DEVELOPMENT OF FLUIDIZED BED COMBUSTION SYSTEM FOR LIGNITE-FIRED INDUSTRIAL BOILERS

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ABSTRACT

A project has been initiated to design and build a multifuel fluidized bed boiler for small scale industrial use. A differentially fluidized sand bed furnace employing a new type of air distributor was incorporated into a 2 MW_{th} conventional D-type water-tube boiler. The system was designed to combust lignite as well as other low-grade fuels efficiently at low combustion temperature between 800 and 900 ^{o}C such that NO_x formation is negligible. An acceptable emission of SO_2 and particulates were controlled by an in situ application of lime stone and by a utilization of a special mechanically induced water-spray scrubber respectively.

The prototype has been installed, test-run, and demonstrated at the Royal Food Processing Plant in Chiang Rai. A study of the system performance and its emission characteristics and controls was then made using lignite with different sizes : -2" and -¾ ". It was observed that for the larger size of lignite, the degree of bed mixing near the feed point was significantly reduced, resulting in a high bed temperature at that region. At a fuel feed rate of 500 kg/h the maximum steam capacity was 3.5 t/h, and the maximum thermal efficiency of the boiler was found to be 80%. The heat transfer coefficients of in-bed and over-bed tubes were in the range of 400-800 and 10-100 W/sq.m ^{o}C respectively. The pollution control system performed properly giving emissions at acceptable levels. The average particulate collection efficiencies of the cyclone and the wet scrubber was 50 and 85% respectively. The size and concentration of particulates in the flue gas discharged to the atmosphere was 3-7 µm and 0.03-5 g/Nm^3, respectively. It was also found that about 20% of SO_2 in the flue gas exiting the boiler was absorbed by water spray in the wet scrubber. Without the addition of limestone, the exhausted SO_2 concentration was in the range of 0.05-0.08% by volume when burning lignite containing 1.2% sulphur. This emission level was reduced well below the acceptable urban standard of 0.04% by volume by adding in limestone using Ca/S ratios between 2-3.

INTRODUCTION

In developing countries such as Thailand, growth of various kinds of industries is quite extensive. Most of these industries rely on fossil fuels in the form of fuel oil, lignite, natural gas and diesel oil for their energy requirements. For a combination of economic and technical reasons, fuel oil and diesel oil are used extensively for energy generation in small industries. These small industries are thus one of the main consumers of petroleum-based fuels which are not available in the country and must always be imported.

Thailand has significant lignite reserves and biomass supply sufficient to meet the energy needs of a large number of medium and small industries. It is thus one of the major goals in the government's industrial development plan to promote the use of lignite and biomass as substitute fuels for petroleum-based fuels. This would help conserve foreign exchange by reducing imports of liquid fuels and crude oils.

OBJECTIVE

Pilot Plant Development and Training Institute or PDTI, an autonomous unit within the King Mongkut's Institute of Technology Thonburi mandated to conduct research, design and development of timely and indigenous technology based on industry needs, has been engaged in the R & D work on finding a suitable and efficient energy generation system capable of utilizing lignite and biomass which is technically and economically feasible as well as being environmentally acceptable. This leads to the development of a Fluidized-bed combustion furnace for small-scale industrial use under the financial support of the Australian Government. Fluidized bed combustion technology is applicable to burning various types of low-grade fuels including lignite/coals with a wide range of sulfur content without the need for stack scrubbers for removal of sulfur dioxide and nitrogen oxides. High heat transfer rate and combustion efficiency as well as minimal gaseous pollutant emission can be maintained by this combustion system.

FLUIDIZATION PRINCIPLE (1,2)

The fluidized bed is simply a volume of inert particles such as sand, ash or limestone which are supported by a grate-like air distributor. When air is blown through the bed mass at sufficiently high velocity, the solid particles are lifted and suspended by the air. At this point the particles can move freely, and the bed behaves like a fluid. This leads to well mixed of fuel particles and high heat transfer rate between bed and surface.

For combustion application, the bed is preheated by a start up burner. After the ignition temperature is reached, the fuel such as coal, biomass or solid waste is then continuously introduced to provide self-sustaining combustion. At any given time, the bed material contains only 1 to 5 percent of combustible material. The heat released can be used for steam and/or power generation

For coal combustion. limestone or dolomite is usually added into the bed to capture sulfur dioxide formed reducing its emission level well below the acceptable standard. A combustion efficiency up to 95% or more can be achieved in a low temperature range of 750-900 Celsius which also minimizes both ash agglomeration and nitrogen oxide emission.

FLUIDIZED BED BOILER DEVELOPMENT

With the advantages of fluidized-bed combustion technology, research and development on this technique has thus been carried out by ASEAN and Thai researchers for the past ten years [3-9]. The results is promissing for a larger scale application, but as yet there has been no attempt outside of the laboratories. Based on these bench-scale knowledges and experiences, and the Australian expertise, the Department of Chemical Engineering and PDTI of King Mongkut's Institute of Technology Thonburi has been able to design and scale up the fluidized bed combustion system into a prototype for practical demonstration. The unit was designed and incorporated into a standard water-tube boiler with 3 tons steam per hour capacity which has been installed and test-run at the Royal Food Processing Plant in Chiang Rai [10-14]. This functional prototype has demonstrated the technical and economical feasibility as well as environmental acceptability of the fluidized bed energy generation system using lignite and biomass as substitutes for the petroleum-based fuels.

BOILER PLANT

The overall boiler plant is shown schematically in Figure 1. The major components of the system are as follows.

Main Boiler

This was a standard 2 MW D-type water tube boiler with the radiant section and bottom stoker being replaced with a fluidizing chamber. The combustion chamber had a dimension of 1.25 x 2.6 m^2 in cross-section and contained a sand bed of 0.30 m in static height. The boiler was designed to produce steam at the rated capacity of 3 ton/h and at pressure of 10 bar gauge.

Fuel Feeding

A magnetic shaker was used to feed lignite coal from a storage bin onto the bed surface. The feeding chute is about 0.60 m above the furnace floor.

Air Moving System

A forced-draft blower (3000 scfm, 40 in. H_2O) was used to supply air for fluidization and combustion and an induced-draft blower (3500 scfm, -16 in. H_2O) helped draw the combustion gas from the freeboard and maintain a negative pressure of about 1 in. H_2O. Air was uniformly distributed into the bed via an air distributor of sparger type. Figure 2 shows details of the distributor.

Entrained Particulates Control

Particles which were entrained with the combustion gases were removed consecutively in the following units; convective tubes section, a dry cyclone, and a mechanical spray scrubber. The wet gas from the scrubber passed through a knock-out drum for droplet separation. Figure 3 depicts the cyclone and the scrubber devices. Also, the scrubbing unit was able to absorb part of SO_2 gas generated from the combustion of fed coal.

TEST PROCEDURE

The fluidized bed furnace was initially preheated by the LPG premixed gas method with the aid of an overbed pilot gas burner. The fluidizing air velocity used during bed heat-up was U - 1.5 U_{mf}, which was found to give the least LPG consumption [11]. When the bed was preheated up to 350^oC, the fuel was slowly added and maintained at the desired feed rate. As the bed temperature went up to

600^oC, the main gas supply was stopped. The required air flow rate was adjusted and the system was allowed to attain steady condition, as indicated by a constant bed temperature. Combustion gas temperature at various locations along the flow path were continuously recorded. Flue gas was analyzed for % CO and carbon loss with entrained particulates determined for computing the combustion efficiency. In addition, the inbed and overbed water tubes were installed at positions 20 and 80 cm above the air distributor and temperature and flow of water were recorded for determining heat transfer coefficients. Tables 1 and 2 summarize the properties of fuels and bed material used in this work.

For emission study, steady conditions at constant bed temperature (800-900^oC, depending on A/F ratio) were maintained. A sampling device as shown in Figure 4 was used to draw gas samples at various positions prior to entering the dry cyclone and after leaving the knock-out drum (see Fig. 3). The sampling time was between 3-5 minutes and the suction rate of the vacuum pump was 4.1 m^3/h. Gas detector tubes were used for analyzing the compositions of O_2, CO_2, CO and SO_2 in the gas samples. The amount of particles collected on the filter paper was determined and the particle size distribution measured by screen analysis for coarse fractions and SEM analysis for fine size ranges. In addition, ash removal rate from the bottom of the convective tube bank, cyclone and knock-out drum underflow were measured.

During the SO_2 removal study, limestone was added continuously with lignite varying Ca/S ratios between 0 to 4. Gas samples were then taken and analyzed as already indicated.

RESULTS AND DISCUSSION

Bed Preheating

Figure 5 compares the rising bed temperature (T_3) during the warm-up period, using lignite feed size of -2" and -3/4". It is noted that the preheating time to attain a constant bed temperature of 800^oC is shorter for the -3/4" feed than the -2" feed (75 min. vs 90 min.). This results from the fact that the smaller size feed has a higher burning rate, due to its larger surface area of contact.

