5470.0	5456.1	5878.2	>5000*	5490.1	>5000*
NO_x, ppm	256.2	222.2	254.3	348.0	359.1	341.6	463.5
Spray Dryer Exit							
O_2, %	N/A	N/A	N/A	N/A	N/A	N/A	N/A
CO_2, %	NM	NM	NM	NM	NM	13.2	NM
CO, ppm	NM	NM	NM	NM	NM	54.4	NM
SO_2, ppm	NM	NM	NM	NM	NM	3486.8	NM
NO_x, ppm	NM	NM	NM	NM	NM	358.8	NM

* Over the detection limit of 5000 ppm for analyzer Mobile 1.

** Not measured.

APPENDIX C

DESCRIPTION OF THE CIRCULATING FLUIDIZED-BED COMBUSTION (CFBC) SYSTEM

DESCRIPTION OF THE CIRCULATING FLUIDIZED-BED COMBUSTION (CFBC) SYSTEM

A cross-sectional view and a flow schematic of the Energy & Environmental Research Center (EERC) circulating fluidized-bed combustion (CFBC) system is shown in Figures C-1 and C-2. The overall system is divided into the following subsystems:

- Combustion air system
- Flue gas system
- Flue gas recirculation system
- Ash-fouling section
- Fuel and sorbent system
- Combustor
- Solids recirculation system
- Natural gas-fired preheater
- Combustor heat exchange system
- External heat exchange system
- Flue gas cooling water system

A forced-draft blower supplies combustion air and secondary air to the combustor. The combustion air heat exchanger is a shell-and-tube heat exchanger that uses hot flue gas to preheat the combustion air before it enters the combustor. Total combustion airflow is controlled by the amount of bypass through the combustion air bypass valve located directly after the combustion air heat exchanger. The secondary combustion air control valve determines the ratio of combustion air that enters the test furnace above the distributor plate to the amount of combustion air introduced into the combustor plenum below the distributor plate. The secondary combustion air can be introduced through manifolds at two different levels, located 5’9” and 10’6” above the distributor plate in Sections 2 and 3, respectively, of the combustor. Four 3-inch manual gate valves at each level are used to select where overfire air is introduced into the combustor.

Flue gas exits the top of the combustor, then flows through a refractory-lined primary cyclone with an inside diameter of 25 inches, the ash-fouling section, an air-cooled flue gas cooler, the combustion air heater, an 18-inch stainless steel secondary cyclone, the ten water-jacketed flue gas heat exchangers (modified for this testing with a hot-water recirculation system), and through either the flue gas bypass or the baghouse. Temperatures and pressures are monitored throughout the flue gas system. Flue gas is drawn through the induced-draft (ID) blower where it finally enters a stack for release to the atmosphere. Flue gas flow is controlled by the amount of air allowed into the ID blower through the ID fan bypass valve. The ID fan bypass valve is computer-controlled and continually adjusted to maintain −2 inches of water pressure at the inlet of the primary cyclone.

The flue gas recirculation blower is used to normally supply fluidizing air to the external heat exchanger, as was the case for this testing. It has the additional capability of supplying flue gas to the external heat exchanger (EHX) and also to the combustor if desired for flue gas

