Chapter 1

Combustion

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1.1 INTRODUCTION

The fluidized bed technology can be used for the combustion and incineration of a wide range of solid and liquid fuels and wastes. However the design of each fluidized bed combustor should be matched with the characteristics of the fuel and the additive applied. That means that the characteristics and the properties of the fuel concerning combustion in a fluidized bed must be known. On the other hand combustion parameters can be varied influencing combustion behaviour of the fuel.

As all coals used in the Dutch industries is imported from countries all over the world, the influence of coal type on the combustion behaviour is important to know so that a range of coal types can be defined for a specific fluidized bed design.

The optimal combustion efficiency which can be obtained, the SO_2 and NO_x emission levels and the amount and composition of the solid residues depend on coal type and on process conditions.

In this chapter attention will be given to the combustion efficiencies of various coal types in relation with the combustion parameters:

-fly ash refiring
-bed temperature, freeboard temperature
-bed/freeboard combustion
-excess air, staged combustion
-bed height
-freeboard height, internals
-particle size bed material
-fuel supply, dry/slurry.

In section 2 the research executed at the TNO 4 MW_{th} Atmospheric Fluidized Bed Boiler (AFBB) facility is reported and discussed.

The research performed at the TU 1 MW_{th} Atmospheric Fluidized Bed Combustor (AFBC) is presented and discussed in section 3.
As these two research facilities have each specific benefits and constraints the results of experimental research are not combined but discussed separately. Next to it in section 4 a simulation of the combustion efficiency as a function of various process parameters is described and used for analyses of experimental results.

The total combustion time of a single coal particle in a fluidized bed can be divided into a relatively short period during which volatiles evolve, followed by a much longer period in which the remained char particle is burning. The combustion of volatiles is supposed to be complete, as turned out from experiments, thus the combustion efficiency is mainly related to the combustion process of char particles. The energy production from volatiles combustion is in general somewhat lower but of the same order of magnitude as that of char combustion. The volatiles burn in the bubble phase (more than in the dense phase) of the bed and in the splash zone of which the heat can be transported into the bed by back mixing of bed particles. (see v.d.Honing [1]).

For the calculation of combustion efficiencies from experimental results the calorific values of the fuel and of the carbon is used, together with the mass flow rates of coal and fly ashes. The combustion efficiency is defined by the formula:

$$\eta_c = \frac{H_o B - H_c(C_R m_D + C_F m_F)}{H_o B}$$

C_R	(-)	Carbon content of fly ash from mechanical dust collector
C_F	(-)	Carbon content of fly ash from baghouse filter
H_o	(MJ/kg)	Gross calorific value of the coal used
H_c	(MJ/kg)	Calorific value of carbon
B	(kg/s)	Flow rate of the coal supplied
m_D	(kg/s)	Flow rate of the drained fly ash form the mechanical dust collector
m_F	(kg/s)	Flow rate of the drained fly ash from the baghouse filter
η_C	(-)	Combustion efficiency

The recycle ratio is defined as the ratio of the total flow rate of the recirculating fly ash to that of the coal fed into the combustor:

$$R = \frac{m_R}{B}$$

m_R	(kg/s)	Flow rate of recycled fly ash
R	(-)	Recycle ratio

1.2 EXPERIMENTAL RESEARCH 4 MWth -BOILER

1.2.1 Experimental conditions

The TNO-AFBB is a 4MW_{th} coal fired Atmospheric Fluidised Bed Boiler, designed for research on the combustion of coal or other fuels such as petroleum coke, oil and gas. A flow diagram is given in figure 1.2.1 .

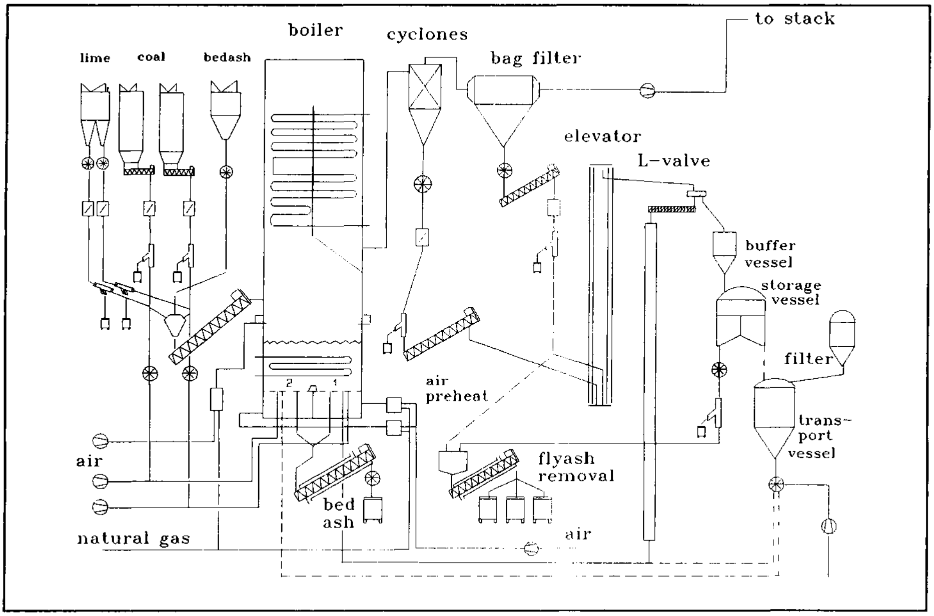


Fig.1.2.1. Flow diagram of the TNO 4MWth AFBB Research Facility.

Coal, transported with screw feeders from two day bunkers, is pneumatically fed to the boiler and injected underbed in the 1.0 x 2.2 m^2 fluidized bed via two nozzles. Overbed feeding is possible via a screw conveyor. Limestone is supplied in the same way, either overbed or underbed.

During a short period coal has been fed overbed as a coal slurry.

In both limestone and coal lines switch valves are installed for sampling and calibration after the flow measuring devices. Installation of a bed ash container and transport lines was necessary to store the bed ash during periods of repair and change of the boiler as well as to supply spent bed material or sand to keep the bed height at the desired level when coal with a low ash content or high attrition tendency is used.

Combustion air, which can be preheated with natural gas burners, is transported via two wind boxes and air nozzles to the fluidized bed. Also, nozzles are mounted in the side walls of the freeboard section, providing secondary air in case of staged combustion. The control of the secondary air mass flow is integrated with the control of the primary air to allow for operation with a constant total air ratio. The capacity of the system is such that 35 % of the total air flow can be added as secondary air.

Both bed and freeboard sections are membrane walled and entirely refractory lined. Consequently the energy loss through the walls is small and cooling of the freeboard section is prevented.

Combustion of the coal takes place in the bed section (expanded height normally 1.05 m) at temperatures varying between 800 and 900°C and in the 3.70 m high freeboard. The energy generated by combustion is extracted from the bed by means of an 7.3 m^2 in-bed tube bank. In the bed tubes, membrane walls and convection tube banks (after freeboard) saturated steam is produced. Steam and water are separated in steam drums.

Surplus bed ash is automatically transported via a cooled bed ash screw to a bed ash container on a weighing platform. To improve the combustion efficiency of the coal, fly ash (cyclone and filter ash) is recycled, using a lock-hopper system, which has recently been replaced by a more simple L-valve device. It is also possible to recycle cyclone ash only, or together with part of the filter ash.

Bed ash samples can be taken from sample ports in the wall of the boiler. Gaseous samples at different levels of the freeboard can be taken with a special developed hot sample probe via sample ports in the front wall of the boiler.

After the freeboard, the dust-laden flue gas is cooled in the first and second convection passes to about 240 °C and cleaned from ash in a multi-cyclone and a bag filter (dust 5 mg/nm^3). Between boiler and cyclones flue gas is sampled and analyzed continuously on SO_2, NO_x, CO_2, O_2, CO and C_xH_y. Generally the C_xH_y content is below 5 mg/nm^3, expressed as CH_4.
In the flue gas line after the bag filter, a sample port for the sampling of stack ash and trace gas components is provided.

Design data of the unit are given in table 1.2.1.

TABLE 1.2.1
Design data of the $4MW_{th}$ AFBB Research Facility.

Output nominal	4 MW_{th}	Bed temperature	800-900°C
Coal supply	broken 0-6 mm dried	Expanded bed height	1.05 m
Coal feed max.	670 kg/h	Freeboard area	2.2 x 1 m
Bed sections	2; 1.1 x 1 m each	Freeboard height	3.7 m
Fluidization velocity	1-3 m/s	Steam production	7 tons/h, 10-15 bar, saturated.
Possibility for oil and gas firing. Start up on gas.		Closed water/steam circuit with forced circulation. Steam is produced in bed and convection sections and side walls.	

Limestone supply: The amount of limestone supplied depends on the SO_2 emission target.

The facility is computer controlled and equipped for unattended operation.

1.2.2 Combustion efficiency versus coal types

In this section an overview is given of the combustion efficiencies achieved with the various coal types used, followed by a discussion about the combustion parameters related to the combustion efficiency.

A variety of coal types has been examined: Polish, Virginian, Columbian, Australian (Queensland), Petroleum coke, Kentucky, Venezuela and Steam Slack. Most of the experiments have been carried out with Polish coal, of which 14 lots are used in total. With respect to the detailed description of the combustion characteristics of the various coal types there should be referred to the specific reports containing the detailed information of the experiments [17].

In table 1.2.2 the analysis data of the coal types used are listed:

TABLE 1.2.2
Analysis data of the used coal types.