Boiler Performance

Figure 6 shows the effect of A/F ratio on the steam production rate. For each running condition, the curve exhibits a maximum steam output near the stoichiometric ratio. The drop in steam production is primarily due to increased heat loss with flue gas at higher air flow rate. For -¾" coal feed, supplying coal at a higher rate gives higher steam production with maximum steam production occurring at 3.7 and 3.3 t/h for coal feed rates at 506 and 435 kg/h, respectively. It is also clear that coal feed size of -2" gives less steam generation than the -¾" feed size, due to its lower burning rate of coal particles in the bed. This result indicates the need to have proper fuel feed size for efficient fuel combustion.

Figure 7 shows the effect of A/F ratio on the thermal efficiency of the boiler. A similar trend is observed as in the case of steam production, as expected. Results of combustion efficiency, which measures the completeness of fuel combustion, are shown in Figure 8. It was found that the combustion efficiency is not a strong function of A/F ratio with the value varying from 75-98% for the two feed rates of -¼" lignite.

Figure 9 shows variation of gas temperature with A/F ratio at various locations of the boiler system; in the bed, above the fluidizing bed, leaving the convective tube bank (flue gas), and leaving the emission control system (stack gas). In general, temperature at each location is relatively uniform, almost independent of varying A/F ratio. It is also noted that for lignite combustion, temperature inside the bed is slightly higher than the freeboard temperature, showing less volatile combustion in the freeboard region.

Heat transfer coefficients of inbed and overbed water tubes were estimated and found to correlate reasonably well with air velocity normalized with minimum fluidizing velocity, U/U_{mf}, in the range of 1.5-4.5. There is a tendency for the coefficient to decrease at a high value of U/U_{mf}. Values of inbed and overbed heat transfer coefficients vary over the range 400-800 and 10-100 $W/m^2\ ^oC$, respectively, depending on U/U_{mf},, and lignite feed rate.

Gaseous Emission

Analyses of CO_2, CO and O_2 composition in the flue gas are presented as a function of air/fuel ratio in Figure 10 and tabulated in Table 3. In general, there is a slight dependence of % CO_2 on the A/F ratio with a maximum gas concentration occuring near the stoichiometric A/F ratio. This maximum concentration was found to be 12.5% and 10% for lignite 1 and lignite 2, respectively. Calculation based on ultimate analysis data showed that for stoichiometric combustion the mean % CO_2 for these fuels was in the order of 16% volume. The variation of O_2 concentration appears to show opposite effect as compared to that of CO_2, as would be expected. The level of CO is considered to be relatively small, indicating almost complete combustion of fuel in the fluidized bed.

The variation of SO_2 in the flue gas with A/F ratio is shown in Figure 11 for the two coals. Again, SO_2 concentration passes through a maximum (~ 1200 ppm) at optimum A/F ratio, corresponding to maximum CO_2 production. SO_2 level for the gas leaving the mechanical wet scrubber is also shown in the same figure for comparison. On the average, the reduction of SO_2 by water spray in the scrubber is about 25%. This gives an SO_2 concentration in the stack gas ranging from 0.05 - 0.08%. These values are close to the standard emission level of 700 ppm (0.07 vol.%) as set by the National Environment Board (NEB) for provincial areas. To have better control of released SO_2, it is necessary to utilize limestone sorbent for inbed capture of this gas.

The results of SO_2 inbed capture test runs using limestone with Ca/S ratios between 0 and 4 are given in Table 5 and Figure 12. The SO2 emission level was reduced well below the NEB urban standard of 0.04 vol% when using Ca/S ratios equal or greater than 2. The overall capture efficiency of limestone and waterspray was founded to be between 60 to 90%.

Particulates Emission

Figure 13 shows the effect of fluidizing air velocity, expressed in terms of U/U_{mf}, on the flow of ash particles from the bottom of the convective tubes, and from the cyclone underflow and knock-out drum. It appears that the removal rate of entrained ash particles from the boiler is not so sensitive to air velocity up to U/U_{mf} 4.5. The results also show that ash particles are collected mostly by in the convective part, followed by the dry cyclone and the wet scrubber. Table 4 compares the average ash removal rate and the input rate of ash with fuel feed. The difference between the input and output rates of ash flow is not great except for lignite 2. The performance of dust collectors is displayed in Figure 14. For both the dry cyclone and wet scrubber, the collection efficiencies increase with air velocity up to $U/U_{mf} \simeq 3.6$ and remain almost constant at higher velocity. For dry cyclone the efficiency varies from 30-65%, while for the wet scrubber it varies from 60-90%.

Figure 15 shows a typical size distribution of ash particles entrained from the combustion chamber at various U/U_{mf}. On the whole, the shape of size distribution appears to be normal with largest particle size being about 250 μm. A single 80%-passing size, deduced from these curves, is used to represent the whole size distribution shown plotted as a function of U/U_{mf} in Figure 16. It is seen that air velocity has no definite effect on the particle size of entrained ash with an approximate 80% size of 180 μm. The size of ash particles leaving the scrubber unit is shown in Figure 17. For lignite coals, the mean particle size increases in the range 2.7 - 8.5 μm for increasing U/U_{mf} from 3.2-4.6.

Figure 18 compares ash particle concentration in flue gas and stack gas. It can be concluded that ash concentration in the flue gas is not affected by the change in the velocity of fluidizing air over the range studied. Approximately, 96% reduction in ash concentration of flue gas is achieved, giving dust concentration in the discharged gas of 0.03-5 g/NM^3.

Economic Analysis

The economic analysis of the 3-ton/hr FBC steam plant (water-tube) as compared to the conventional oil fired boiler (fired-tube) was made. The results in terms of payback time and IRR are concluded in Figure 19. Assumption and criteria used in the calculation are given in Table 6. The economic advantage of the FB boiler depends on both the capacity factor and the lignite price which varies according to the transporation cost. For the capacity factor above 60%, the new FBB system is economically attractive for the lignite price upto 1,000 bahts/ton.

CONCLUSIONS

The performance of a 2 MW FBC boiler burning and lignite of size -¾" and -2" has been evaluated and the following results can be drawn :

- Combustion of smaller size feed of -¾" gave less preheating time of the bed, hence saving LPG consumption.

- There was an optimum A/F ratio which gave maximum steam production but at the same lignite feed rate, the coarser size feed (-2") generated less amount of steam due to its smaller surface area available for combustion.

- Boiler thermal efficiency varied in the range of 60-80% and 30-70% for lignite feed size of -¼" and -2", respectively, and depended on A/F ratio in the same fashion as the steam production rate.

- Carbon combustion efficiency of lignite was well above 80% and slightly dependent on A/F ratio.

- Temperature of gas at various locations in the furnace was found to be independent of changing A/F ratio. For lignite combustion, temperature inside the bed was higher than the overbed temperature.

- Heat transfer coefficients of inbed and overbed water tubes were in the order of 400-800 and 10-100 $W/m^2\ {}^oC$, depending on the normalized air velocity, U/U_{mf}.

Evaluation of emission characteristics of the 2 MW fluidized bed boiler burning lignite gave the following results.

- Analysis of flue gas composition showed that percentage of CO is relatively low (< 0.20%), signifying high conversion of combustible fuel in the bed.

- The reduction of SO_2 by inbed capture using limestone and by water spray of the wet scrubber was sufficient to lower the concentration to the standard emission of 400 ppm.

- Air fluidizing velocity appeared to have little effect on the ash removal rate, entrained ash particle size and ash particle concentration.

- Particle size and concentration of particles in the stack gas were 2.7-8.5 μm and 0.03-5 g/NM^3, respectively.

The overall performance data of the prototype fluidized bed boiler is given below.

Combustion efficiency	>	90%
Thermal efficiency	>	70%
Overall heat transfer coefficient (bed-to-wall)		100-300 $w/m^2\ {}^oK$
SO_2 emission	<	200 ppm
Particulate emission	<	1.5 g/Nm^3

In the near future, the application of the Fluidized-bed Combustion Technology will help conserve foreign exchange by reducing imports of liquid fuels and crude oils as well as control gaseous emission at minimal level at a reasonable cost. This is one further step in the development for better energy production for Thai industry in future.