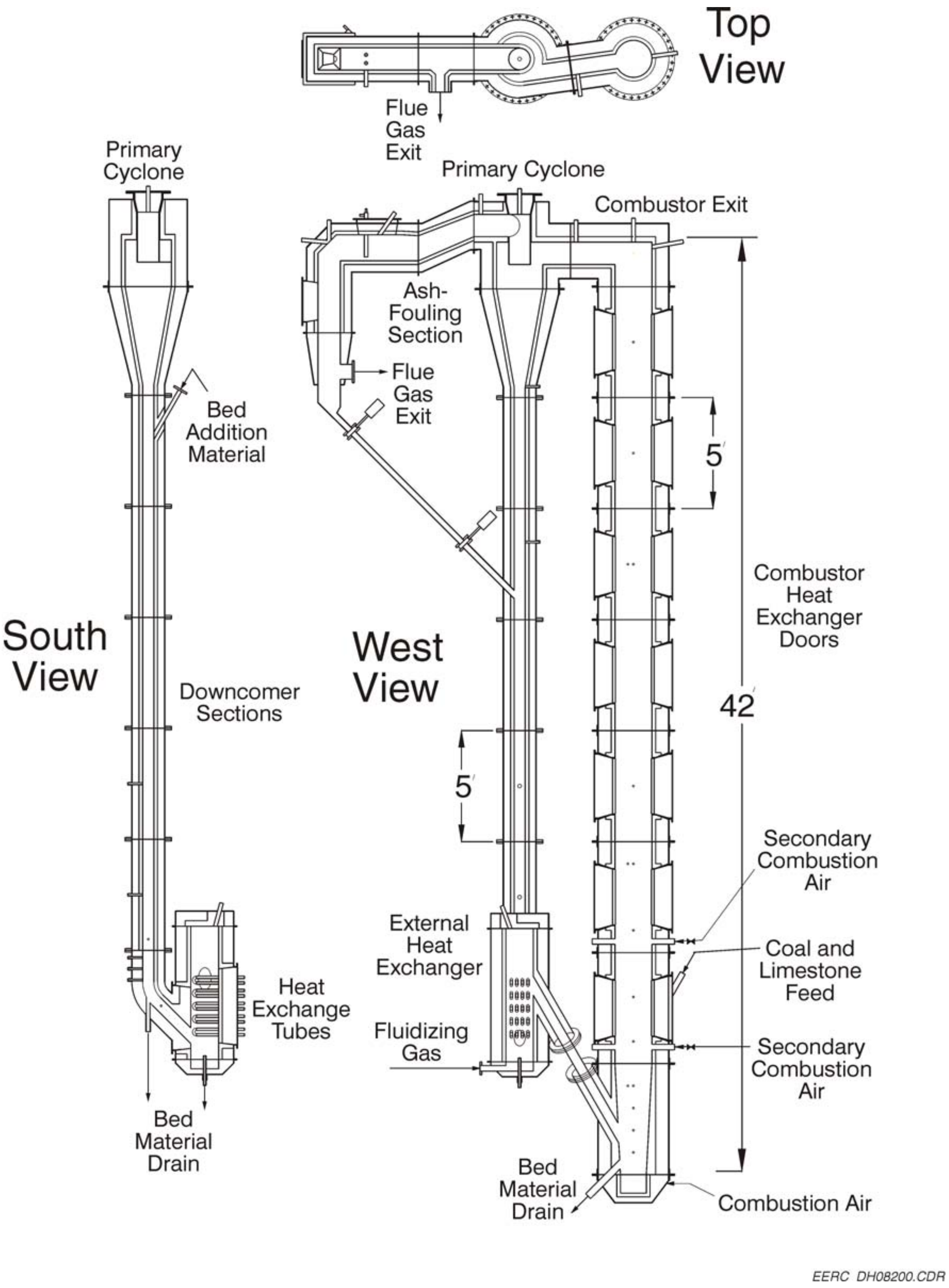

Figure C-1. Cross-sectional view of the EERC circulating fluidized-bed combustor.