Coal types		Polish	Virginia	Columbia	Australia (Queensland)	Petroleum Cokes	Kentucky	Venezuela	Steam Slack
On dry									
C	%	75.3	77.1	71.4	73.3	87.5	70.6	73.2	71.4
O	%	7.3	6.3	10.6	6.0	1.6	8.1	10.1	10.8
H	%	4.5	4.4	4.5	4.1	3.5	4.9	4.2	3.9
N	%	1.4	1.4	1.4	1.4	1.4	1.6	1.5	1.7
S	%	0.77	1.52	0.82	0.60	5.53	2.13	0.73	0.60
Ca	%	0.46	0.19	0.13	0.30			0.13	0.40
Ash	%	10.9	9.7	10.5	13.5	0.5	13.1	10.3	11.6
Volatile	%	30.7	27.3	40.3	25.5	11.5	33.5	30.8	29.7
HHV	MJ/kg	30.75	30.98	29.51	29.30	34.94	29.38	29.2	29.2
Dry and ash free									
C	%	84.5	85.4	79.8	84.7	88.0	81.2	81.6	80.8
O	%	8.2	6.9	11.8	6.9	1.6	9.3	11.3	12.2
H	%	5.1	4.9	5.0	4.7	3.5	5.6	4.7	4.4
N	%	1.5	1.6	1.6	1.6	1.4	1.8	1.7	1.9
S	%	0.86	1.68	0.92	0.69	5.56	2.45	0.81	0.68
Volatile	%	34.4	30.2	45.0	29.5	11.6	38.6	34.3	33.6

With respect to the combustion efficiency a target value of 98.5 % was set.

In fig. 1.2.2. the measured maximum combustion efficiencies of the coal types tested are presented.

The highest combustion efficiencies of 99.3 and 99.5 % refer to the Polish and Australian (Queensland) coal types respectively. With the other coal types investigated, maximum values of 98 % to 98.3 % are achieved. Differences between the various coal types tested are small.
A lower value for the combustion efficiency of 97 % is measured for the Virginian coal. A closer examination of the combustion regime in the boiler shows that the combustion efficiency in the bed section of the Virginian coal, compared to the Polish or Columbian coal, is 3 % lower. This is thought to be the result of a lower reactivity of the Virginian coal. This results in a higher carbon content in the freeboard which gives rise to a higher contribution of the freeboard combustion. From a less reactive coal it can be expected that the residence time in the bed of the finer refired particles is too short to contribute to the in-bed combustion. This explains the independency of the in-bed combustion of the Virginian coal with respect to the recycle ratio (fig 1.2.4).

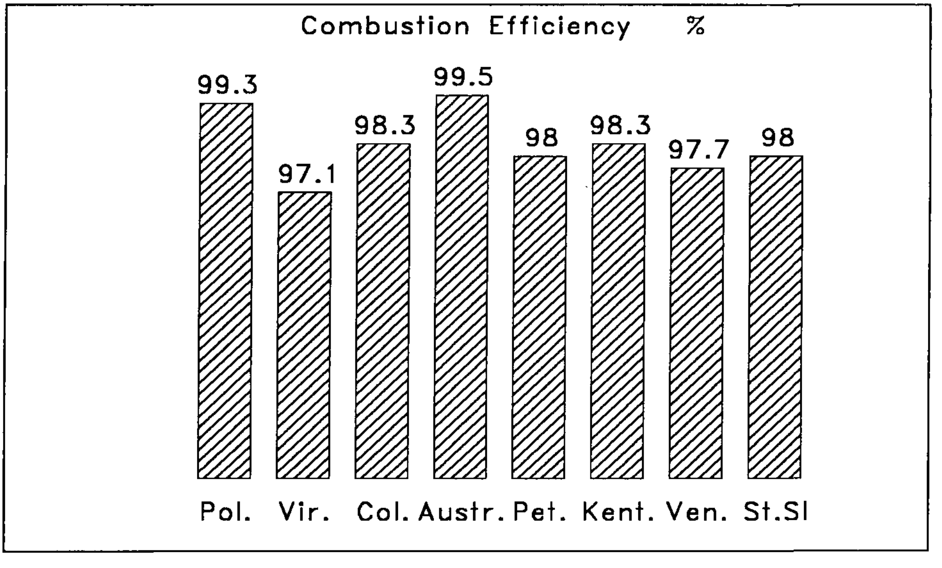


Figure 1.2.2. Combustion efficiency of coal types tested.

Pol = Polish	Pet. = Petroleum cokes
Vir = Virginia	Kent = Kentucky
Col = Columbia	Ven = Venezuela
Austr = Australia	St.Sl = Steam Slack

1.2.3. Fly ash refiring; bed-, freeboard combustion

Achieving a high combustion efficiency is important for overall plant efficiency and consequently for plant economics. Carbon loss occurs through the carry-over of unburnt particles and is mainly found in the collected fly ash (cyclone ash and filter ash). The loss in combustion efficiency due to the incomplete combustion of carbon to carbon monoxide is, during normal operating conditions, less than 0.1% and of minor importance. Generally, when no excessive quantities of bed ash have to be disposed of, the efficiency loss via the bed ash removal is also small and can be ignored.

The loss in combustion efficiency with the collected fly ash can be compensated for by recycling cyclone and filter ash. By recycling fly ash with unburnt char, the combustion efficiency is increased as a result of the enhanced residence time for the combustion of the unburnt fines. This facility is incorporated within the TNO 4MW_{th} test facility. In this way the TNO bubbling bed AFBB can be seen as an attractive intermediate between the so called stationary atmospheric fluidized bed combustor and the circulating fluidized bed combustor (CFB).

In the next figures the effect of fly ash recycling on the in-bed and freeboard combustion and, as a result of both, on the overall combustion efficiency is shown for Polish, Virginian and Columbian coal.

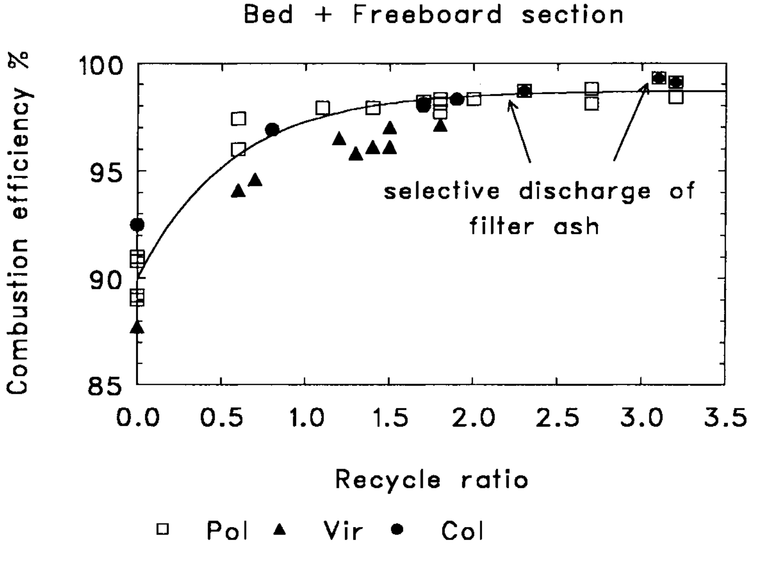


Figure 1.2.3. Overall combustion efficiency of Polish, Virginian and Columbian coal vs. recycle ratio. Bed temperature 845 °C; air ratio 1.2; fluidizing velocity 2.1 m/s; bed height 1.07 m; Ca/S molar ratio 1-3.

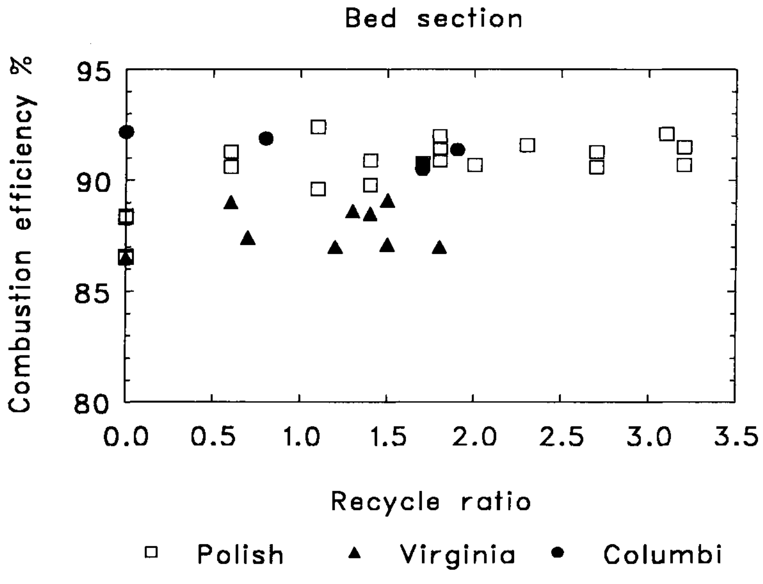


Figure 1.2.4. Combustion in bed section.