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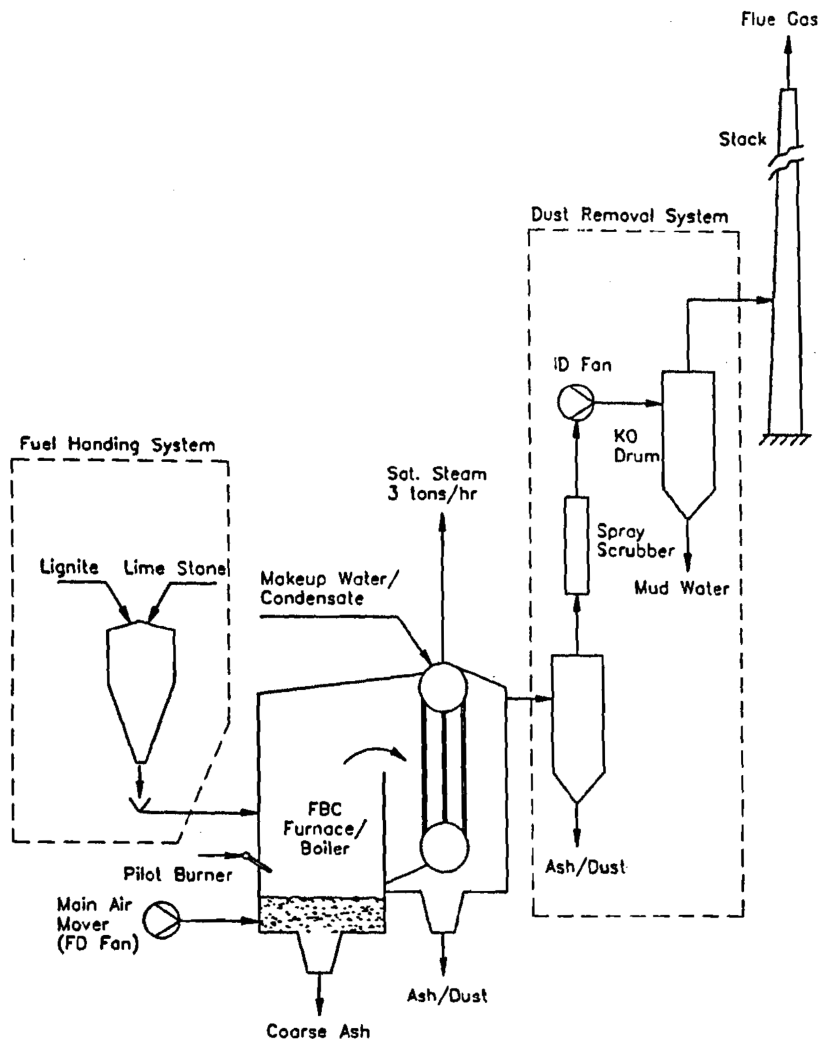


Figure 1 Fluidized-bed steam plant diagram

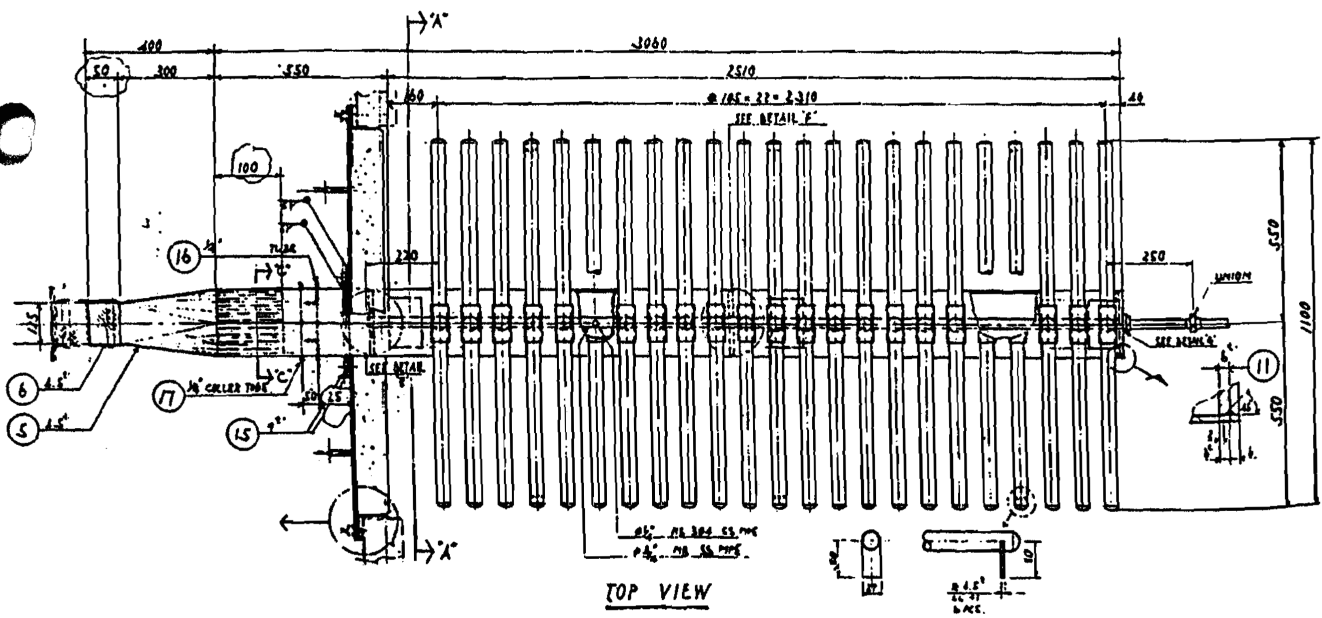


Figure 2 Differential multispout air distributor

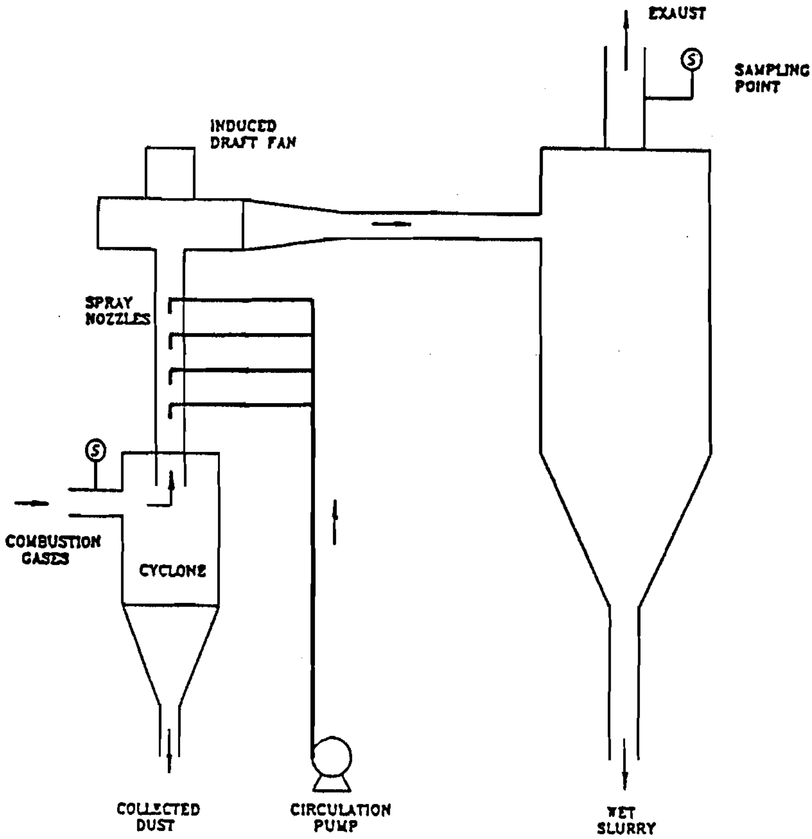


Figure 3 Dry cyclone and mechanical wet scrubber

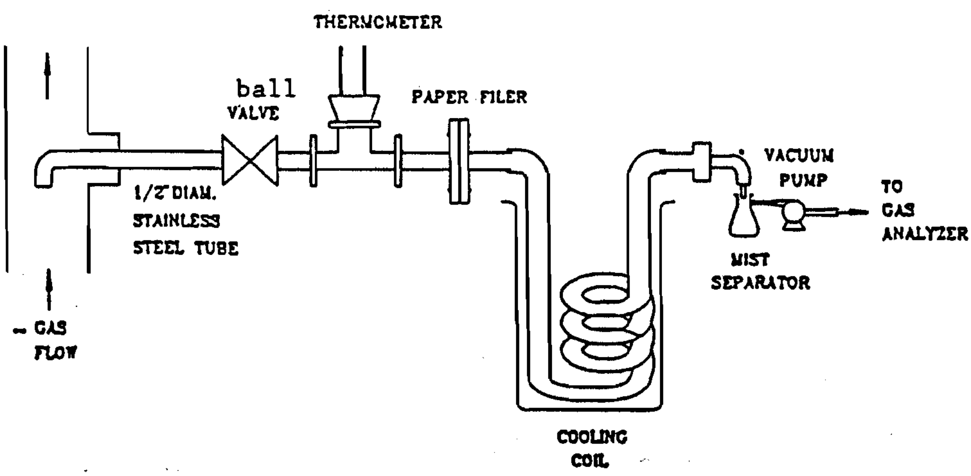


Figure 4 Sampling device for particulate and gas analyses

TABLE 1 PROPERTIES OF FUELS

Ultimate Analysis, %	Lignite #1	Lignite #2
C	47.20	41.91
N	1.12	0.87
H	5.19	4.58
O	23.30	16.79
S	0.98	1.19
ash	11.61	21.46
moisture	10.60	13.20
Heating value, kcal/kg	4730	4460
Feed size	-¾"	-¾" (-2")
Feed rate, kg/h	506	430 (435)

