

01-7993292 c.2  
CPUB

Energy, Mines and Resources Canada  
Energie, Mines et Ressources Canada

**CANMET**  
Canada Centre for Mineral and Energy Technology  
Centre canadien de la technologie des minéraux et de l'énergie

ERL 87-067 (OPJ) c.2  
87-67 (OPJ) c.2

COMBUSTION PERFORMANCE OF ONAKAWANA LIGNITE  
IN A PILOT SCALE FLUIDIZED BED COMBUSTOR

E.J. ANTHONY, D.L. DESAI, F.D. FRIEDRICH AND M. BEAL

March 1987

01-7993292 c.2

Presented at 37th Canadian Chemical Engineering Conference Montreal, Quebec, 18-22 May 1987.

Published in Conference Proceeding of the 37th Canadian Chemical Engineering Conference.

ENERGY RESEARCH LABORATORIES  
DIVISION REPORT ERL 87-67(OPJ)

MLU

COMBUSTION PERFORMANCE OF ONAKAWANA LIGNITE  
IN A PILOT SCALE FLUIDIZED BED COMBUSTOR

by

E.J. Anthony\*, D.L. Desai\*\*, F.D. Friedrich\* and M. Beal\*\*

ABSTRACT

Onakawana lignite containing up to 47% moisture has been burnt in two pilot scale atmospheric fluidized bed combustors. Bed parameters including fluidizing velocity, bed temperature and fuel moisture were varied to study their effect on combustion efficiency and emissions. Carbon combustion efficiency varied between 95 and 98% without recycle, the highest value being achieved above 900°C. With fly ash recycle, combustion efficiencies of >99% were achieved at temperatures as low as 800°C, providing that the ratio of "fines to lumps" in the coal feed stock did not exceed 2:1.

NO<sub>x</sub> emissions typically varied between 17 and 90 ng/J input, well within the limit of 258 ng/J input specified by the National Guidelines for Stationary Sources. On the other hand, SO<sub>x</sub>, varied from 270 to 1400 ng/J depending on operating conditions. The specified limit of 258 ng/J can be achieved by adding small amounts of limestone at about 850°C.

---

\*Research Scientist and \*\*Research Engineer, Combustion and Carbonization Research Laboratories, CANMET, Energy Mines and Resources Canada, Ottawa, Canada K1A 0G1.

## INTRODUCTION

Under a cost-shared agreement with Onakawana Development Limited, the Combustion and Carbonization Research Laboratory (CCRL) evaluated the feasibility of using lignite from the Onakawana deposit, located south of James Bay, as a fuel in utility boilers and industrial furnaces.

The joint project formed part of CANMET's Energy Program and was carried out in three phases. Phase I studies were conducted in CCRL's pilot-scale boiler under conditions representative of those in large, pulverized-fuel steam generators (1). Phases II and III parametric studies on combustion performance were carried out in the Mark I pilot-scale atmospheric fluidized bed combustor (AFBC) and the Mark II AFBC.

In 1975, CCRL initiated an R&D program on atmospheric fluidized bed combustion by erecting a small (240 mm diam) pilot-scale AFBC (Mark I). This combustor was used for a number of studies of fuel performance, but its design, configuration and size placed severe restrictions on the range of variables which could be examined.

As both the potential applications of FBC technology to the Canadian energy picture and the need for supporting R & D became apparent, it was decided to build a larger, more versatile pilot-scale combustor at CCRL, the Mark II. Both combustors have been described elsewhere (2,3).

This report compares, where possible, the Phase II and III studies, which were designed to evaluate the effect of parameters such as fluidizing velocity, bed temperature and fuel moisture on combustion performance. The effects of adding limestone and fly ash recycle were also investigated.

## EQUIPMENT, FEEDSTOCKS AND TEST PROGRAM

RESEARCH OBJECTIVES

Onakawana lignite has not yet been utilized on a commercial scale, hence information on its combustion performance is very limited. In the case of its utilization in a AFBC no prior work has been done therefore the main objective of these tests was to fill this knowledge gap.

It was agreed that as a first effort the following objectives should be addressed:

1. to determine the effect of bed temperature on combustion efficiency and emissions of  $\text{NO}_x$ ,  $\text{SO}_2$ , CO and particulates;
2. to determine the effect of coal moisture, if any, on combustion performance;
3. to determine whether ash particles tend to elutriate or remain in the bed;
4. to determine the effect of limestone addition on  $\text{SO}_2$  emissions; and
5. to study the effect of fly ash recycle on combustion performance and sulphur capture.

#### COAL AND LIMESTONE ANALYSIS

The Onakawana lignite used for the AFBC trials was part of a 15 tonne sample delivered to CCRL in sealed, plastic-lined drums. It had an "as received" moisture content of approximately 50% which made handling by mechanical equipment very difficult. This was alleviated by air-drying the coal to about 30% total moisture (including surface moisture of only 5%).

Average analytical data are given in Table 1. It should be noted that the ash is high in calcium, giving a natural Ca/S molar ratio of 1.38 for the fuel used in the Mark I trials, and a Ca/S molar ratio of 1.82 for coal used in the Mark II trials. Also most of the sulphur exists in an inorganic form, primarily as pyrite. The ash fusion temperatures are sufficiently high so that no slagging or sintering was expected over the normal temperature range employed in fluidized bed combustors ( $\sim 750$ - $1000^\circ\text{C}$ ).

The limestone used was Havelock limestone from New Brunswick. This has been shown to be moderately good sulphur sorbent and is the standard limestone employed at CCRL for FBC studies. It contains typically 95%  $\text{CaCO}_3$ . The limestone particle size was 6 mm x 0.

The bed material used for the tests was brown sand which had a particle density of  $2580 \text{ kg/m}^3$ , a bulk density of  $1570 \text{ kg/m}^3$  and voidage fraction measured at 0.39. The size range of the sand was 2 mm x 0 with a surface mean particle size of 1 mm.