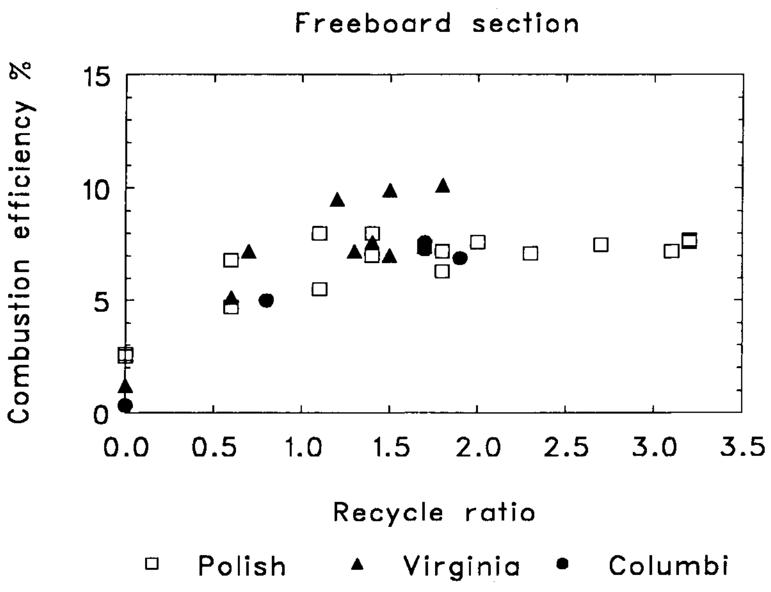


Figure 1.2.5. Combustion in freeboard section.

The effect of fly ash recycling on combustion efficiency is clearly demonstrated. The largest increase is measured between recycle ratio 0 and 1 (fig. 1.2.3). Up to a recycle ratio of 2.5 an increase of combustion efficiency can be achieved of approximately 9%. For example, without recycling the combustion efficiency of Polish coal amounts to 90%, whereas with a recycle ratio of 2.7 the combustion efficiency is increased to 99%.

The combustion of the Polish coal in the bed section slightly increases to a maximum at a recycle ratio of 1.0 (fig. 1.2.4). At higher recycle ratios the combustion in the bed section shows hardly any increase. The freeboard combustion (fig. 1.2.5), however, shows a larger increase and is more dependent on the recycle flow because the recycled fines mostly burn in the freeboard after being heated up in the bed section.

The overall combustion efficiency of Columbian coal shows the same pattern as for Polish coal with exception of the in-bed combustion without recycling. The reason for this is not clear.

Concluding from the overall efficiency, the Virginian coal seems to be less reactive than the Polish and Columbian coal. This results in a higher carbon content in the freeboard which gives rise to a relatively higher contribution of the freeboard combustion. From a less reactive coal it can be expected that the residence time of the finer particles in the bed is too short to contribute to the in-bed combustion. This explains the independency of the in-bed combustion of the Virginian coal with respect to the recycle ratio.

With a selective discharge of filter ash, the efficiency loss, compared to the combined removal of filter ash and cyclone ash from the recycle flow, can be reduced. This has been tested with Polish coal. A combustion efficiency of 99.3% (see figure 1.2.3) has been achieved when only part of the filter ash was discharged and not recycled, as filter ash had lower carbon content than cyclone ash. The capacity of the recycle system was about 1500 kg/h. By limiting the filter ash flow to a maximum of 200 kg/h the cycloon ash flow could increase to 1300 kg/h. In this way relatively more cyclone ash with higher carbon content was recycled. By this the efficiency loss due to the cyclone ash disposal deminished from 1.4 % (no selective filter ash discharge) to 0.3 % with selective filter ash discharge.

Differences in the combustion behaviour cannot be explained at first sight, by differences in composition, as shown by the proximate and ultimate analysis data. However, when the analysis data are expressed in the atomic ratio H:C:O, a higher degree of coalification of the Virginian coal can be concluded, related to differences in molecular structure. Differences in molecular structure can result in differences in porosity, effecting the diffusion controlled transport properties, or in a change of the kinetics of the combustion process.

With respect to the correlation between coal rank and combustion behaviour, though in many cases a quantitative correlation cannot be found, in general a higher coal rank (using an equal particle size distribution) leads to a higher ignition temperature and a longer burning time [2].

Much work is done at the Twente University in the Netherlands [3], to correlate the combustion characteristics and coal properties for use in a mathematical model developed by Brem [4]. This model describes the coal combustion process and is implemented in an overall reactor model, as reported in chapter 5.

1.2.4 Bed temperature, freeboard temperature

The bed temperature shows hardly any effect on the combustion efficiency, as is shown by figure 1.2.6. within the tested temperature range from 820 to 880 °C.

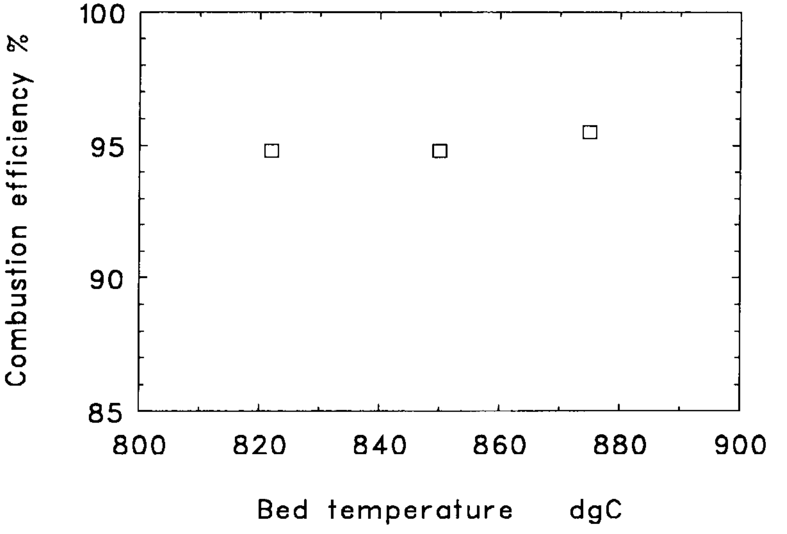


Figure 1.2.6. Combustion efficiency vs. bed temperature. Polish coal; air factor 1.15; fluidizing velocity 2.15 m/s, bed height 1.05 m; recycle ratio 1.7 (cyclone ash only).

As discussed in chapter 5, for fluidized bed conditions with typical bed temperatures varying from 800 to 900 °C and mm-sized coal particles, the combustion mainly takes place in the boundary layer combustion regime, i.e. pore diffusion and external diffusion control, and is less kinetically controlled [4]. This explains the small temperature dependency of the combustion.

1.2.5 Excess air, staged combustion

To achieve a considerable reduction in NO_x-emission, staging of the combustion air is applied. The combustion air is split into primary and secondary air. The primary air flow is reduced to obtain substoichiometric conditions in the bed section, by this the fluidizing velocity is reduced too.

The secondary air can be supplied into the freeboard, 1.65 and 2.65 m above the air distributor.

The stoichiometry is expressed as the air factor, i.e. the ratio of the used amount of combustion air to the theoretical mass balanced quantity of combustion air required for complete combustion.

The effect of air staging and reduced primary air conditions on the emission values is discussed in chapter 2.5. Here the effect on combustion efficiency is concerned.

A lower primary air ratio with staged combustion, within the tested range from 0.87 to 1.20, does hardly have any negative effect on the combustion efficiency, as is illustrated in figure 1.2.7.

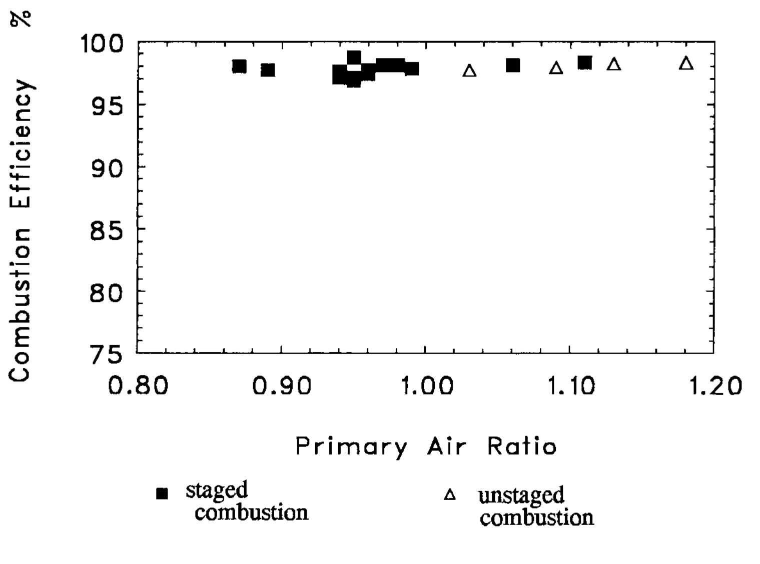


Figure 1.2.7. Combustion efficiency with staged and unstaged combustion vs. primary air ratio. Polish coal; total air ratio 1.2.

However the process conditions were not the same for staged and unstaged combustion, as the boiler is controlled via the bed temperature and by lowering the primary air flow through the bed the mass flow rate of the coal is also lowered to some extend. For this reason the load of the boiler, and also the mass flow rates of coal and air, was under staged combustion conditions lower than for unstaged combustion. This results in some positive effects on the combustion efficiency for staged combustion. The lower fluidizing velocity in the bed causes less elutriation and attrition of carbon particles and besides this, lower total air flow results in longer residence time in the freeboard section and better burnout of carbon particles. A negative effect is that lower air velocities causes less mixing and lower mass transport rates around a carbon particle.
Probably various effects compensate each other so no clear difference appeared from these experiments.