TABLE 2 BED CONDITIONS

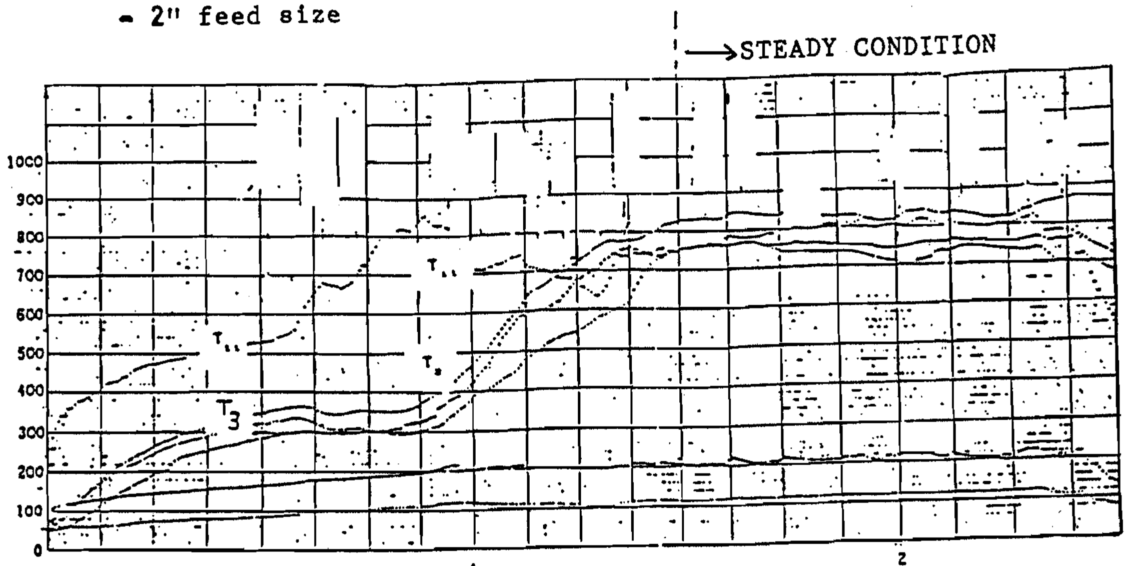
Bed material	Sand particles
Mean particle diam., mm	0.571
Particle sphericity	0.81
Particle density, gm/cm^3	2.16
Bed porosity	0.468
Static bed heigth, m.	0.30
Min. fluidizing velocity, m/s	0.27 (27 ^{o}C, 1 atm)

TABLE 3 COMPOSITIONS OF CO_2, CO AND O_2 IN FLUE GAS FROM FLUIDIZED COMBUSTION CHAMBER

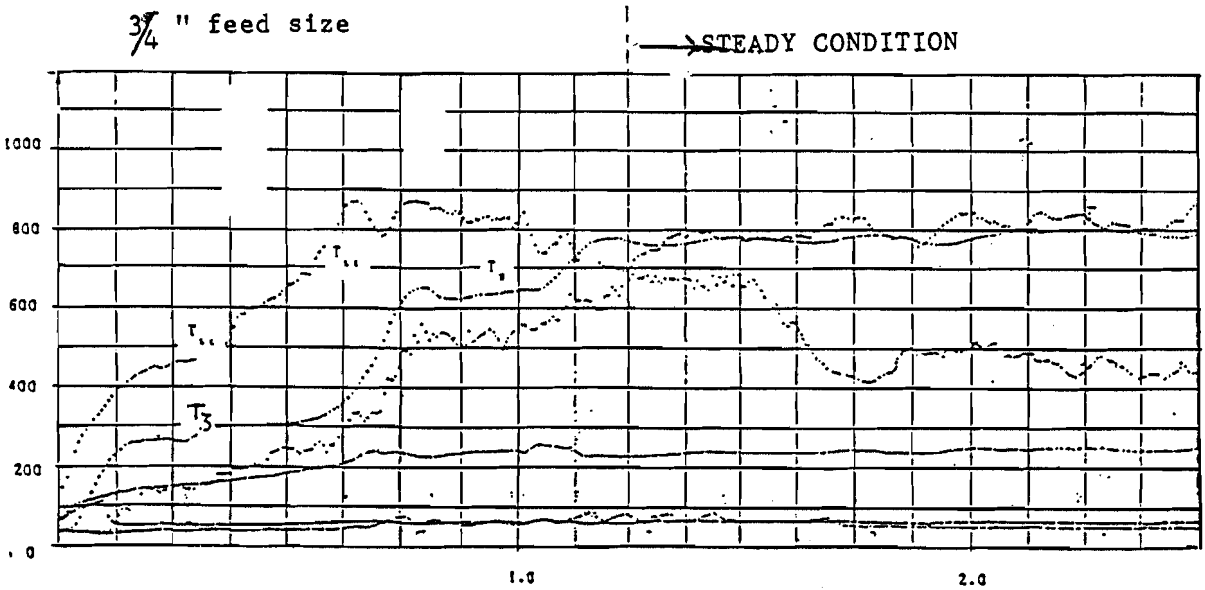
Fuel	Feed rate kg/h	Excess air, ratio	Range of gas composition, vol%			
			%	CO_2	O_2	CO
Lignite#1	506	5.4-8.4	-17-29	10-12.5	6-9.1	0.02-0.2
Lignite#2	430	6.8-8.2	15-30	6-10	9-15	0.1-0.4

TABLE 4 ASH REMOVAL RATE VERSUS ASH LINPUT RATE

Fuel	Average total ash removal kg/h	Ash input with fuel kg/h
Lignite 1	42	50
Lignite 2	50	90
Corn cob	6.0	6.1



(a)



(b)

Figure 5 Variation of bed temperature during bed preheating period for coal feed size of -3/4" and d-2"

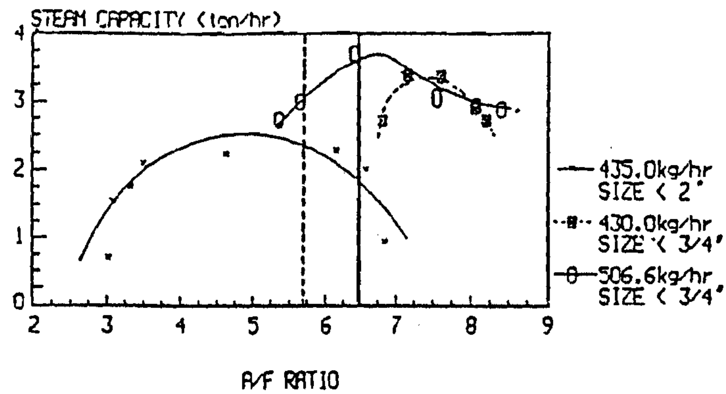


Figure 6 Effect of A/F ratio on steam production rate for coal combustion

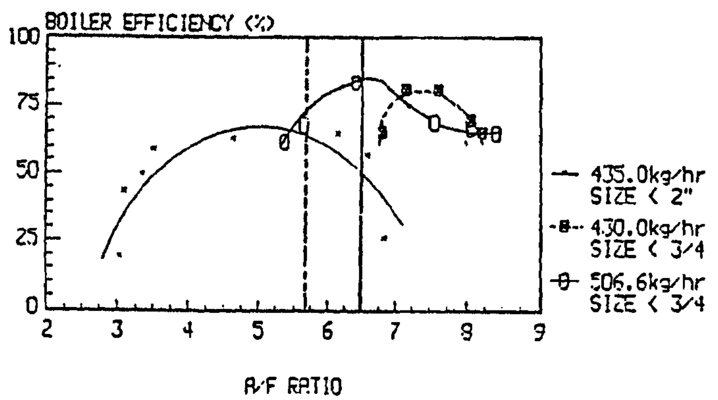


Figure 7 Effect of A/F ratio on boiler thermal efficiency for coal combustion.