#### MARK I COMBUSTOR

The Mark I pilot-scale AFBC facility has been described elsewhere (2). It consisted of a 1.2 m high combustor; a 1.73 m high freeboard section; a multi-cyclone dust collector, an electronic weigh-scale upon which sat a fuel hopper, screw feeders and a stack (Fig. 1).

The combustor consisted of a stainless steel shell surrounded by another cylinder of mild steel. The annulus between the cylinders formed a water jacket which absorbed some of the heat of combustion. As a further means of controlling the bed temperature, water could be injected directly into the bed.

The inside of the stainless steel shell was lined with a castable refractory insulation (Fiberfrax Variform B) which in turn was protected from abrasion due to the bed by means of an inconel liner. The final internal diameter of the combustor was 240 mm. The combustor and freeboard were penetrated by various ports to allow gas sampling and pressure and temperature measurements.

The distributor plate was comprised of a 6 mm thick stainless steel plate fitted with 36 stainless steel oil burner nozzles, which served as bubble caps. These were arranged in concentric circles of 63.5, 127 and 190.5 mm diam containing 6, 12 and 18 nozzles respectively. This configuration was found to give an even distribution of air and good fluidization.

The expanded freeboard section (500 mm in diam) was lined with insulating fire brick and attached to a conical section which tapered to the combustor diameter. The entire freeboard section was insulated by a fiberfrax blanket on the outside. The freeboard provided for some disengagement of elutriated material and allowed more residence time for secondary combustion.

The Mark I combustor was CCRL's first venture into FBC technology, and the following major limitations led to the design and installation of a larger, more versatile unit, the Mark II:

1. the heat input rate was restricted to about 400 MJ/h, too small for reliable extrapolation of results to industrial-size units;
2. the combustor design incorporated an inconel liner which had a service life of only 100 to 200 h;
3. the combustor was circular in section (which is not typical of full-scale AFBC equipment), and made probing of the bed and freeboard regions more difficult;
4. no in-bed cooling made it difficult to control bed temperature and impossible to measure bed-to-tube heat transfer coefficients;
5. the unit was not equipped with a fly ash recycle facility an important feature for maximizing combustion efficiency and sulphur capture.

## MARK II COMBUSTOR

The Mark II facility has been described elsewhere (3). It consists of a rectangular combustor (380 x 406 mm), a multicyclone dust collector with means for fly ash recycle, a gas-to-air heat exchanger downstream of the dust collector, a baghouse and a stack (Fig. 2). The fuel and limestone are screw fed into the bed from two hoppers each placed on an electronic weigh scale.

The combustor and freeboard consist of nine sections stacked one upon another and bolted together by means of flanges. The joints are sealed by fiberfrax rope gaskets and silicon caulking compound. The assembly is approximately 4.8 m high, has external dimensions of 0.94 x 0.97 m, and its internal cross-section is 0.154 m<sup>2</sup>.

The bed and freeboard are lined with an inner course of firebrick backed by two courses of high-density insulating brick. This brickwork is encased by a mild steel shell and a layer of insulating castable refractory which fills the space between the brickwork and the shell.

The distributor plate is comprised of a mild steel plate 380 mm<sup>2</sup> and 9.5 mm thick fitted with 100 bubble caps arranged in a 35 x 38 mm rectangular pitch. The bubble caps are comprised of stainless steel bolts. A hole drilled along its axis up to the bolt head with four holes drilled at right angles through the side of each bolt just below the head allows egress of the fluidizing air. The air nozzles are located about 20 mm above the plate, leaving a dead zone of non-fluidized bed material on the distributor plate in order to protect and insulate it from the combustion zone.

In-bed cooling is provided by a variable number (up to 48) of stainless steel "1/2 in." schedule 40 pipes 380 mm long. They can be connected in series or parallel circuits as desired by hoses, and water flow is measured in each circuit by rotameters.

The freeboard region extends above the active bed zone for about 2.2 m and provides a near adiabatic zone in which secondary combustion may go to completion. It also permits some of the elutriated solids to disengage from the gas stream. At the top of the freeboard two sections containing water-cooled heat exchangers reduce the flue gas temperature by 170-230°C. These were removed after the Onakawana lignite studies had been completed.

The instrumentation is standard and consists of NDIR analyzers to continuously measure CO, CO<sub>2</sub> and SO<sub>2</sub> concentrations, a paramagnetic oxygen analyzer and a chemiluminescent NO<sub>x</sub> analyzer. The temperatures in the bed,

freeboard and cooling circuits are measured by "K type" thermocouples; air and gas-side pressures are measured by means of manometers. With the exception of pressure measurements all data are recorded every five minutes by an electronic data logger.

#### TEST PROGRAM

Combustion efficiency, here expressed in terms of carbon loss or carryover, is the most important parameter for low-sulphur coals. It is strongly affected by bed temperature and to a lesser extent by fluidizing velocity. The test programs were planned accordingly with bed temperature as the primary input variable, ranging from 750 to 950°C in the Mark I tests and from 800 to 900°C in the Mark II tests. Fluidizing velocities were allowed to range from 2.3 to 3.8 m/s for the Mark I tests and were varied between 1.4 and 2.1 m/s for the Mark II tests.

As fuel moisture at the high levels typical of lignite is important, some tests were planned in which additional moisture was introduced at the feed screw to bring the moisture levels back to those representative of the "as received moisture" of 50%. However, all of the lignite had been previously air dried to 30% moisture in order to make it possible to handle and feed the material.

For the Mark I trials it was planned to hold the heat input constant, except for test 7 where at bed temperature of 950°C and a high moisture level in the coal, it was anticipated that a high coal feed rate would be required. For the Mark II trials, the effect of fines was studied since the fuel had been shown to be very friable. Tests 8 and 9 were performed with 4:1 and 2:1 ratios of "fines to lumps" in the feedstock, where fines are particles 2 mm x 0 and lumps are particles of 30 x 6 mm. For the Mark II and Mark I trials the fuel sizes were 30 mm x 0 and 6 mm x 0 respectively.