1.2.6 Bed height

No influence on the overall combustion efficiency has been observed with respect to the bed height, as has been verified from 0.7 to 1.35 meters. A more detailed analysis of the combustion split between bed and freeboard section reveals a decreasing combustion in the bed section with decreasing bed height but over the whole range the loss of combustion efficiency in the bed section is entirely balanced by the freeboard combustion.

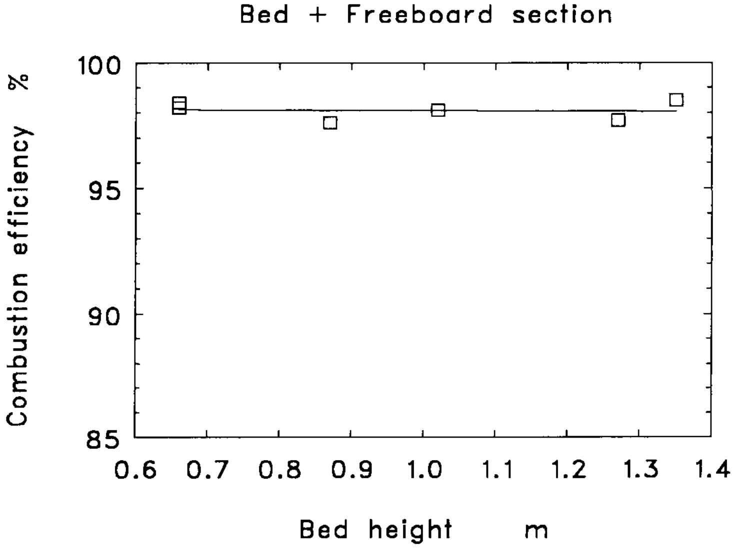


Figure 1.2.8. Combustion efficiency vs. bed height. Bed temperature 841 °C; Fluidizing velocity 2.1 m/s; air ratio 1.25; recycle ratio 1.7.

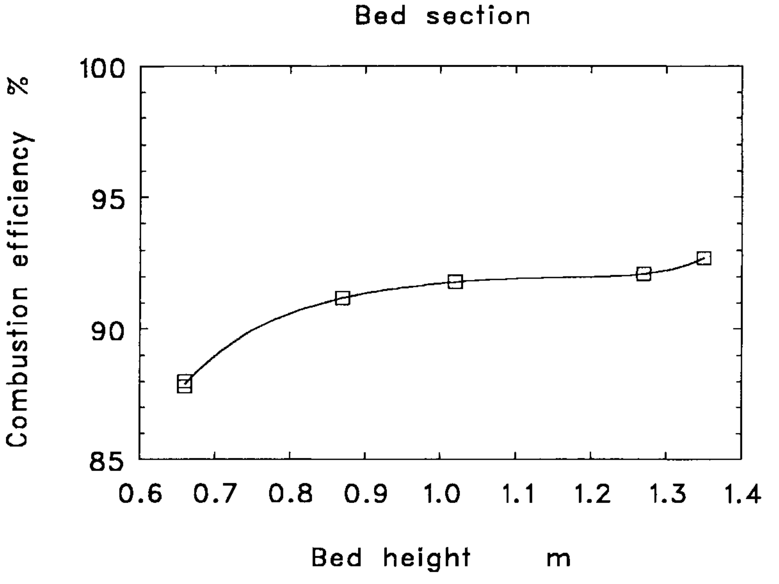


Figure 1.2.9. Combustion efficiency bed section.

Below 0.9 meter expanded bed height, the combustion in the bed section shows a strong decline. This phenomenon can be explained by gas flow jets between the bed bundle pipes. The upper layer of the bed bundle is positioned at 0.65 meters above the air distributor. Below a critical distance of about 0.20 meters between the upper surface of the bed and the heat transfer bundle the entrainment of the bed particles seems to be enlarged by these gas flow jets.

1.2.7 Freeboard height, internals

The major loss in combustion efficiency in AFBC is due to elutriation of the unburnt carbon fines. The fines may come directly from the feed or be generated during combustion by attrition (abrasion) or be the remaining part of a burnt coal particle, which is blown out of the bed.

In order to improve the overall combustion efficiency, refiring of fly ash is applied. As an alternative for a refiring system, the possibility has been investigated to enlarge the combustion efficiency in bed and freeboard, apart from refiring [8]. Apart from coal quality and particle size, an underbed feed system gives better results than an overbed feed system. The combustion efficiency can also be increased with a higher bed temperature but, as discussed before, not significantly effective in a temperature range between 800 and 900°C. Besides, the bed temperature is more or less fixed around 850°C because of the optimum conditions with respect to the sulphur capture.

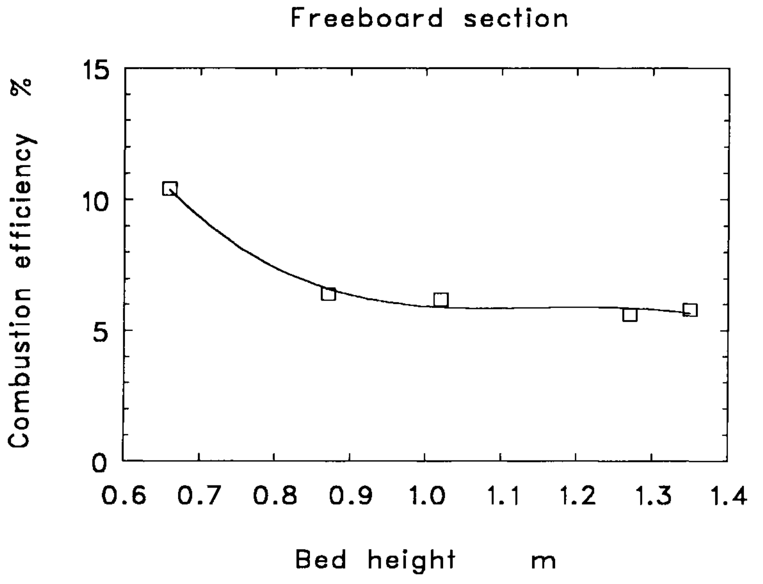


Figure 1.2.10. Combustion efficiency freeboard section.

For complete combustion of all elutriated coal particles the required residence time might be as high as 10 seconds, which means a freeboard height of 25 m for a fluidization velocity of 2.5 m/s.

To avoid the design of a tall boiler and yet to have reasonable efficiencies it has become good practice to recirculate cyclone and/or fabric filter ash into the fluidized bed. The coal particles, however, are cooled down in this way and have to be heated up in order to reignite. Therefore injection should be preferably done in the lower part of the bed while combustion itself takes mainly place in the freeboard.

However, ash recycle systems are expensive and complex, and therefore a main cause for trouble with a FBC-plant.

A different approach is to increase the residence time of the char particles in the freeboard itself before they leave the boiler. Therefore freeboard internals have been developed by TNO and a prototype has been tested.

Because TNO's AFBB was fully occupied for a long time the internals were built in a 3 MW_{th} FBC-test facility of "Royal Schelde" at Vlissingen with a 1.2 x 1.2 m^2 bed area. Coal was pneumatically fed through one nozzle in the bed. Cyclone ash could be recirculated through another nozzle in the bed. Freeboard height was 4 meters, measured from the 1 m expanded bed level to the convection section.

Based on aerodynamic research with a three dimensional perspex flow model and engineering considerations, a prototype was constructed consisting of rectangular channels in which baffles were mounted causing an increase of particle residence time.

An internal assembly of 2 m height was mounted in the freeboard of the 3 MW_{th} FBC-test facility 0.8 m above the expanded bed and 1 m below the convection heat transfer section. The arrangement of these internals is presented in figure 1.2.11.

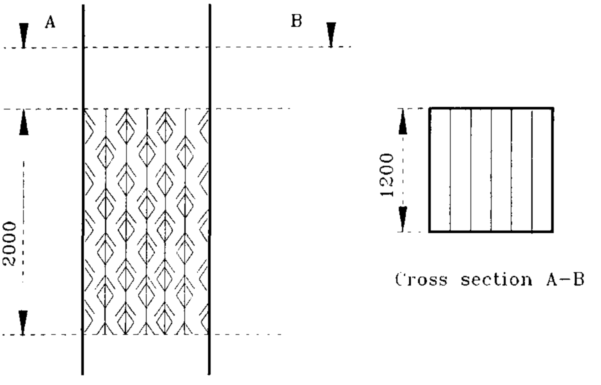


Figure 1.2.11. Arrangement of freeboard internals in testboiler.

Experiments were carried out at bed temperatures of 825 and 875°C, first with and without recycling of cyclone ash and after that with the use of internals. The results with respect to the combustion efficiency are summarized in table 1.2.3.

TABLE 1.2.3
The effect of freeboard internals on the combustion efficiency.

Bed temperature	825 °C	875 °C
	Combustion efficiency	
Once through	89 %	91 %
With cyclone ash refiring	98 %	98 %
With internals and without refiring	95 %	96 %

It is observed that the combustion efficiency improves with 8 % when refiring cyclone ash and 5 % with the use of freeboard internals only. As expected, somewhat higher combustion efficiencies are measured at a bed temperature of 875 °C than at 825 °C.

It can be concluded that freeboard internals have an equivalent effect as ash refiring on combustion efficiency, but ash refiring proves to be more effective.

1.2.8 Particle size bed material

The bed material is subject to attrition. The attrition is thought to be the result of the following size reduction mechanisms:

- size reduction due to fracture caused by thermal shock effects and/or collision with other particles and with immersed heat transfer surfaces.