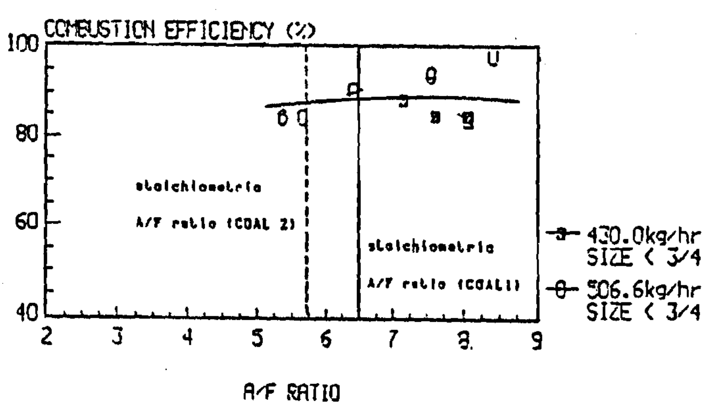


Figure 8 Effect of A/F ratio on combustion efficiency for coal combustion

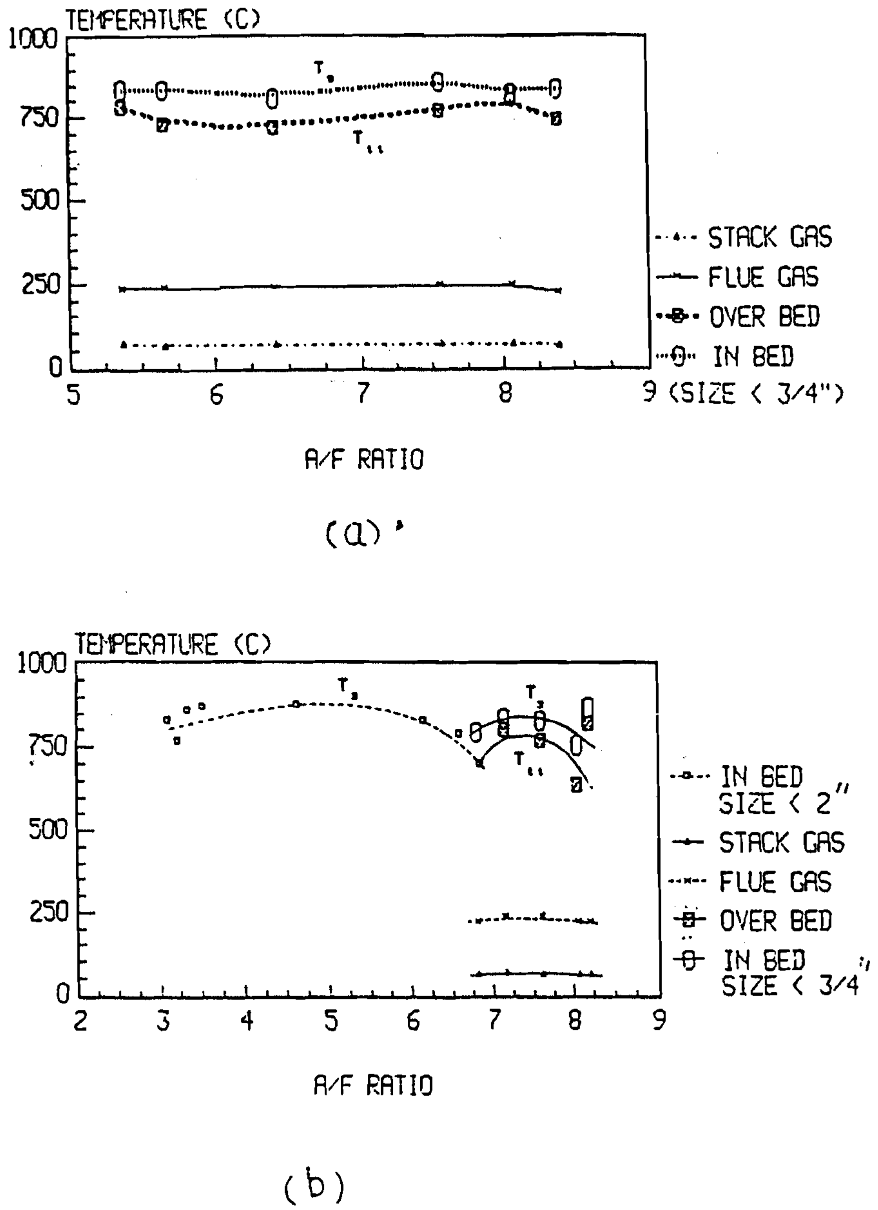


Figure 9 Effect of A/F ratio on gas temperature at various locations in the furnace system for coal combustion; (a) size - $\frac{3}{4}$" (b) size - $\frac{3}{4}$" and -2"

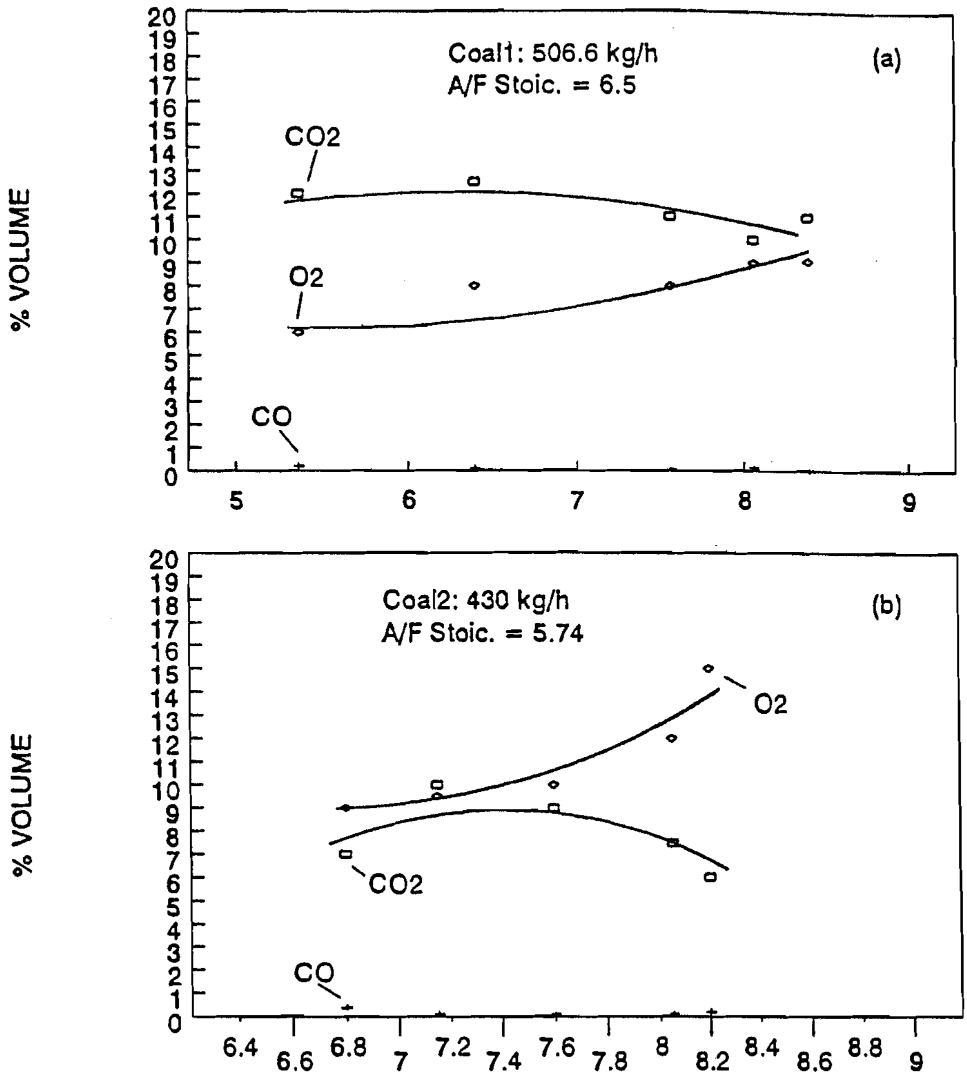


Figure 10 Effect of air-fuel ratio on flue gas composition (CO, CO_2 and O_2 gas)