As past experience has shown excess air has relatively little effect on combustion performance except near stoichiometric conditions or at abnormally high levels. A target of 4% O<sub>2</sub> in the flue gas was chosen for all tests.

The test program was also planned to include trials with limestone to bring the total Ca/S molar ratio to 3. Unfortunately in the Mark II tests, it was discovered that Ca/S ratio in the fuel varied from 1.25 to 2.25 hence for the limestone tests the Ca/S ratio varied from 2.7 to 4.7. It is likely

that Ca/S ratios also varied in the Mark I tests, but it was not possible to carry out fuel analysis for each test. Therefore, for the Mark I tests, one sample was taken and analyzed and all subsequent samples were assumed to have the same S and Ca contents. The limestone used for these tests was Havelock, from New Brunswick, which contains 95%  $\text{CaCO}_3$  and has been shown to be a moderately good sulphur sorbent. The limestone particle size was 6 mm x 0. The matrices of target conditions for the Mark I and II studies are shown in Table 2.

## RESULTS AND DISCUSSION

Prior to each test, the combustors were brought as closely as possible to the specified operating conditions given in Table 2. Upon reaching equilibrium at those conditions, data were then taken during the "steady state" period only. On the whole, target conditions were maintained fairly closely, the greatest variations occurring in the fuel moisture content and excess air level. Thus steady state periods of 1 to 2 h proved adequate for the accumulation of representative data. The actual mean bed conditions and flue gas concentrations are given in Tables 3 and 4.

### FLUE GAS EMISSIONS

Concentrations of CO in the flue gases varied widely both in terms of the average concentration and in the degree of fluctuation and appear to be independent of fluidizing velocity and bed temperature. However, tests with fly ash recycle showed lower CO concentrations and less variation in the CO levels.

Test 8 of the Mark II trials, which was performed using coal with a 4:1 ratio of "fines to lumps", shows the highest CO concentrations. This suggests that the presence of fines and freeboard burning contributes significantly to overall CO emissions.

$\text{NO}_x$  concentrations from the Mark II combustor were, on average, higher than those from the Mark I. The  $\text{NO}_x$  levels observed with the Mark I increased with increasing bed temperature whereas no such clear correlation exists for the Mark II results.

The highest average concentrations of  $\text{NO}_x$  were 138 and 211 ppm for the Mark I and II combustors respectively, corresponding to 61 ng/J and



93 ng/J. These values compare favourably with the 258 ng/J allowed under the national guidelines (4).

For the Mark I combustor tests an apparent air leak of 5 to 10% occurred as the  $O_2$  and  $CO_2$  levels were too low and too high respectively when compared with the theoretical values calculated from the ultimate analysis, dry feed rate and measured air flow. The Mark II figures showed variations from theoretical values, but these were not statistically significant using a 95% confidence level. The results for  $SO_2$  are discussed later.

#### COMBUSTOR TEMPERATURE PROFILES AND FREEBOARD COMBUSTION

Temperatures in the bed and freeboard were monitored by thermocouples located at various heights above the distributor plate. The mean values for the Mark I and II units are given in Tables 5 and 6. Fly ash recycle seemed to enhance the degree of freeboard combustion.

Presumably the recycled unburnt fuel, which is light and friable, tends to burn in the freeboard. In addition, high velocities (greater than 3 m/s) seemed to reduce significantly the degree of temperature fluctuation (as shown by the standard deviations associated with mean freeboard temperatures). It is reasonable to suppose that this reflects the effects of increasing turbulent mixing in the flue gas. Calculations indicate that the flow should be turbulent in both units at velocities above 1.3 m/s.

#### CARBON CARRYOVER

One of the most important bed parameters is combustion efficiency. For fluidized bed systems, since all the volatile materials in the fuel are effectively burnt, we can discuss combustion efficiency in terms of the degree of unburnt carbon or carbon carryover, designated as fcc. Data for this parameter are presented in Tables 7 and 8 together with the values of the combustion parameters which might be expected to influence combustion performance, i.e., excess air level expressed as the mixture strength (which is the ratio of air used to that required for stoichiometric combustion); dimensionless fluidizing velocity, fuel moisture and bed temperature.

Attempts to correlate the bed parameters with combustion efficiency led to the following equation for the tests without fly ash recycle:

$$fcc = 0.452 Z^{0.4} X^{1.5} \exp \frac{3950}{T}$$

Multiple correlation coefficient:  $R^2 = 0.54$

where fcc is the carbon carryover; Z is the dimensionless fluidizing velocity defined as  $gL/U^2$ ; g is the acceleration due to gravity; L is the bed diameter or length; U is the superficial fluidizing velocity; X is the fuel moisture content expressed as a mass fraction; and T is the bed temperature (K). Although this correlation shows considerable scatter, its form is similar to that seen with other fuels (5,6). The values of fcc predicted by the equation are given in Tables 7 and 8.

Mixture strength appears to have no detectable effect while fluidizing velocity and moisture content appear to have comparable effects in the ranges in which they are varied. Bed temperature is, however, clearly the dominant parameter. Thus all the tests at about 950°C have fcc of <2% or combustion efficiencies of >98%. It should be noted that tests with fly ash recycle had combustion efficiencies >99% even though they were carried out as low as 800°C. Clearly fly ash recycle contributes to high combustion efficiency.

Test 8 of the Mark II trials produced a higher value for fcc than other tests carried out under similar conditions. This is almost certainly due to the relatively large proportion of fine material in the coal (4:1 "fines to lumps"). This conclusion is supported by the observation that this test was associated with the highest CO emissions in the Mark II trials, presumably due to freeboard combustion of fines.

It is also interesting to note that the results of test 9 in the Mark II series, with 2:1 "fines to lumps" give combustion efficiencies comparable to other tests, with a fcc of 2.46% whereas a 4:1 ratio (test 8) gave rise to a fcc of 4.97%; a variation of 100%. Although there is a temperature difference of a 100°C between these two tests, this cannot explain the much higher fcc. In particular, the lower fluidizing velocity in test 8 should have reduced the carryover. Also, tests 1 and 3 of the Mark II trials, which have similar moisture levels and fluidizing velocity, show a 10% variation in fcc for the same temperature difference. The much higher value for fcc must therefore be due to the high fines content of coal in test 8.