- size reduction due to superficial abrasion of the particles in the bed section.

Most studies [15] indicate a relation between attrition rate and the difference between fluidizing velocity u_{fl} and minimum fluidization velocity u_{mf}. The mass rate of production of fines is proportional to the impulse of the particle sand to the total bed weight.

Based hereon the following correlation has been tested with experiments in the TNO AFBB:

$$\Phi_{att} = K_{att} (u_{fl} - u_{mf})^{0.5} M_b^2 \cdot 3600 \qquad [kg/h]$$

where:

Φ_{att}	rate of production of fines	[kg/h]
K_{att}	attrition rate constant	[kg^{-1} $m^{-0.5}$ $s^{-0.5}$]
u_{fl}	fluidizing velocity	[m/s]
u_{mf}	minimum fluidizing velocity	[m/s]
M_b	mass of bed material	[kg]

The results of this model are presented in table 1.2.4.

TABLE 1.2.4
The attrition rate constant K_{att} vs. coal type.

	K_{att} [$kg^{-1}m^{-0.5}s^{-0.5}$]
Virginia	$11 \cdot 10^{-10}$ ± 8 %
Polish 3	$15 \cdot 10^{-10}$ ± 10 %
Polish 4	$13 \cdot 10^{-10}$ ± 12 %

For these experiments the fluidizing velocities has been varied from 1.5 to 2.6 m/sec and the bed height from 1.0 to 1.3 m.
The attrition rate constant is a direct indication of the friability of the bed material. The conclusion can be that the ash of the American Virginian coal is apparently better resistant to fracture and abrasion than the ash of the Polish coal.

With the use of the tested correlation a variety of parameters important for boiler design can be predicted depending on coal and limestone types as well as composition and particle size distribution. In this way parameters can be calculated, such as fly ash flow, bed ash flow, minimum ash content of the coal and minimum desired limestone flow required to keep bed height at the desired level, equilibrium between bed height and fly ash flow when no bed ash is discharged or when no limestone is supplied.

Based on the attrition model, calculations have been made of the ash flows to be expected from the 4 MW_{th} TNO AFBB research facility and from a 90 MW_{th}-boiler (see chapter 9) and confirmed by measured values.

1.2.9 Fuel supply, dry versus slurry feed

To feed an AFBC-boiler various overbed feed systems and pneumatic underbed feed systems are in use. For large capacity boilers pneumatic feed via a number of injection points is necessary to get the required distribution over the bed area. With a large number of injection points, the system gets rather complicated. Moreover the coal has to be prepared, i.e. crushed and dried, to prevent blockage in the feed lines.

For this reason it seems interesting to test a system that offers basically a potential for a simpler set up, based on a coal slurry that can be pumped into the boiler [9].
Depending on the diameter of the coal nozzles usually it will still be required to crush the coal. The slurry is made by mixing coal with water.
To transport the slurry two pump types are available, the monoscrew and the piston pump. In this case a piston type pump has been used. Together with the manufacturer, a pump, originally designed to pump concrete, has been adapted for this purpose. The complete unit, including splitters and coal nozzles, has been successfully tested in a closed circuit.

The required water content of the slurry appeared to be 23-25 %. At a lower water content the "viscosity" is getting too high, at higher contents sedimentation occurs very quickly. The right mixing of the coal and water appeared to be crucial to achieve a stable slurry; a blender was used for this purpose.
The slurry was fed into the boiler through one nozzle mounted in a side wall of the freeboard.

The combustion efficiency of coal fed as a slurry (fig. 1.2.12) is approximate 3 % lower than for the dry coal feed. This is thought to be the result of the lower combustion efficiency in the bed section, i.e. 87 % combustion without ash recycling (R = 0), while with dry coal feed a bed combustion of 90 % is measured [10].

This lower bed combustion is thought to be caused by the experimental setup of the slurry feed supply system. The total water content of the slurry has been in the range of 26% up to 30%, which is a little higher than the maximum of 25% originally planned. This may have had some effect on the experimental results, because the spray pattern of the coal nozzle will be effected by the lower "viscosity" of the slurry. Probably the lower combustion efficiency that has been recorded might be caused (to some extend) by this phenomenon. A possible explanation for this effect is that the spray of wet slurry produces more fine droplets than a "normal" slurry. If these fine droplets are elutriated instead of reaching the bed section, the residence time might be too short for complete combustion.

It is expected that the combustion characteristics with a coal slurry feed system will be equal to these of a dry system when the slurry supply is optimised with regard to water content and location of the nozzles. In that case slurry feed allows for a simplified layout of the feed system compared to pneumatic coal feed.

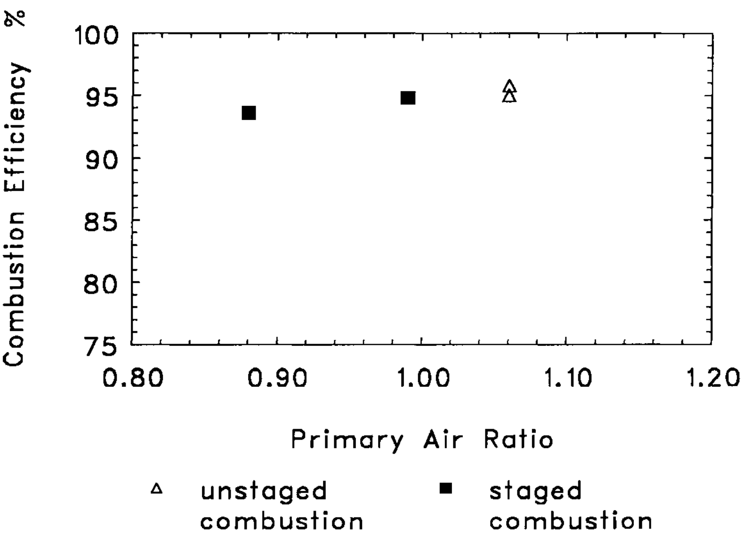


Figure 1.2.12. Combustion Efficiency of coal slurry with staged and unstaged combustion vs. primary air ratio.

1.3. SUPPLEMENTARY RESEARCH 1 MWth TU.

The experimental research described in this section is performed at the Twente University. The combustion efficiency related to the recycle ratio for various coal types was the first objective. Besides that special attention is given to the influence of some related process parameters: the fluidizing velocity, the bed temperature, the excess air level and the particle size of the coal used. Interim results of this research have been presented on conferences [18-20].

1.3.1. Research facility.

The layout of the experimental combustor and ancillaries is presented in figure 1.3.1. The 0.6 m square combustor is refractory lined, the height from the air distributor up to the freeboard cooling section is 4.4 m. The bed section is equipped with five rows of in-bed water tubes. Most of these tubes are hair pin bend, so up to 40 percent of the heat exchange surface can be withdrawn from the bed during operation. Above the freeboard section the flue gases are cooled down by a heat exchanger.

The bed material is a mixture of sand, limestone and ash of which the mean particle size is 1.6 mm. Later on also a finer bed material is used for NO_x reduction experiments. (mean particle size is 0.7 mm). The expanded bed height, measured by pressure probes inserted into the bed, was 1 m. During a test run bed material can be drained or sampled protected by inert gas. The flow rates of coal and limestone supplied to the bed are controlled by screw feeders and measured by weighing belts. They are injected together into the bed 0.15 m above the distributor plate.

An over-bed feed system is available for coal and limestone supply 0.5 m above the expanded bed surface.

The fluidizing air is supplied by a forced draft fan, metered and introduced into the bed through an air distributor plate with nozzles. Secondary air can be supplied through 4 over-fire air ports into the freeboard 2.2 m above the distributor plate.

Balanced draft at the freeboard is realized by an induced draft fan.

Downstream the freeboard cooler the flue gases pass a mechanical cyclone dust collector and a baghouse filter. The fly ash separated by the cyclone dust collector can be re-injected into the bed 0.15 m above the air distributor. The fly ash recycle system is designed in a way which enables a partial recycle of the collected fly ash, so that the recycled mass flow rate and the recycle ratio can be adjusted to a desired value. The recycled and drained fly ash mass flow rates are measured continuously.

A gas sampling unit for continuous flue gas sampling from various points in the freeboard and the stack is available. The main sample point is situated in the flue gas duct between the freeboard cooler and the cyclone dust collector.

1.3.2. Experimental program.

In order to compare various coal types, series of similar experiments are carried out. The range of experimental conditions is comprised in table 1.3.1. and the analysis and particle size of the coals investigated in table 1.3.2. The results of these experimental series are summarized in figures in the next sections.