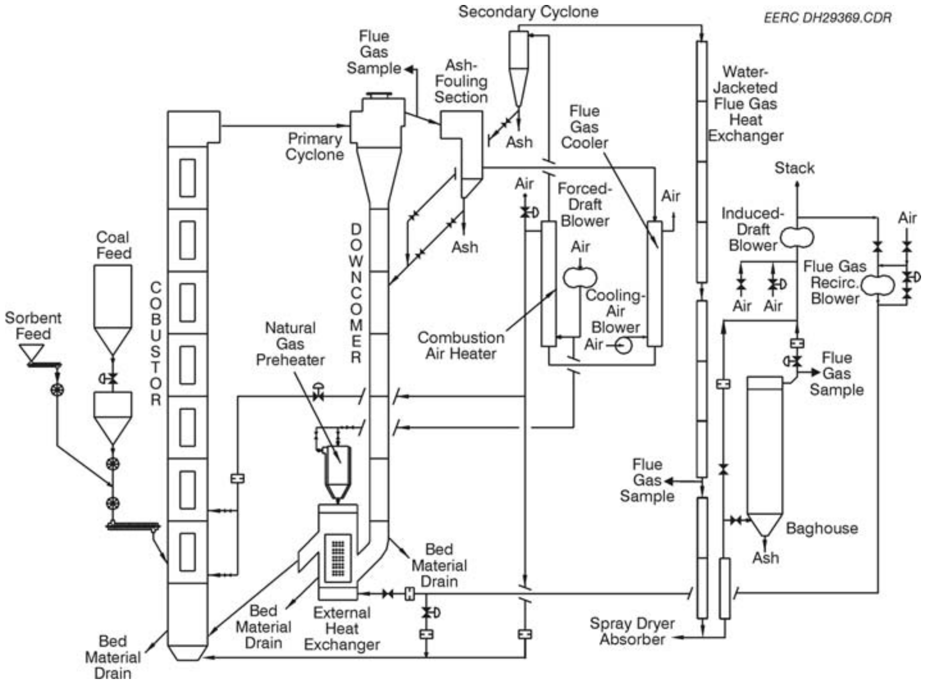

Figure C-2. Schematic of the EERC circulating fluidized-bed combustor.

recirculation testing. Manual gate valves located upstream of the blower allow either air or flue gas to enter the blower.

Primary and secondary combustion air, flue gas recirculation, and flue gas flow rates are measured using orifice plates. Instrumentation is interfaced with the data acquisition/control system to record and display the flow rates. Orifice differential and static pressures are also monitored with magnehelic pressure gauges.

The ash-fouling section is located at the exit of the 25-inch primary cyclone. Two air-cooled stainless steel probes maintained at 1000°F are present in the ash-fouling section to detect potential ash deposition or slagging. A hopper attached to the bottom of the ash-fouling section is connected to the downcomer via a drain leg containing two pneumatically actuated gate valves for ash recirculation. Three pneumatically actuated gate valves are used to allow the solids collected downstream by the secondary cyclone to be either routed back into the downcomer or to a collection barrel located on the ground floor. The length of time that any of these five pneumatic valves are open or closed is controlled with the data acquisition/control system.

The fuel storage hopper has a capacity of about 3000 pounds, which is transferred to a permanent feeder in approximately 600-pound increments. A gate valve is used to recharge the fuel feed hopper. The fuel feed hopper is suspended from a load cell; approximate fuel feed rates are calculated from the weight loss of the hopper over time. At the bottom of the weigh hopper, a rotary valve with an electronic speed controller is used to control the fuel feed rate.

The combustor is a series of refractory-lined sections bolted together. Each section has 2 inches of hard, abrasion-resistant refractory used in combination with 7 inches of insulating refractory. The bottom plenum section has the primary combustion air entrance and a bed material drain. The first combustor section (Section 1) has the solids recirculation return from the EHX. A removable stainless steel nozzle distributor plate is installed between the plenum and first combustor section. The next seven sections (Sections 2–8) each have two doorways on opposite sides for the installation of either blank refractory doors or heat exchange panels. At this time, twelve of the possible fourteen heat exchanger panels are installed in the combustor, two each in Sections 2, 3, 4, 7, and 8, and one each in Sections 5 and 6. Section 2 has the entrance for gravity feed of fuel and sorbent and the first set of secondary combustion air ports. Section 3 has the second set of four secondary combustion air ports. Section 9, the combustor exit, connects to the primary refractory-lined cyclone. Thermocouple and pressure taps are present in all of the combustor sections. All pressure taps are continuously purged with air to keep them open for accurate pressure measurements.