The distribution of unburnt carbon was also examined. Bed residue was found to contain <0.3% carbon, hence carbon loss is almost entirely due to elutriation and quenching of combustion after the freeboard. Typically, 84 to 96% of the elutriated carbon was trapped in the cyclone; fly ash recycle reduced the amount of carbon in the cyclone down to 54 to 78%, the rest of the carbon escaped the cyclone. Attempts to correlate bed parameters with the percentage of carbon trapped in the cyclone or escaping it, were unsuccessful as the data showed no regular pattern of dependence on moisture content, fluidizing velocity, bed temperature or limestone addition.

#### SULPHUR NEUTRALIZATION

The Onakawana lignite used in the Mark I test trials had a sulphur content of 1.3% on a dry basis and hence a natural Ca/S ratio of 1.38. Unfortunately no further measurements were made for this test series. For the Mark II tests, samples of the fuel were taken before each test, and the sulphur content was shown to vary from 0.8 to 1.44%, which meant that the natural Ca/S ratio ranged from 1.25-2.25 with an average of 1.82.

Table 9 gives the measured and theoretical SO<sub>2</sub> for the Mark I and II trials respectively. These tests show that the degree of sulphur capture by natural Ca components in the fly ash is inversely dependent on bed temperature and is best at 800°C and lower. This effect has been described before by Goblirsch et al. (7). The degree of sulphur captured by the natural Ca content in the ash can also be enhanced by fly ash recycle. For instance test 4 of the Mark II trials shows 79% SO<sub>2</sub> capture at a bed temperature of 800°C. Thus, recycle has the same effect as adding 3 to 4% by weight limestone per kilogram of fuel to bring the nominal Ca/S ratio to about 3:1.

Thus, there would appear to be two strategies to keep SO<sub>2</sub> emissions below 258 ng/J heat input. Operate at low temperatures (below 800°C) and employ fly ash recycle. Alternately, operate at 850°C with limestone. Fly ash recycle might still be used in order to help minimize the amount of sorbent required.

The National Guidelines for Stationary Sources states the following emission limits for SO<sub>2</sub>:

For sulphur dioxide - Generating units emitting more than 258 ng/J (0.6 lb/10<sup>6</sup> BTU) of heat input when uncontrolled.

- a) Those units emitting between 258 and 2580 ng/J (0.6 and 6.0 lb/10<sup>6</sup> BTU respectively) should be controlled such that the final emission does not exceed 258 ng/J (0.6 lb/10<sup>6</sup> BTU).
- b) Units emitting more than 2580 ng/J (6.0 lb/10<sup>6</sup> BTU) should be controlled so that a minimum of 90% of the uncontrolled emission is captured before release to the atmosphere.

Most of the tests performed had SO<sub>2</sub> concentrations above the required limit for stationary sources, i.e., 258 ng/J. However, two tests with limestone addition were performed where acceptable levels of 117 and 217 ng/J were achieved. The corresponding Ca/S ratios were 3.7 and 3.0 at a bed temperature of 850°C. It should be noted that this Ca/S ratio corresponds to a limestone feedrate of 7% or less of dry coal feed rate.

As part of the effort to carry out sulphur balances, the weight of the bed material for each test was determined along with its carbon and sulphur contents. This process demonstrated that there was no substantial accumulation of material in the bed for tests without limestone. For limestone tests, there was some accumulation of material in the bed, but this could be accounted for in terms of the weight of the limestone added with appropriate corrections for calcination and sulphation. This suggests that most or all of the ash component of Onakawana lignite elutriates.

#### CONCLUSIONS

Onakawana lignite is a reactive fuel which can be expected to burn well in an AFBC. Combustion efficiencies of 98% without fly ash recycle can be achieved at fluidizing velocities as high as 3 m/s providing the bed temperature is maintained at 950°C. Similar combustion efficiencies can be achieved at bed temperatures as low as 850°C providing the superficial velocity is 2 m/s or less, the fuel moisture is limited to about 30%, and the ratio of "fines to lumps" in the feedstock does not exceed 2:1. With fly ash recycle and a fluidizing velocity of 2 m/s, combustion efficiencies above 99% can be achieved with bed temperatures as low as 800°C.

The concentration of NO<sub>x</sub> in the flue gas fluctuated widely and seems to result from fluctuation in freeboard combustion. The highest levels encountered were 138 and 211 ppm from the Mark I and II combustors respectively,

which correspond to 61 ng/J and 93 ng/J. For both, this is well below the guidelines limit of 258 ng/J. It seems likely that the more uniform, elevated temperatures in the freeboard of the Mark II combustor are the cause of the consistently higher  $\text{NO}_x$  level found in that test rig.

Concentrations of CO fluctuated rapidly from several hundred to several thousand parts per million; greater variations were seen in the Mark I combustor, presumably due to greater fluctuation in freeboard combustion. The proportion of 4:1 "fines to lumps" produced the highest concentration of CO with a mean concentration of 2750 ppm. In a full-scale combustor, the CO concentrations could be minimized through appropriate freeboard design and adroit use of secondary air.

Onakawana lignite has a variable sulphur content, from 0.8 to 1.44% in the samples used for these tests. As a result, the natural Ca/S molar ratio varied from 1.25 to 2.25. Providing the bed temperatures were at or below 850°C, the calcium in the fuel appeared to capture about 50% of the potential  $\text{SO}_2$  emissions. Fly ash recycle improved the capture to 70-80% providing the bed temperature was maintained at 850°C or less, with the best capture occurring at 800°C or less.

However, neither low bed temperatures nor fly ash recycle reduces  $\text{SO}_2$  emissions to the 258 ng/J limit required by the national guidelines. This can only be achieved by adding limestone. In two tests, increasing the Ca/S molar ratio from approximately 1.3 to 3 and 1.82 to 2.67 gave a sulphur capture of 83 and 87%. The final  $\text{SO}_2$  emissions were then 218 and 117 ng/J respectively.