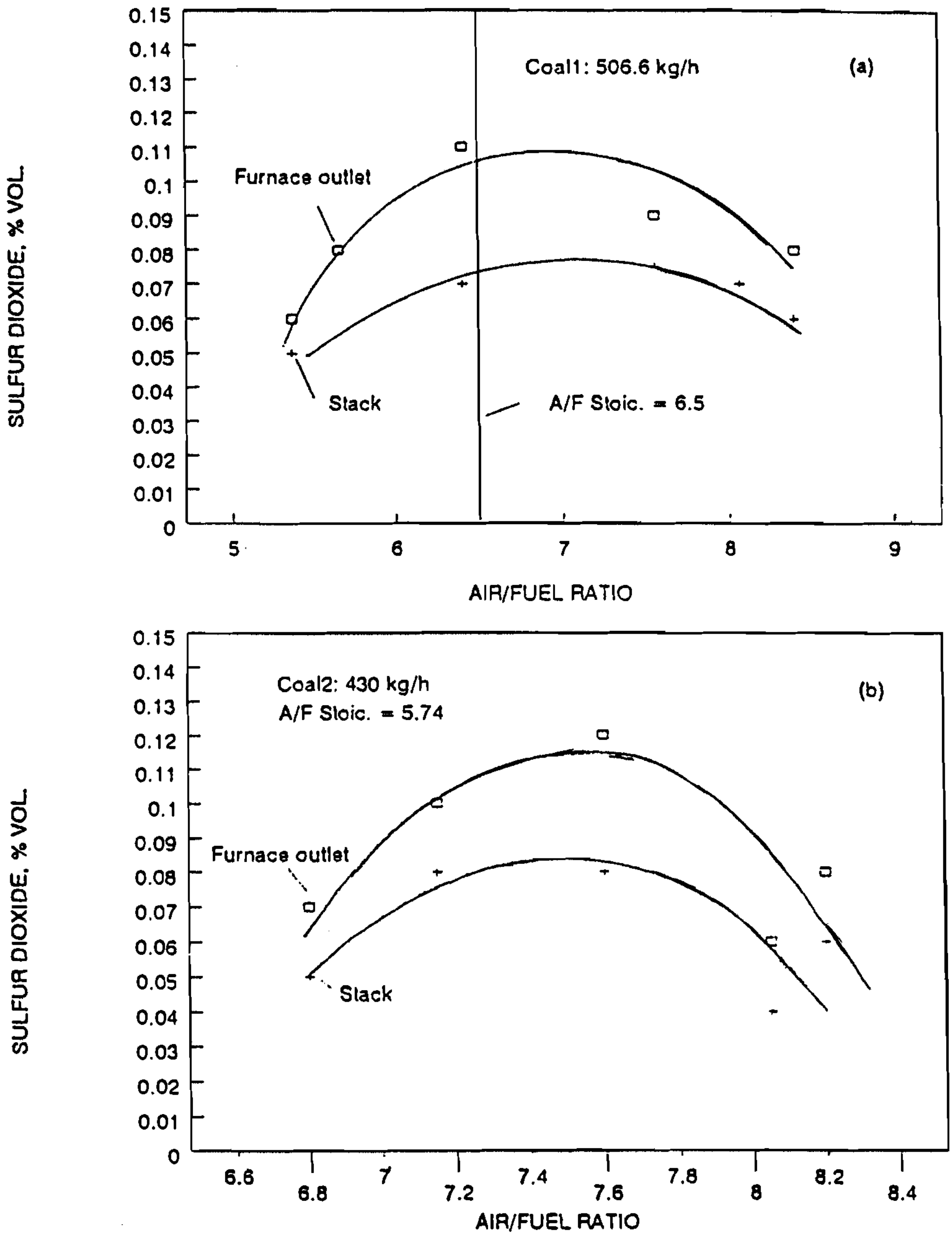


Figure 11 Effect of air-fuel ratio on sul-fur dioxide level

Table 5 Data of Sulfur dioxide removal with limestone in 2 MW (th) Fluidized Bed Boiler

Sulfur content in coal (%) =	1.467	Mean limestone particle size (mm) =	1
Ca content in limestone (%) =	36.67	Bed temperature (deg.C) =	830-840
Coal feed rate (kg/hr) =	300	Static bed height (m) =	0.3
A/F ratio =	11.055	Sand particle size (mm) =	0.571

Limestone feed rate (kg/hr)	Ca (kg/hr)	Ca/S (moles ratio)	SO2 (1)	SO2 (2)	1-[SO2 (1)/SO2]	1-[SO2(2)/SO2(1)]	1-[SO2(out)/SO2] (overall efficiency)
0	0	0	723	299	0	0.586445367	0.586445367
24	8.8	1.6	513	141	0.290456432	0.725146199	0.804979253
30	12	2	351	202	0.514522822	0.424501425	0.720608575
36	13.2	2.4	265	220	0.633471646	0.169811321	0.69571231
42	15.4	2.8	252	132	0.651452282	0.476190476	0.817427386
48	17.6	3.2	137	119	0.810511757	0.131386861	0.835408022
50	22	4	145	95	0.79944675	0.344827586	0.868603043

Remarks : (1) = Gas concentration before entering dry cyclone (ppm)

(2) = Gas concentration at wet cyclone exit (ppm)

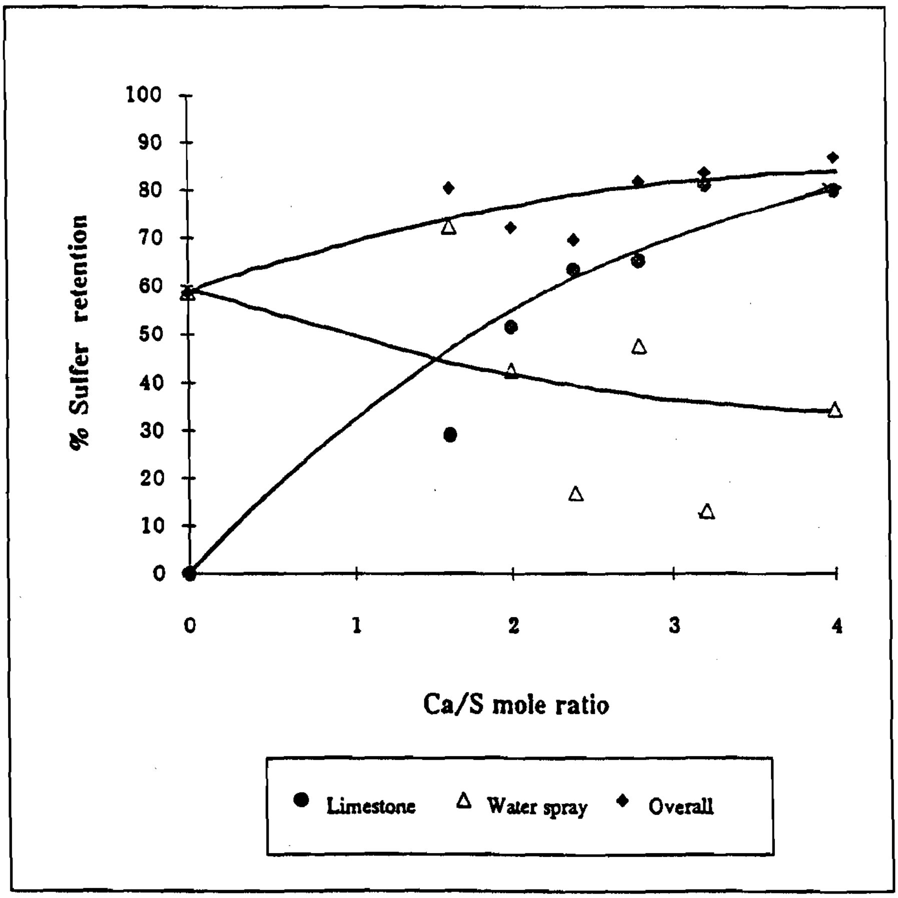


Figure 12 Sulfur Retention in the Fluidized Bed Boiler

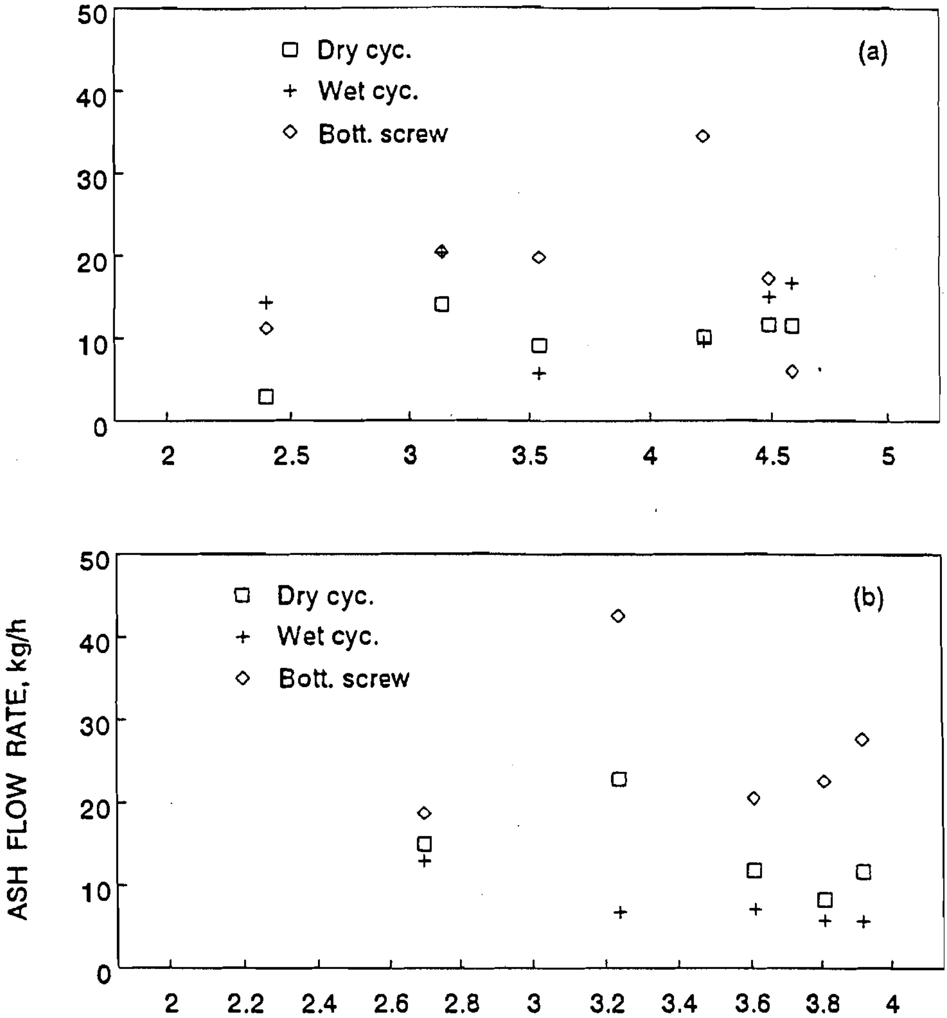


Figure 13 Effect of air fluidizing velocity on removal rate of ash particles; (a) coal 1 (506 kg/h). (b) Coal 2 (430 kg/h)