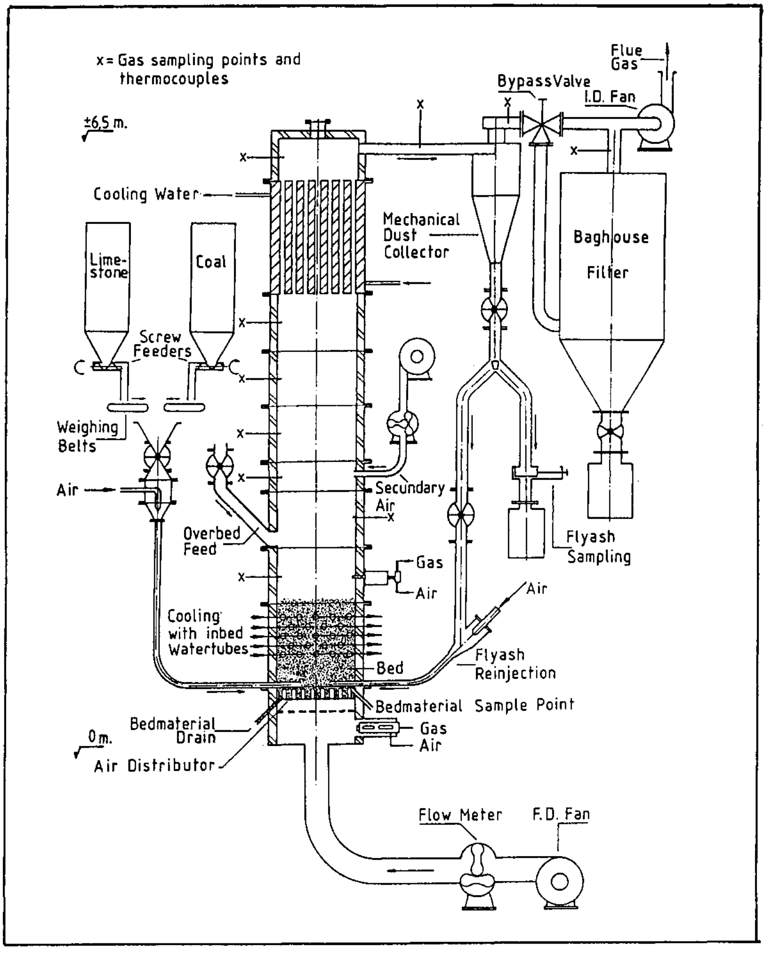


Figure 1.3.1. The fluidized bed combustor used for the experimental program at Twente University.

TABLE 1.3.1. Range of experimental conditions.

Bed temperature	750 - 900 °C
Fluidizing velocity	1.8 - 2.6 m/s
Exess air level	10 - 30 %
Bed height expanded	1.0 m
Recycle ratio	0 - 3
Coal feed rate	70 - 110 kg/h
Limestone feed rate	zero (except for section 1.3.9)
Mean particle size bed material	1.6 mm

TABLE 1.3.2. Analysis and particle size of coals investigated.

COAL TYPE		A		B		C	D		E	
		Polish		Belgian		German brown coal	Virginia USA		S. African Rietspruijt	
Ultimate analysis (dried fuel)										
carbon	%	74.13		80.85		64.22	75.30		70.11	
hydrogen	%	4.52		5.09		4.94	4.42		3.43	
oxygen	%	9.90		8.24		24.34	6.50		7.65	
nitrogen	%	0.71		1.76		0.60	1.51		1.58	
sulphur	%	0.78		0.79		0.48	1.56		1.03	
ash	%	9.96		3.27		5.42	10.59		16.19	
Proximate analysis (as received)										
volatile matter	%	29.73		23.73		44.06	27.24		22.68	
moisture	%	2.24		6.09		15.93	2.36		4.51	
fixed carbon	%	58.29		67.11		35.45	60.06		57.35	
Calorific value (gross)	MJ/kg	29.37		31.2		21.6	30.1		26.4	
Size range	mm	0-10	4-10	0-4	2-10	0-4	0-12	3-12	0-12	3-12
mean diamter	mm	2.8	6.5	1.8	6.8	1.0	3.4	6.3	3.8	7.7
< 1mm	%	27	0	13.8	0.3	63.1	24.0	0	23.7	0
Symbol used in figures		▼	∇	♦	◊	+	■	□	●	○

1.3.3. Combustion efficiency versus coal type.

For the five above mentioned coal types the combustion efficiencies as a function of the recycle ratio is presented in figure 1.3.2. It must be noted that for the values of the combustion efficiencies shown here no limestone was supplied.

As can be seen from this figure the combustion efficiency measured at a certain combustor differs for various coal types, with the largest differences for the process without recycling fly ash.

1.3.4. Combustion efficiency versus excess air level.

The influence of the excess air level on the combustion efficiencies at various recycle ratios was investigated for the excess air levels from 10 % up to 30 %.

Here only the results of the experiments without recycling are presented in figure 1.3.3. Changes of the excess air level within this range does not give remarkable differences in combustion efficiencies. When fly ash is recycled the experimental results show still no relation with the excess air level.

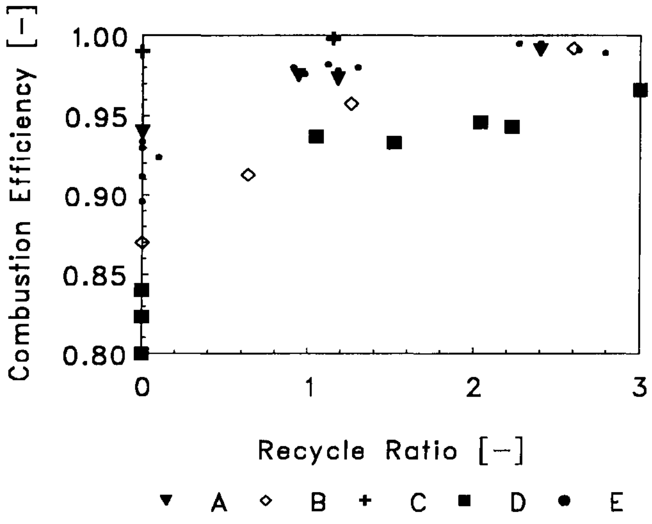


Figure 1.3.2. Combustion effiency of different coal types as a function of recycle ratio, 20 % excess air, 850 °C. The symbols represent different coal types, see table 1.3.2.

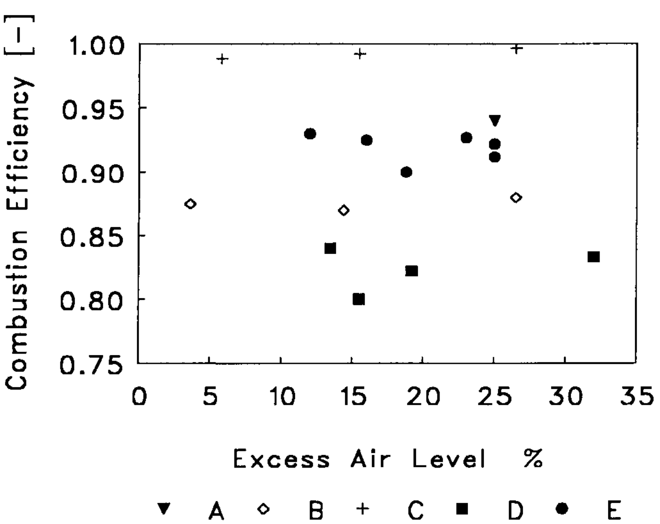


Figure 1.3.3. Combustion efficiency as a function of excess air level, no recycling, 850 °C, for different coal types. See table 1.3.2. for the symbols.

1.3.5. Combustion efficiency versus bed temperature.

The bed temperature is changed in two steps from 750 °C to 900 °C at which the combustion efficiency is measured for various recycle ratios. In figure 1.3.4 some results are given for Belgian coal. When no fly ash is recycled the influence of the bed temperate is larger than for the process with recycling, higher recycle ratios show less influence of the bed temperature. All coal types, except German Brown coal show the same trend. At recycle ratios higher than 1.5 no clear dependence on bed temperature is measured.

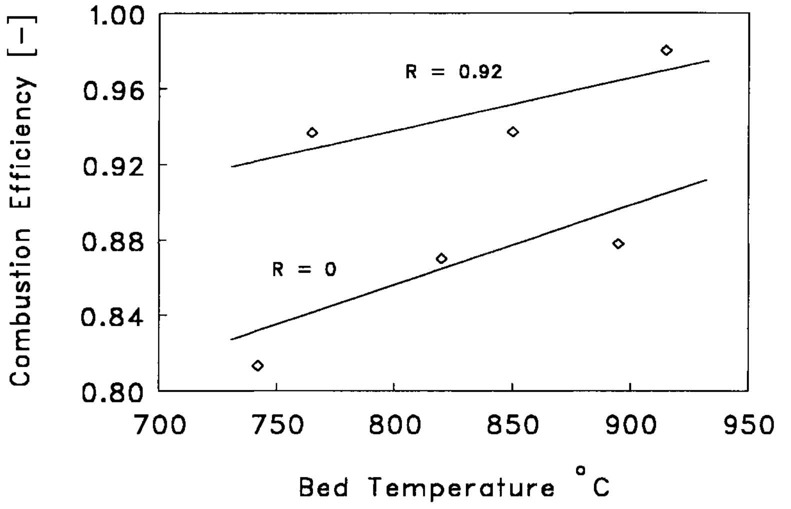


Figure 1.3.4. Combustion efficiency as a function of bed temperature, Belgian coal, mean particle size 6.8 mm, excess air 18 %.

1.3.6. Coal mean particle size.

For some coal types mentioned in table 1.3.1. two different mean particle sizes are investigated; one size range with fines and one without fines. (German brown coal only with fines). Figures 1.3.5 and 1.3.6. show results from experiments illustrating that for this parameter the coal can be divided into two groups. The mean particle size is important for Virginia coal but not for South African coal. Polish coal shows the same behaviour as Virginia coal, Belgian coal the same as South African coal. Visual observation of Belgian coal in a small fluidized bed combustor indicates a fast fragmentation of coal particles soon after that they entered the bed, during devolatilization. Such a fragmentation is not observed during similar experiments with Polish coal.

The results in these two figures are obtained from experiments at a bed temperature 850 °C and excess air levels from 10 % to 30 %, as the influence of the latter can be disregarded.