It is concluded that a full-scale AFBC would successfully burn Onakawana lignite. With a bed temperature of about 850°C, and fly ash recycle, combustion efficiencies would be 98 to 99%, and  $\text{NO}_x$  emissions would be well below the national guidelines. The required sulphur capture could be achieved by adding limestone at a rate of less than 10% of the weight of dry fuel. For optimum operation, the ratio of "fines to lumps" should be held at less than 2:1.

## REFERENCES

1. Prokopuk, R., Banks, G.N., Lee, G.K. and Whaley, H. "Pilot-scale trials with Onakawana lignite Phase I: Pulverized fired research boiler"; Division Report ERP/ERL 80-61(CF); Energy Research Laboratories, CANMET, Energy, Mines and Resources Canada, October 1980.
2. Anthony, E.J., Desai, D.L. and Friedrich, F.D. "The design and operation of the fluidized bed combustor developed at the Canadian Combustion Research Laboratory"; Division Report ERP/ERL 83-09(TR); Energy Research Laboratories, CANMET, Energy, Mines and Resources Canada, March 1982.
3. Hector, D.R., Desai, D.L., Friedrich, F.D. and Anthony, E.J. "Description of the Mark II combustor at the Canadian Combustion Research Laboratory"; Division Report ERP/ERL 81-55(TR); Energy Research Laboratories, CANMET, Energy, Mines and Resources Canada, August 1981.
4. "Thermal power generation emission - National Guidelines for New Stationary Sources"; Canada Gazette; April 25, 1981.
5. Anthony, E.J., Desai, D.L. and Friedrich, F.D. "The fluidized bed combustion of a medium volatile bituminous coal from British Columbia"; La Rivista Dei Combustibili XXXVIII:261-269; 1984.
6. Stover, N.S.H., Anthony, E.J., Desai, D.L. and Friedrich, F.D. "Combustion performance and sulphur capture of a western sub-bituminous coal and an eastern bituminous coal"; Proc ASME Winter Annual Meeting, Washington, DC; November 15-20, 1981.
7. Goblirsch, G., Vander Molen, R.H., Wilson, K. and Hajic, D. "Atmospheric fluidized bed combustion testing of North Dakota lignite"; 6th International Fluidized Bed Conference; Atlanta, Georgia; p 850-862; 1980.

Table 1 - Average analytical data for Onakawana lignite

Analysis			
Proximate (dry basis, wt %)			
Ash		24.47	
Volatile matter		38.59	
Fixed carbon		<u>36.94</u>	
		100.00	
Ultimate (dry basis, wt %)			
Carbon		52.62	
Hydrogen		3.78	
Sulphur		1.30,	[0.8 - 1.44 (Mk II tests)]
Nitrogen		0.78	
Ash		24.47	
Oxygen (by diff)		<u>17.05</u>	
		100.00	
Calorific value (MJ/kg)			
		20.18	
Hardgrove grindability index			
		46	
Moisture - "as received" wt %			
		31-38%	
		"as fired" wt %	31-47%
Ash fusion temperatures (°C)			
	Oxidizing	Reducing	
Initial deformation	1182	1149	
Spherical softening	1249	1232	
Hemispherical softening	1282	1249	
Fluid	1449	1393	
Ash composition (wt %)			
SiO <sub>2</sub>	44.95	SO <sub>3</sub>	13.33
Al <sub>2</sub> O <sub>3</sub>	11.47	Na <sub>2</sub> O	1.01
Fe <sub>2</sub> O <sub>3</sub>	7.93	K <sub>2</sub> O	1.07
TiO <sub>2</sub>	0.86	SrO	0.12
P <sub>2</sub> O <sub>5</sub>	0.28	BaO	0.23
CaO	12.85	Loss on firing	1.99
MgO	3.49		

Table 1 (cont'd)

Analysis	
Rank	Lignite A
Sulphur forms, % of total S	
Sulphate	9.6
Pyritic	72.3
Organic	18.1



Table 2 - Target operating conditions for Mark I and Mark II tests

Test No.	Heat input		Bed temp °C	O <sub>2</sub> in flue gas %	Fuel moisture %	Actual Ca/S mole ratio
	kW	MW/m <sup>2</sup>				
Mark I tests						
1	73.3	1.6	850	4	30*	1.38**
2	73.3	1.6	850	4	30*	1.38**
3	73.3	1.6	750	4	30*	1.38**
4	73.3	1.6	950	4	30*	1.38**
5	73.3	1.6	850	4	45	1.38**
6	73.3	1.6	750	4	45	1.38**
7	***	***	950	4	45	1.38**
8	73.3	1.6	850	4	30*	3.00†
Test No.	Fluidizing velocity		Bed temp °C	O <sub>2</sub> in flue gas %	Fuel moisture %	Actual Ca/S mole ratio
	m/s					
Mark II tests						
1	2.13		900	4	30*	2.22*
2	2.13		900	4	30*	2.25*
3	2.13		800	4	30*	2.25*
4	2.13		800	4	30*	1.40*
5	2.13		850	4	30*	3.67†
6	2.13		850	4	30*	2.69†
7	2.13		800	4	30*	4.67†
8	2.13		800	4	30*	1.25*
9	2.13		900	4	30*	1.56*

\* Air dried

\*\* Ca in ash

\*\*\* Unspecified in order to achieve high bed temperature with added moisture

† Ca in ash plus limestone

Table 3 - Mean bed operating conditions and flue gas concentrations for Mark 1 tests