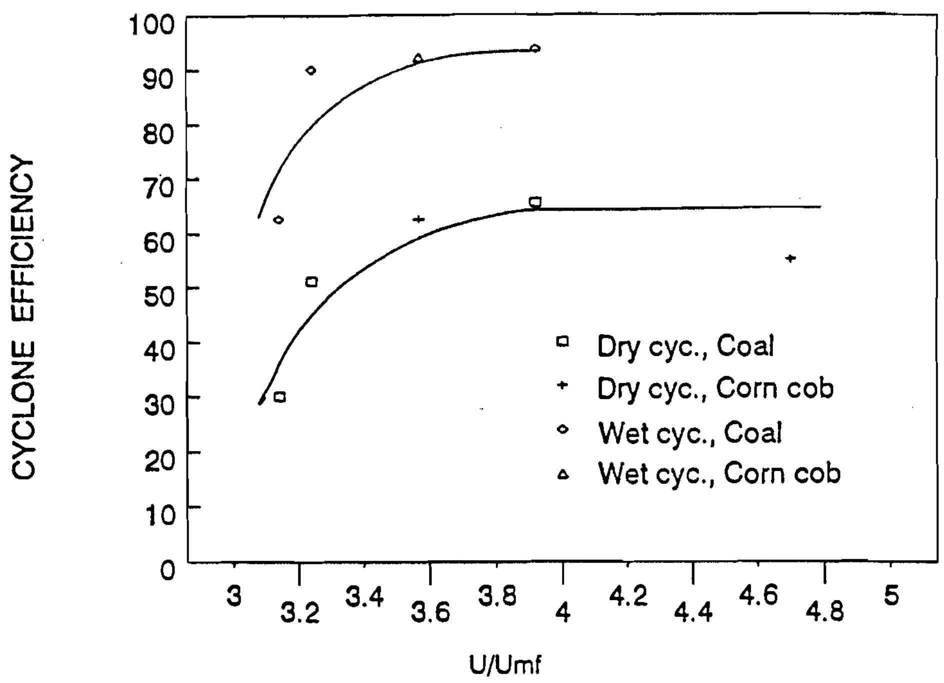


Figure 14 Particulate collection efficiency of dry cyclone and Wet scrubber at various U/U_{mf}

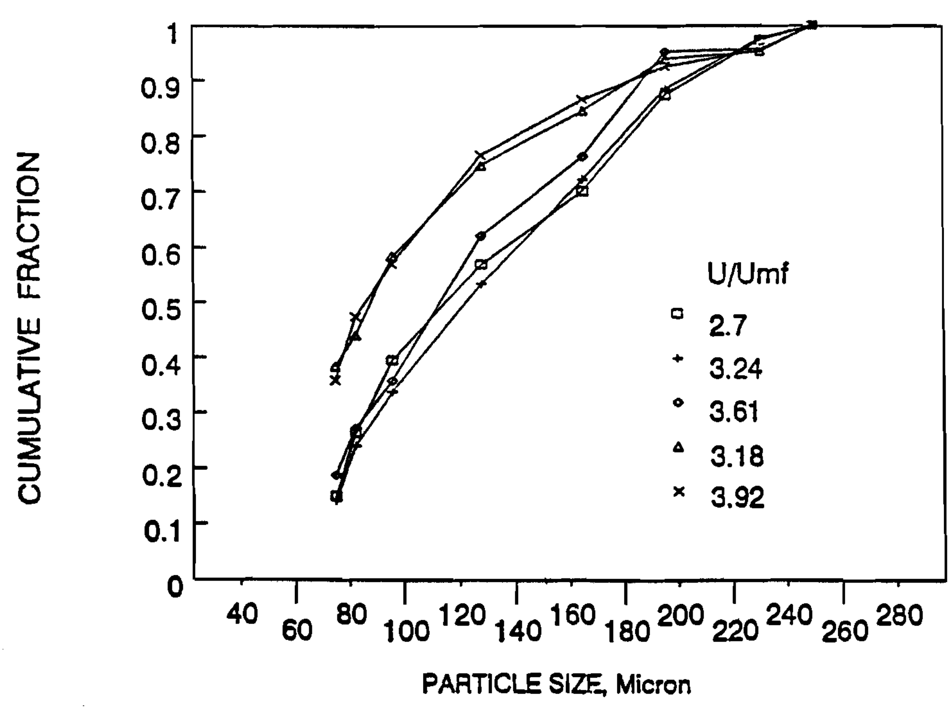


Figure 15 Size distribution of entrained particles from combustion chamber

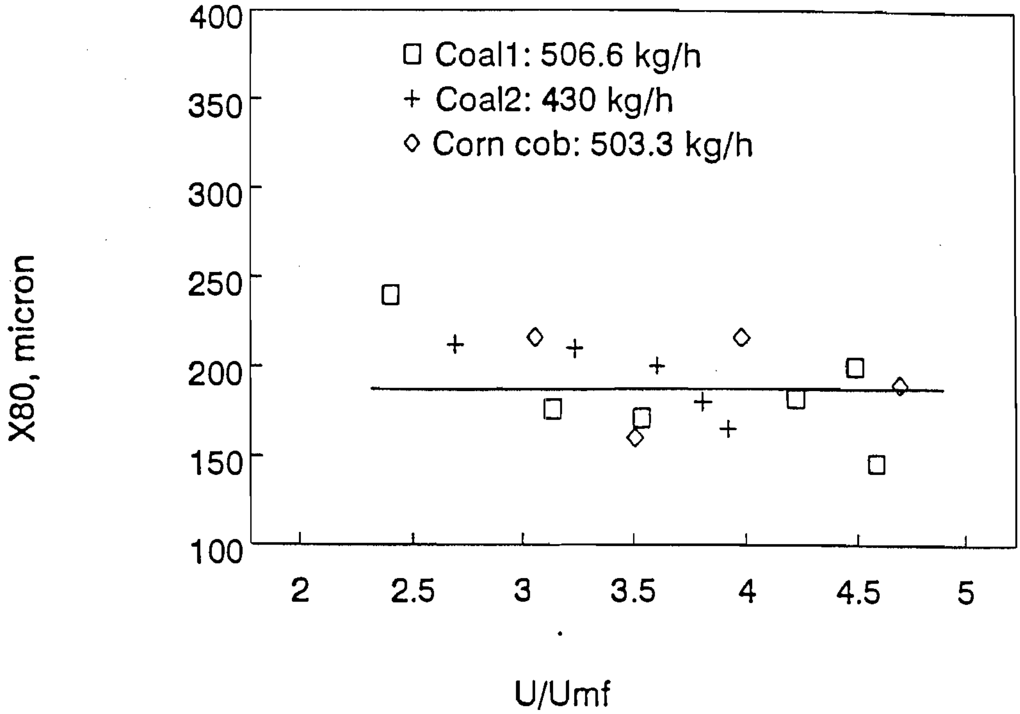


Figure 16 Effect of U/U_{mf} on 80% - passing size of entrained particles from combustor

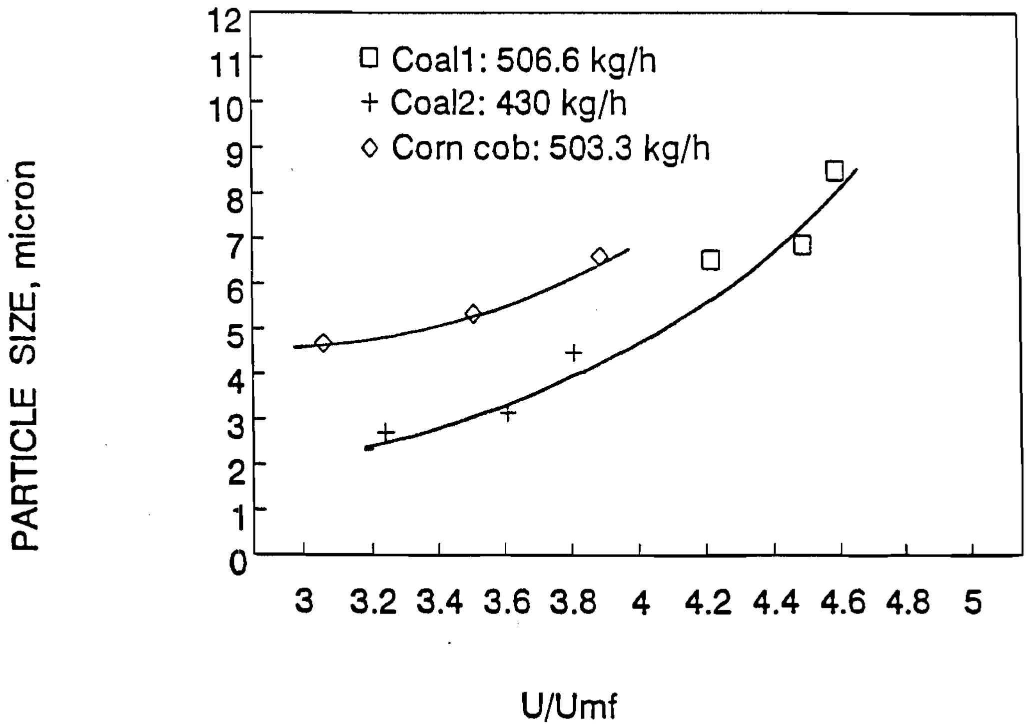


Figure 17 Effect of U/U_{mf} on the average size of particles discharged to atmosphere