The refractory-lined components of the solids circulation system include the primary cyclone, the downcomer, and the EHX. Solids that are captured by the primary cyclone drop into the downcomer and travel downward into the EHX. Thermocouples monitor the temperature at the entrance and exit of the primary cyclone. The EHX has a plenum section into which either air or flue gas can be introduced. A removable stainless steel distributor plate is installed between the plenum and the main body of the EHX. The natural gas-fired preheater, described later, is attached to the top section of the EHX. Sixteen U-shaped stainless steel water-cooled heat exchanger tubes are installed in a removable refractory-lined door in the EHX. Thermocouple and pressure taps are distributed along the sections of the downcomer and in the EHX.

The preheater combustion chamber is constructed with inner and outer stainless steel shells. The natural gas-fired burner is bolted to the top of the preheater and fires downward. To maintain an acceptable operational temperature on the inside surface of the preheater, air is circulated through a baffled cooling jacket. Cooling air enters at the top of the preheater and flows downward, where it combines with the combustion gases at the bottom of the preheater transition cone. Preheater combustion air and the cooling jacket air are supplied by the forced-draft (FD) blower. A butterfly valve in the 4-inch supply line from the FD blower and a gate valve between the preheater and the EHX isolate the system when it is not being used. There are butterfly valves in the combustion air and cooling air lines for control purposes. There are also orifice plates in each line with magnehelics to monitor the flow rates. The flow of natural gas to the main and pilot burners are controlled with flowmeters located in the control room. A flame safety system is located in the control room to shut off the flow of natural gas to the preheater if 1) a flame is not present in the preheater, 2) combustion air is not being supplied to the preheater or cooling jacket, or 3) the combustion air pressure is greater than the natural gas pressure supplied to the preheater.

The rate of water flow to the combustor heat exchangers (CHX) is measured individually for each door by flowmeters and is controlled by globe valves installed above the flowmeters in the CHX panel boards. Total flow is measured with an in-line turbine flowmeter, which includes a bypass to allow for maintenance or repair during operations. An air system is connected to the inlet manifolds of each of the heat exchange panels. Air is used to cool the heat exchanger panels during operation prior to the introduction of water.

There are sixteen heat exchange coils available for water-cooling in the external heat exchanger door. Each U-shaped heat exchanger is constructed of 1-inch stainless steel pipe with ½-inch stainless steel tubing at each end. Each of eight circuits has a flowmeter and flow control valve mounted in a panel board to monitor and control the flow of water. Total flow is measured with an in-line turbine flowmeter, installed with a bypass to allow for maintenance or repair during operation. Flowmeters control flow to either single-circuit or dual-circuit sets of cooling coils with a thermocouple located at the exit of each circuit to measure the water exit temperature.

To eliminate or reduce the impact of any mercury reduction as a result of flue gas coming in contact with cold metal surfaces for this testing, the flue gas water-cooling system prior to the baghouse was modified. Previously water-cooling was with a once-through system using "city" water entering the heat exchangers at an average temperature of approximately 40° to 60°F, depending upon the time of year. The system was converted to a hot-water cooling system for the eight existing water-jacketed flue gas heat exchangers and the two newly installed water-jacketed flue gas heat exchangers. Water flow is supplied from an existing water pump connected to a water storage tank located on the 5th floor. Three manual control valves were installed for FGHXs 1–3, 4–6, and 7–10 to control the water flow rate to these three sets of heat exchangers. Fresh makeup water to the water storage tank, for controlling the inlet recirculation water temperature, is by use of two existing water flowmeters. A bypass from the pump to the tank is also available if required. For testing an inlet temperature of 190°F or higher to the heat exchangers, an exit temperature approaching 205°F was maintained to ensure no mercury condensation in the heat exchangers. Each set of heat exchangers was equipped with a relief valve that opens if the temperature exceeds 212°F.