Test No.	Steady state test length h	Fuel feed			Air rate m <sup>3</sup> /h**	Bed temp °C**	Flue gas analysis (dry gas basis)					Coal feed (dry basis) kg/m <sup>2</sup> h
		rate kg/h* (kW)	Fuel moisture (%)	Limestone feed rate kg/h			O <sub>2</sub> %**	CO <sub>2</sub> %**	SO <sub>2</sub> ppm**	NO <sub>x</sub> ppm**	CO ppm**	
1	2.0	16.92 (65)	31.4	0	93.3 (5.70)	814 (25)	5.80 (1.01)	14.0 (0.96)	663 (194)	130 (26)	571 (395)	254
2	1.5	19.33 (74)	31.4	0	96.7 (2.08)	850 (7)	4.00 (0.58)	14.50 (0.46)	701 (97)	93 (13)	921 (596)	291
3	1.25	19.04 (73)	31.4	0	94.9 (1.22)	757 (13)	4.20 (0.65)	15.20 (0.83)	532 (86)	70 (12)	1083 (680)	286
4	0.75	22.13 (77)	37.9	0	110.0 (0.66)	938 (6)	4.05 (0.30)	15.20 (0.49)	1288 (48)	88 (11)	2125 (629)	300
5	1.75	20.91 (74)	41.6††	0	100.4 (8.76)	841 (19)	3.91 (0.50)	14.70 (0.43)	1071 (138)	85 (35)	2425 (2724)	291
6	1.25	22.24 (74)	46.9††	0	87.9 (1.69)	754 (9)	4.15 (0.37)	15.27 (0.23)	608 (285)	43 (6)	1367 (634)	291
7	1.25	29.28 (104)	40.5††	0	131.6 (2.38)	943 (6)	3.92 (0.35)	15.48 (0.48)	1142 (86)	138 (19)	1383 (833)	407
8	2.25	23.38† (79)	37.0	0.95	100.5 (19.39)	831 (34)	4.24 (0.74)	14.68 (0.84)	229 (213)	71 (33)	2810 (327)	308

\* Fuel feed on air dried basis

\*\* Brackets indicate standard deviation

† Fuel is a coal-limestone mixture and the weight includes both

†† Test with moisture added at the feed screw

Table 4 - Mean bed operating conditions and flue gas concentrations for the Mark II test

Test No.	Steady state test length h	Fuel feed				Bed temp °C**	Flue gas analysis (dry gas basis)					Coal feed (dry basis) kg/m <sup>2</sup> h
		rate kg/h (kW)*	Fuel moisture (%)	Limestone feed rate kg/h	Air rate m <sup>3</sup> /h**		O <sub>2</sub> %**	CO <sub>2</sub> %**	SO <sub>2</sub> ppm**	NO <sub>x</sub> ppm**	CO ppm**	
1	2.23	63.0 (225)	36.3	0	291.8	900 (9)	4.23 (0.77)	16.36 (0.79)	804 (185)	211 (21)	655 (344)	260
2***	2.06	69.98 (250)	36.3	0	291.8	900 (10)	3.74 (0.67)	16.51 (0.82)	725 (179)	206 (25)	579 (416)	289
3	2.0	74.75 (282)	32.6	0	319.0	804 (6)	3.91 (0.65)	15.37 (0.68)	503 (125)	200 (8)	652 (373)	327
4***	1.98	74.72 (280)	32.6	0	319.0	800 (11)	4.13 (0.47)	14.19 (0.54)	316 (94)	211 (16)	460 (173)	324
5†	1.80	75.98 (284)	33.3	2.6	305.0	852 (5)	3.94 (0.45)	14.73 (1.58)	592 (193)	145 (20)	998 (609)	329
6††	1.97	70.86 (265)	33.3	2.6	305.0	847 (10)	3.66 (1.0)	15.03 (1.24)	378 (148)	130 (30)	610 (540)	306
7†	2.7	71.1 (185)	46.7	2.35	294.3	809 (12)	4.06 (0.92)	15.56 (1.37)	500 (203)	159 (25)	116 (1382)	246
8	2.33	54.55 (206)	32.6	0	205.2	807 (20)	5.04 (2.56)	19.53 (3.25)	752 (445)	160 (31)	2750 (2005)	238
9	2.53	63.53 (238)	33.2	0	281.6	904 (11)	4.72 (1.97)	14.59 (2.67)	809 (405)	202 (39)	825 (886)	275

\* Fuel feed on air dried basis

\*\* Brackets indicate standard deviation

\*\*\* Test with fly ash recycle

† Test with limestone

†† Test with limestone and fly ash recycle

Table 5 - Mean temperature profiles in the Mark I fluid bed combustor

Test No.	Height above distributor (m)					
	Bed		Freeboard			
	0.10	0.30	0.61	0.91	1.80	2.72
	Temperature °C*					
	T <sub>1</sub>	T <sub>2</sub> **	T <sub>3</sub>	T <sub>4</sub>	T <sub>5</sub>	T <sub>6</sub>
1	823 (24)	814 (25)	829 (26)	843 (25)	741 (24)	840 (48)
2	857 (7)	850 (7)	859 (6)	871 (6)	776 (11)	866 (15)
3	762 (21)	757 (13)	766 (12)	787 (12)	775 (5)	872 (9)
4	946 (6)	938 (6)	948 (6)	963 (6)	903 (6)	922 (6)
5	849 (19)	841 (19)	851 (18)	859 (18)	812 (11)	850 (19)
6	760 (9)	754 (9)	764 (11)	778 (8)	772 (6)	815 (4)
7	948 (6)	943 (6)	953 (6)	964 (6)	929 (8)	945 (8)
8	839 (36)	831 (34)	843 (39)	- -	775 (36)	783 (43)