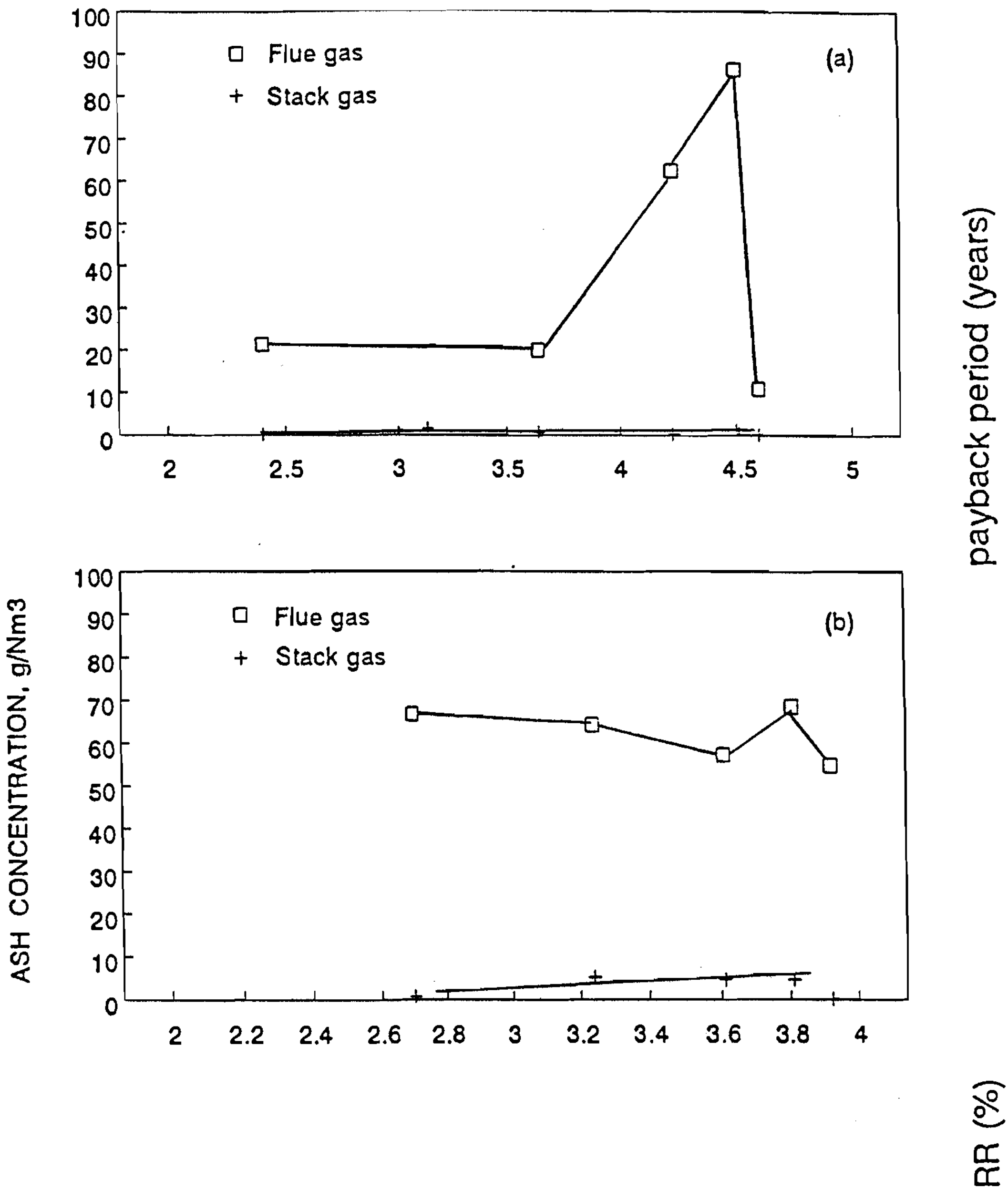


Figure 18 Effect of U/U_{mf} on dust concentration in flue gas and stack gas : (a) Lignite 1, (b) lignite 2

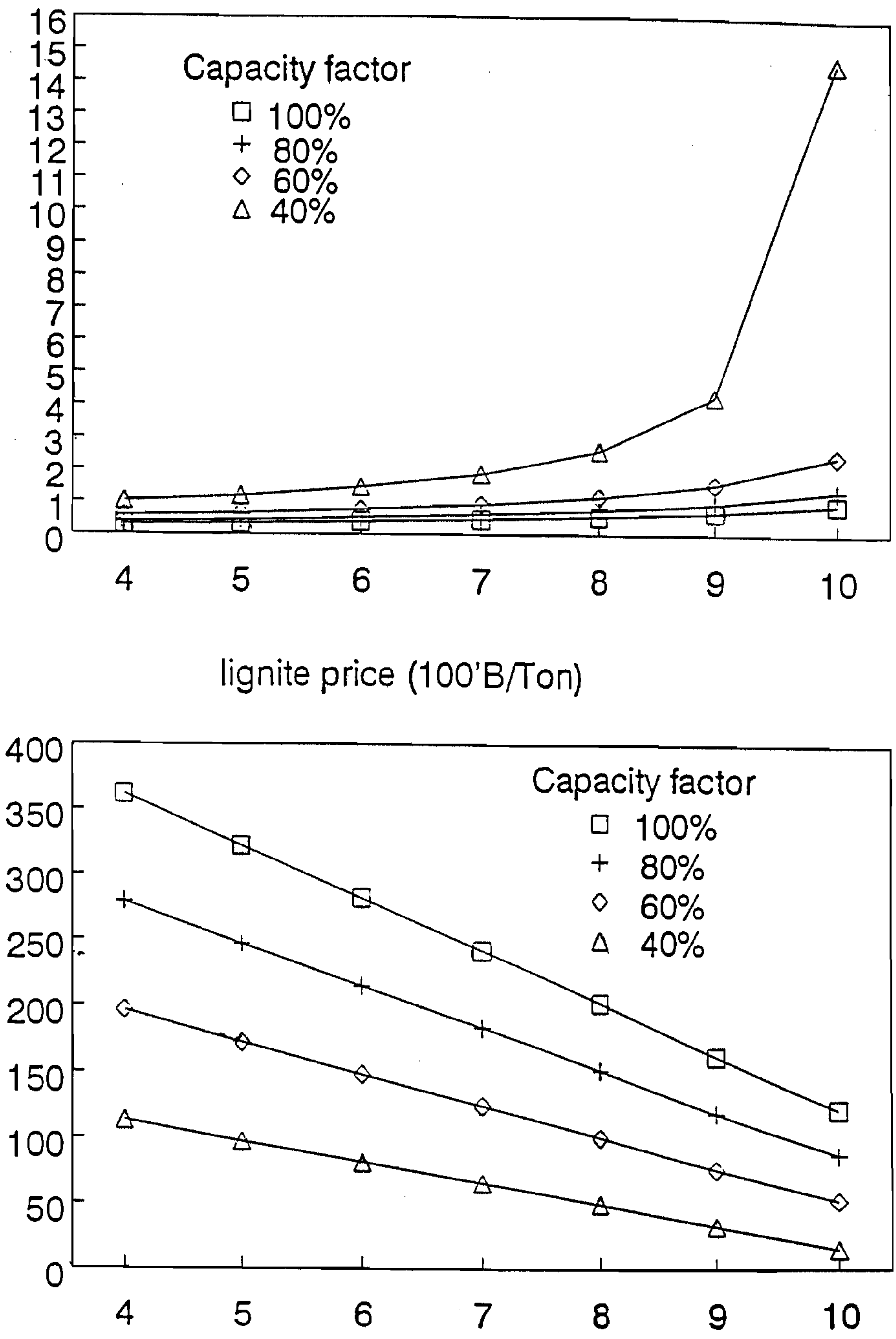


Figure 19 Economic Analysis of the 2 MW_{th} Fluidized Bed Boiler as Compared to a Fired-Tube Boiler.

TABLE 6 ASSUMPTIONS AND CRITERIA USED IN THE ECONOMIC ANALYSIS CALCULATION

1. Total Investment Costs for Boiler 3t/h (1993 prices) - Oil Fired Boiler - Fluidized Bed Boiler	 2500000 Baht 3500000 Baht
2. Operating & Maintenance Cost (excluded fuel cost) - Oil Fired Boiler - Fluidized Bed Boiler	 3% of fuel cost 15% of TIC
3. System Life - Oil Fired Boiler - Fluidized Bed Boiler	 15 years 15 years
4. Discount Rate	15% per year
5. Fuel Price - Fuel Oil - Lignite	 3.3 Baht/l 400 to 1000 Baht/ton
6. HHV of Fuel - Fuel Oil - Lignite	 39647 kJ/l 41441.4 kJ/kg 4000 kCal/kg 16748 kJ/kg
7. Boiler Efficiency - Oil Fired Boiler - Fluidized Bed Boiler	 80% 80%
8. Fuel consumption - Oil Fired Boiler - Fluidized Bed Boiler	 211 l/h 500 kg/h
9. Operating Hours per Year	8000 hrs/year (Capacity factor = 100%)
10. Salvage Value	0 Baht