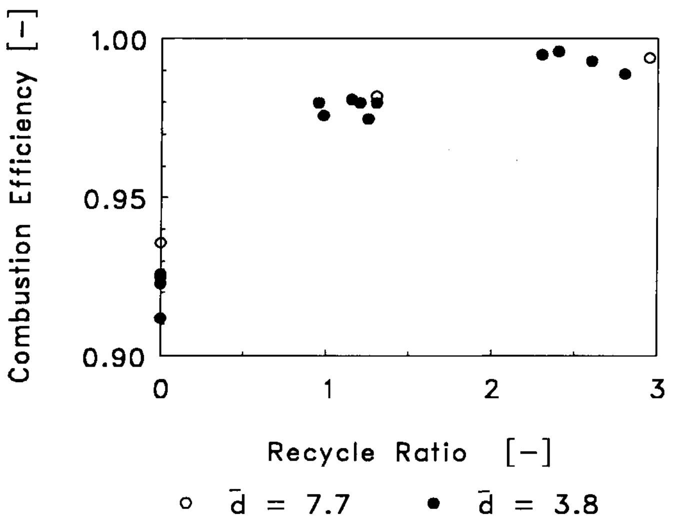


Figure 1.3.5. Combustion efficiency as a function of recycle ratio for two mean coal particle sizes, South African coal.

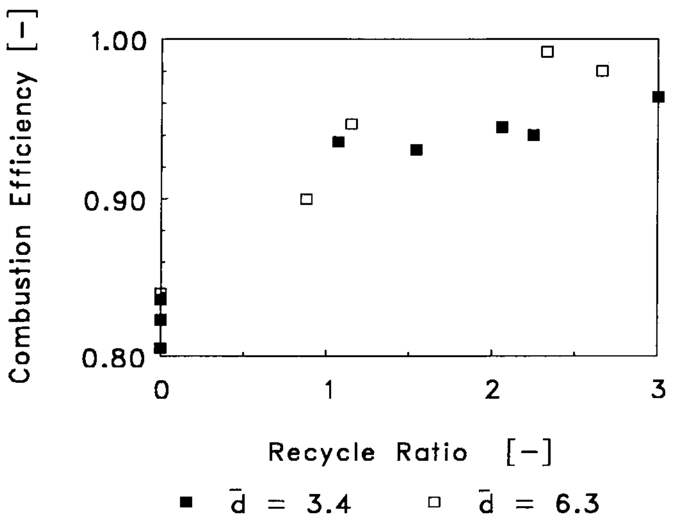


Figure 1.3.6. Combustion efficiency as a function of recycle ratio for two mean coal particle sizes, Virginia coal.

1.3.7. Freeboard temperature.

For most experiments the temperature of the fluidized bed is the process parameter which is controlled at a constant value. The appearing freeboard temperature will differ from the bed temperature as it depends on the heat production and heat transfer occurring under given process conditions.

The measured differences between freeboard temperature and bed temperature as a function of the recycle ratio for various coal types is presented in figure 1.3.7. In this experimental facility the freeboard temperature appears to be lower than the bed temperature for all experiments without recycling. Recycling fly ash results in freeboard temperatures higher than the bed temperature, caused by the combustion in the freeboard section of fine coal and char particles present in the fly ash .

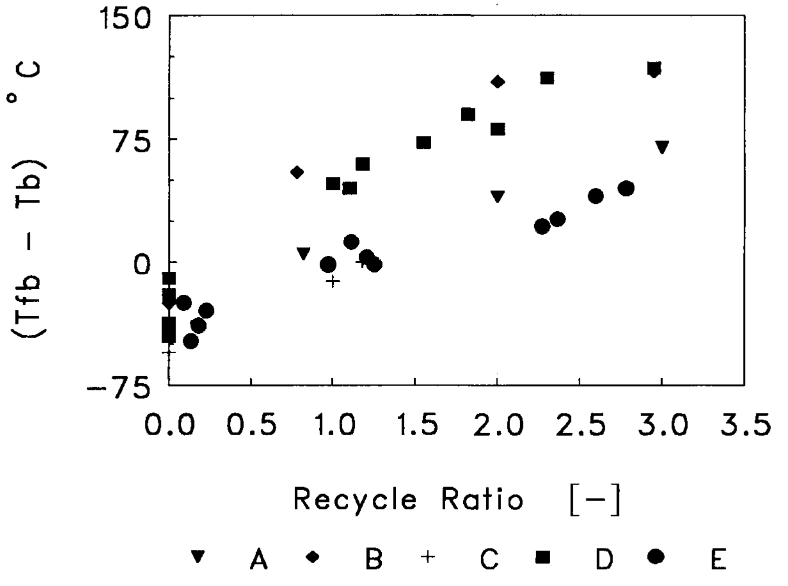


Figure 1.3.7. Difference between freeboard and bed temperature versus the recycle ratio. The symbols indicate the coal type according table 1.3.2

The differences between the two temperatures depend also on coal type, the Belgian coal and the Virginia coal produce a higher freeboard temperature than the other three coal types. As will be shown in the next section the carbon content in the recycled fly ash is in the same way related to the freeboard temperature for these coal types.

1.3.8. Carbon content in fly ashes.

As for a certain recycle ratio the combustion efficiency is different for various coal types one can expect that the amount of fly ashes and the carbon content of these ashes depend on coal type, when all burned in the same combustor, at same process conditions. The measured carbon contents of the fly ash from the cyclone dust separator is presented in figure 1.3.8, and those of the baghouse filter in figure 1.3.9. These results are obtained from experiments at a bed temperature of 850 °C and at measured excess air levels from 5 % to 35 %.
A comparison of these two figures indicates that in general the carbon content of the cyclone fly ash is higher than that of the baghouse filter. In both figures a relation with coal type and recycle ratio is seen.
For the whole series of experiments of which the results are presented in the two figures showed above the mean values of the ash flow rates are calculated for each coal type from all experiments together, with the objective to give some information about the magnitude of these mass flow rates. In figure 1.3.10 a diagram is given of these mean values of ash from coal supplied and ash (without carbon, but with elutriated bed material) discharged from the cyclone separator and the baghouse filter.

Comparing figures 1.3.8 and 1.3.10 it can bee seen that e.g. for Belgian coal the carbon content is rather high, however the drained fly ash mass flow rates are relatively small. And for S. African coal the carbon content is much lower together with much higher drained mass flow rates of ashes.

From figure 1.3.10, though it are mean values from a range of experiments, some information can be obtained about the ash balance of each coal. For Polish, German and S.African coal the ash supplied is somewhat higher than the ash drained, that means that part of this ash remains in the bed, resulting in an increase of bed height; and for the other two the bed height will be decreased a few at the end of the series of experiments. However when also limestone is supplied for sulfur retention for most combustion processes the bed inventory will increase.

In this figure also an indication is given of the discharged mass flow rates of ashes from the cyclone separator compared to that from the baghouse filter. For these series of experiments the mean value of recycled mass flow rate is of the same magnitude as the mass flow rate of the coal supplied, this is for the above described experiments about 90 kg/hour, and thus much larger than the drained mass flow rates.

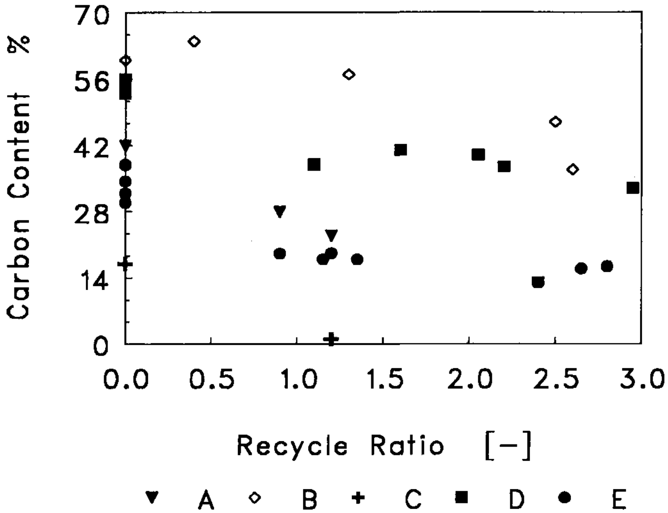


Figure 1.3.8. Carbon content of fly ashes from cyclone dust collector as a function of the recycle ratio.