\* Brackets indicate standard deviation

\*\* T<sub>2</sub> is quoted as the bed temperature

Table 6 - Mean temperature profiles in the Mark II fluid bed combustor

Test No.	Height above distributor (m)					
	Bed		Freeboard			
	0.13	0.32	0.61	0.87	1.47	2.44
	Temperature °C*					
	T <sub>1</sub>	T <sub>2</sub> **	T <sub>3</sub>	T <sub>4</sub>	T <sub>5</sub>	T <sub>6</sub>
1	890 (9)	899 (9)	916 (9)	919 (9)	904 (10)	878 (12)
2	892 (9)	899 (10)	914 (11)	920 (10)	923 (8)	910 (9)
3	795 (6)	804 (6)	822 (7)	830 (7)	828 (6)	819 (12)
4	794 (11)	799 (11)	813 (12)	821 (12)	836 (9)	843 (6)
5	844 (5)	852 (6)	867 (7)	874 (6)	868 (6)	852 (8)
6	843 (10)	846 (10)	861 (12)	866 (11)	877 (9)	875 (9)
7	796 (14)	809 (12)	825 (12)	832 (11)	829 (14)	820 (20)
8	793 (19)	807 (20)	825 (21)	831 (20)	815 (25)	782 (33)
9	891 (11)	904 (11)	920 (12)	921 (12)	907 (17)	866 (22)

\* Brackets indicate standard deviation

\*\* T<sub>2</sub> is quoted as the bed temperature

Table 7 - Carbon carryover and related bed parameters for Mark I tests

Test No.	Mixture strength*	Dimensionless fluidizing velocity		Moisture %	Bed temp °C	Carbon carryover %	Predicted carbon carryover %
		Z	(m/s)				
1	1.48	0.4451	(2.3)	31.4	814	4.83	2.20
2	1.34	0.3430	(2.62)	31.4	850	1.76	1.77
3	1.33	0.3649	(2.54)	31.4	757	2.75	2.48
4	1.47	0.2482	(3.08)	37.9	938	1.6	1.60
5	1.38	0.3303	(2.67)	41.6	841	5.92	2.73
6	1.21	0.4054	(2.41)	46.9	754	4.22	4.78
7	1.30	0.1648	(3.78)	40.5	943	1.45	1.48
8	1.30	0.3510	(2.59)	37.0	831	4.76	2.42

\* Mixture strength =  $\frac{\text{total air supplied}}{\text{stoichiometric requirement}}$

Table 8 - Carbon carryover and related bed parameters for Mark II tests

Test No.	Mixture strength*	Dimensionless fluidizing velocity		Moisture %	Bed temp °C	Carbon carryover %	Predicted carbon carryover %
		Z	(m/s)				
1	1.57	0.8779	(2.13)	36.3	899	2.05	2.73
2**	1.27	0.8779	(2.13)	36.3	899	0.45	-
3	1.23	0.8779	(2.13)	32.6	804	1.85	3.13
4**	1.22	0.8779	(2.13)	32.6	799	0.75	-
5	1.14	0.8779	(2.13)	33.3	852	1.92	2.76
6**	1.24	0.8779	(2.13)	33.3	847	0.58	-
7	2.02	1.0236	(1.97)	46.7	809	5.61	5.60
8	1.08	2.1220	(1.37)	32.6	807	4.97	4.37
9	1.33	0.9386	(2.06)	33.2	904	2.46	2.42

\* Mixture strength =  $\frac{\text{total air supplied}}{\text{stoichiometric requirement}}$

\*\* Tests with fly ash recycle

Table 9 - Comparison of measured and theoretical SO<sub>2</sub> concentrations in the flue gas

Test No.	Measured SO <sub>2</sub>		Theoretical SO <sub>2</sub> ppm	Ratio of measured to theoretical		SO <sub>2</sub> retained %	Bed temp °C
	ppm	ng/J		%	%		
Mark I trials							
1	663	716	1206	55	45	814	
2	701	686	1332	53	47	850	
3	532	519	1337	40	60	757	
4	1288	1383	1214	106	0	938	
5	1071	1086	1285	83	17	841	
6	608	537	1477	41	59	754	
7	1142	1085	1373	83	17	943	
8*	229	218	1363	17	83	831	
Mark II trials							
1	804	788	827	97	3	899	
2**	725	640	909	80	20	899	
3	502	429	940	53	47	804	
4**	316	272	1494	21	79	798	
5*	145	117	1139	13	87	852	
6†	378	344	1499	25	75	847	
7*	500	568	764	65	35	809	
8	752	566	1068	70	30	807	
9	809	724	1288	63	37	904	

\* Tests with added limestone

\*\* Tests with fly ash recycle

† Test with limestone and fly ash recycle



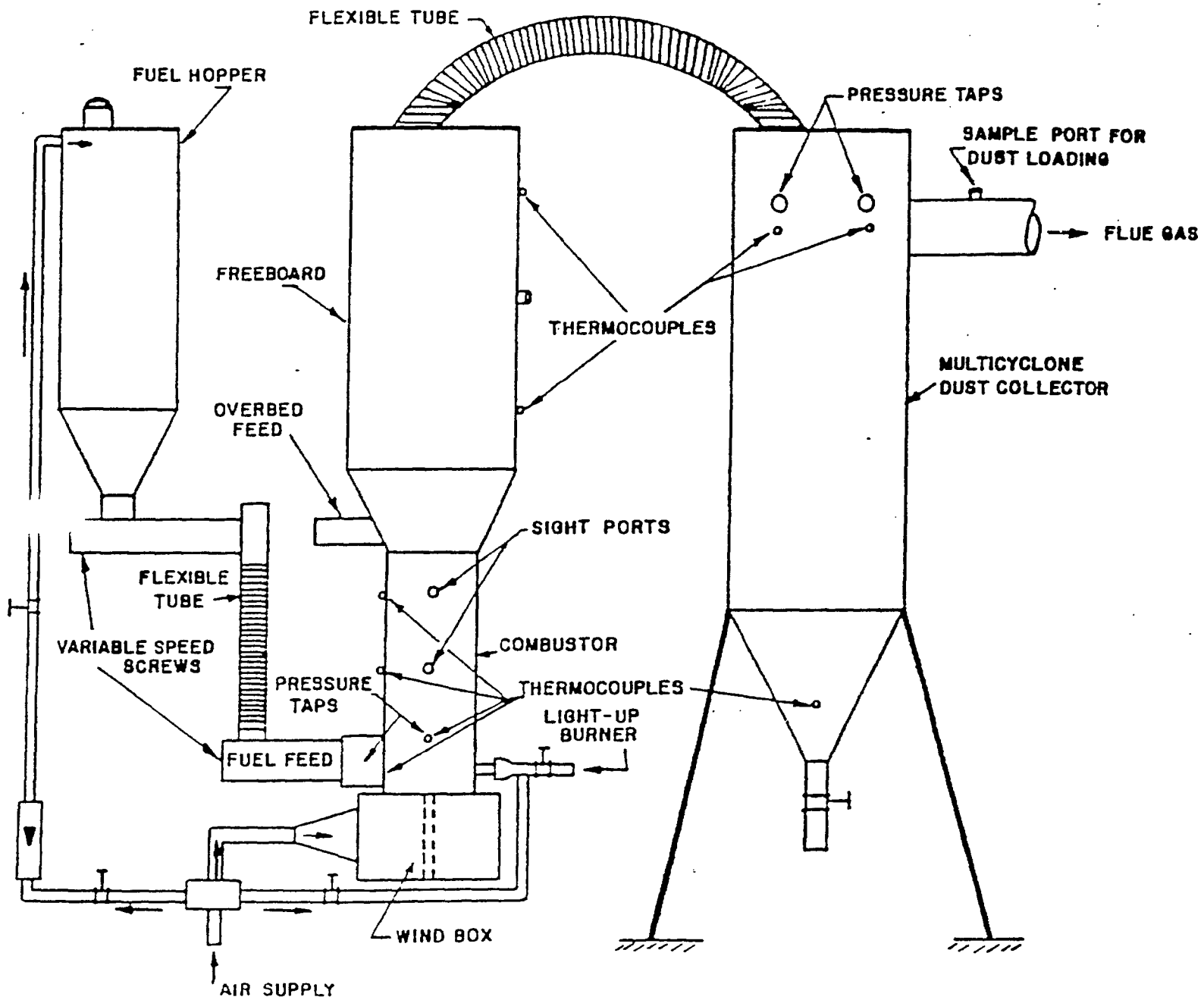


Fig. 1 Schematic Of Mark I Fluidized Bed Combustor

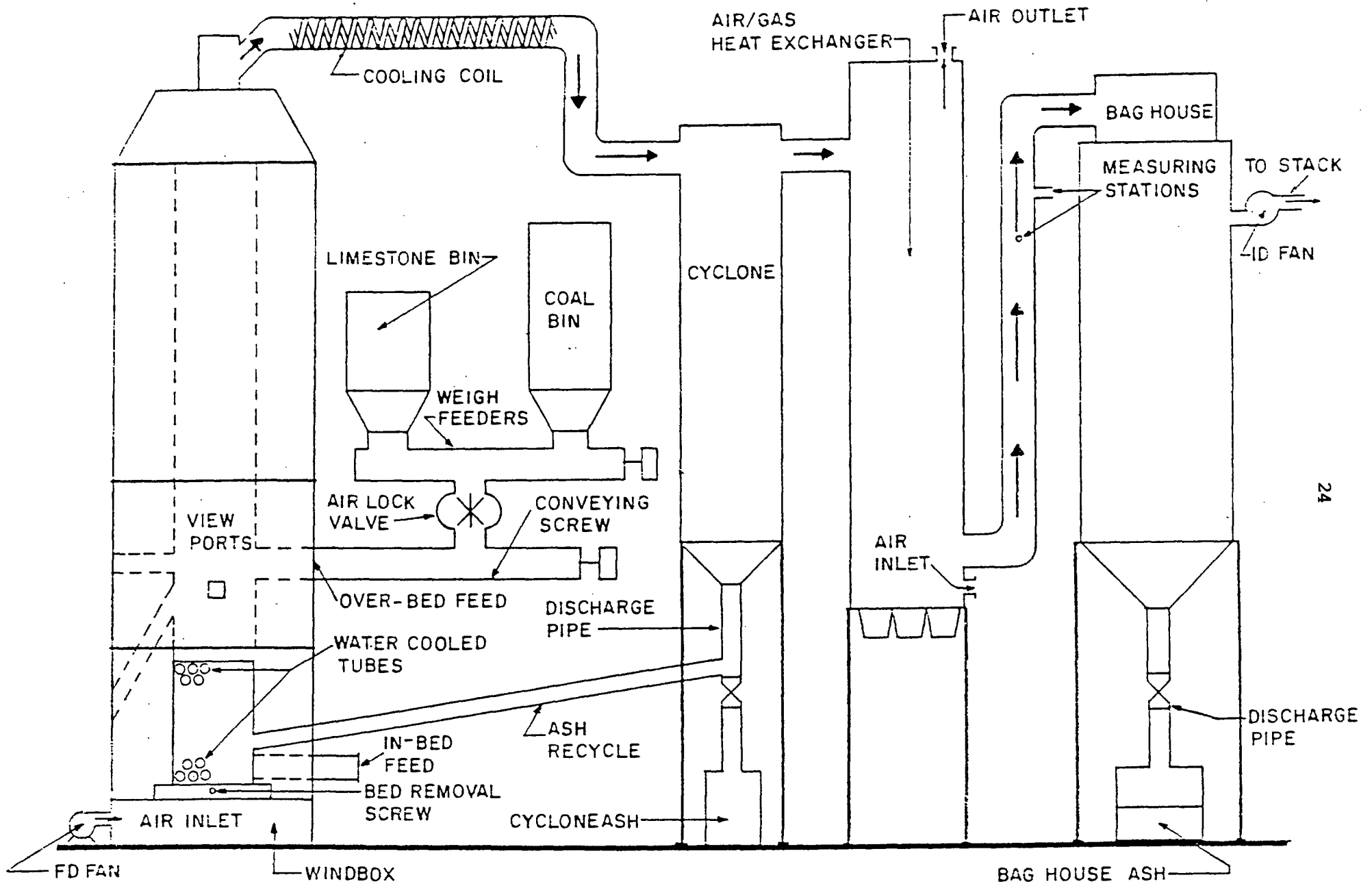


Fig. 2 Schematic Of Mark II Fluidized Bed Combustor