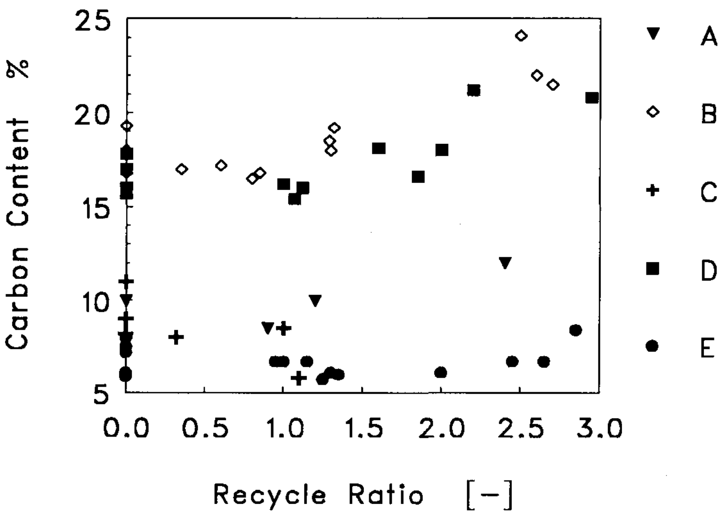
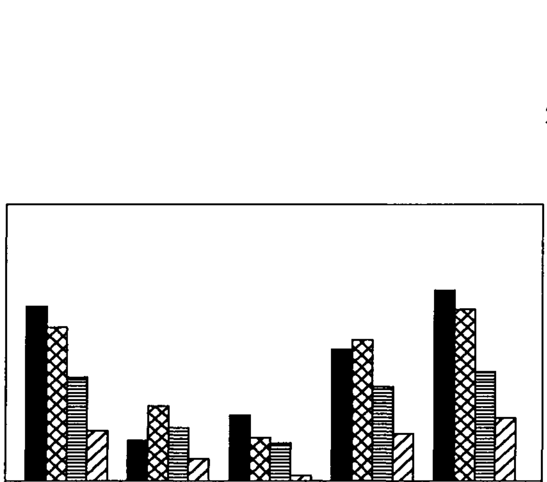


Figure 1.3.9. Carbon content of fly ashes from baghouse filter as a function of the recycle ratio. The symbols represent different coal types from table 1.3.2.



	Pol.	Bel.	Br.k	Virg.	S. A.
Coal	12.6	2.9	4.7	9.5	13.8
Cyc.+ Bagh.	11.1	5.4	3.1	10.2	12.4
Cyclone	7.5	3.8	2.7	6.8	7.9
Baghouse	3.6	1.6	0.4	3.4	4.5

Figure 1.3.10. Mean mass flow rates of the whole series of experiments of ashes (without carbon, but with elutriated bed material) for each coal type in kg/hour.

1.3.9 Staged combustion conditions. Freeboard temperature.

As is showed in the next chapter lower NO_x emissions can be achieved by staged combustion. For this the air used for combustion is split into a primary air flow supplied to the bed through the air distributor and a secondary air flow entering the freeboard section above the bed, as indicated in figure 1.3.1 In continuation of the research described above special attention is given to this subject with the main objective to optimise the process conditions for minimal emission levels.

Within the context of this research the range of coal types is extended with some other coals, for which the combustion efficiencies are also measured under various staged combustion conditions, but with limestone supply for sulfur retention. Some results of this investigation is presented here.

During these experiments the fluidizing velocity is set to a fixed value of 2 m/s so the primary air flow rate supplied to the bed has a fixed value, and the mass flow rate of the coal is increased to maintain the same excess air level for the sum of the primary and secondary air flow. The extra heat produced is extracted from the bed by changing the amount of heat transfer tubes in the bed, so the bed temperature is kept at a fixed value. The capacity of the facility increases in this way, and also in the freeboard more heat is produced. With Polish coal not all experiments originally planned could be performed as the freeboard temperature increased to over 950 °C when the primary air stoichiometry was turned down below 80 percent. For this reason a second freeboard cooler was installed for some test runs. This cooler was situated 3 m above the air distributor plate and consists of three rows of total 14 water tubes.

Results from experiments executed with British Marine coal are presented here for a range of staged combustion conditions; for this coal type the freeboard temperature is not too high for the lower air stoichiometries at the experiments without extra freeboard cooling. The limestone used for sulfur retention is hard, the mean diameter of which is 1.6 mm. Analysis of coal type used for these experiments are mentioned in table 1.3.3.

TABLE 1.3.3. Analysis of British Marine coal.

Coal Type		British Marine Coal
Ultimate analysis (dried fuel)	[%]	
Carbon	C	81.84
Hydrogen	H	4.54
Oxygen (by diff.)	O	3.20
Nitrogen	N	1.39
Sulphur	S	0.95
Proximate analysis (as received)	[%]	
Volatile matter		23.40
Moisture		1.00
Fixed Carbon		67.65
Ash		7.95
Calorific value (gross)	[MJ/kg]	32.06
SOXth	[g/GJ]	589
NOXth	[g/GJ]	1510
Size range	[mm]	0-12
Mean diameter	[mm]	2.50
< 1 mm	[%]	22.0

In figure 1.3.11 the combustion efficiency is presented as a function of the primary air ratio for staged combustion conditions for two series of experiments, one without freeboard cooler and one with freeboard cooler. The total air ratio has been 110 percent as a base value for all these experiments, so the results given for the air ratio = 1.1 in this figure are from experiments for which the secondary air flow is zero.

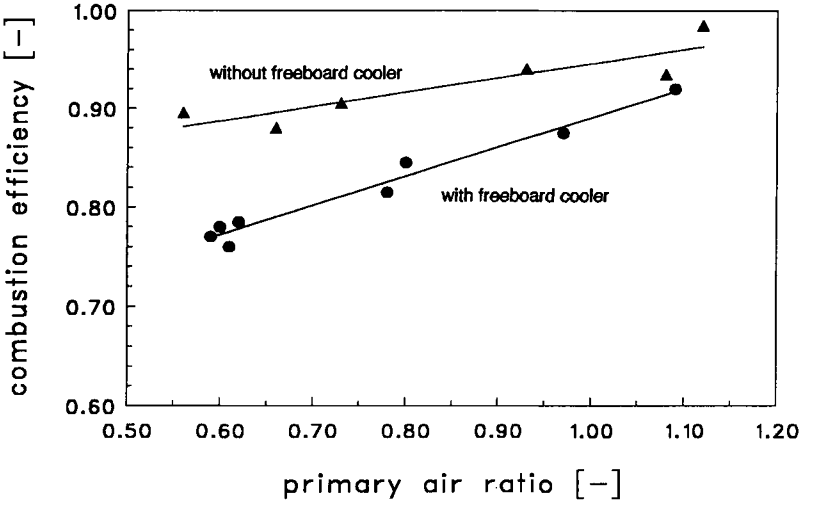


Figure 1.3.11. Influence of primary air ratio on combustion efficiency at staged combustion. Total air ratio: 1.1; bed temperature: 850 °C; R = 1.5; Ca/S = 1.5.

Next to it the bed temperature has been 850 °C, the recycle ratio R = 1.5 and Calcium to sulfur ratio is set on 1.5 as mean process conditions.

This figure demonstrates that at staged combustion conditions the combustion efficiencies decreases with decreasing primary air ratios. This may be caused by the fact that more coal is supplied to the bed with as a consequence lower oxygen concentration in the bed, lower carbon conversion rates and higher carbon contents in the bed resulting in more attrition loss of carbon. Besides this the residence time of carbon particles in the freeboard is reduced by the enhanced flow velocity in the freeboard as a consequence of keeping the fluidizing velocity in the bed at a fixed value.

At moderate staged combustion conditions the combustion efficiency can still attain rather high values, for the experiments without freeboard cooler.

Injection of secondary air in the freeboard improves the mixing of gases and solids and thus has a positive influence on gas-solid reaction like char combustion.

Next to it figure 1.3.11 shows a remarkable difference in combustion efficiencies between experiments with and without an extra freeboard cooler, caused by differences in freeboard temperatures. In general for experiments without extra freeboard cooler the freeboard temperatures are about 50 °C higher then the bed temperature and for those with extra freeboard cooler about 70 °C lower than the bed temperature.

At these lower temperatures and for small carbon particles the reaction rate is kinetically controlled so temperature changes affect the combustion efficiency since temperature has a larger influence on kinetics than on oxygen diffusion.

1.3.10 Particle size bed material

During experiments with other research objectives it turned out that particle size of bed material is a parameter influencing the NOx emission level at coal combustion in fluidized beds. In chapter 2 this subject is described and results of experiments are given; and in section 2.4.2 an explanation for this phenomenon based on different heat and mass transfer conditions is presented.

From experiments no significant differences in combustion efficiencies for one coal type using two different bed particle sizes is found. The bed material mean particle sizes used is equivalent to those described in chapter 2: 0.7 and 1.6 mm.

However during other experiments with pellets made from waste carbon a remarkable difference in combustion behaviour is observed caused by the bed particle size. This waste carbon is a by-product of the production process of carbon anodes used in aluminium melting industries and consists of fine particles low reactive carbon. This carbon originates from petcoke and contains less than 7.52 percent volatiles. Combustion experiments with these fine particles turned out that it is not possible to maintain constant process conditions, the bed temperature decreases while the freeboard temperature increases to unacceptable high values, as these fine particle are elutriated out of the bed and burn mainly in the freeboard. For that reason this fine material is mixed with hydrated lime and pelletized during which also steam is supplied. The diameter of these cylindrical pellets is 4.8 mm and the length is zero to three times the diameter.

These pellets are investigated in the above described experimental facility of Twente.This research has demonstrated that a rather good combustion efficiency can be obtained and stable combustion behaviour is possible only with the use of coarse bed material. In hot drained bed material from these experiments glowing pellets were observed.

It turned out that for this fuel the heat and mass transfer rates reaches critical values for the finer bed particle sizes so that stable combustion of the pellets could not be obtained.

1.4. SIMULATION AND ANALYSIS OF THE COMBUSTION EFFICIENCY

1.4.1 Description model

With the objective to get more insight in the influence of various combustion parameters on the combustion efficiency for different coal types, a calculation is presented here for a simplified flow scheme for the combustion process with fly ash recycling. At the same time some related parameters will be discussed.

The experimental facilities of which the results will be analyzed and discussed in this chapter are described in section 1.2 and 1.3. Both facilities can be represented by the same flow diagram shown in figure 1.4.1.

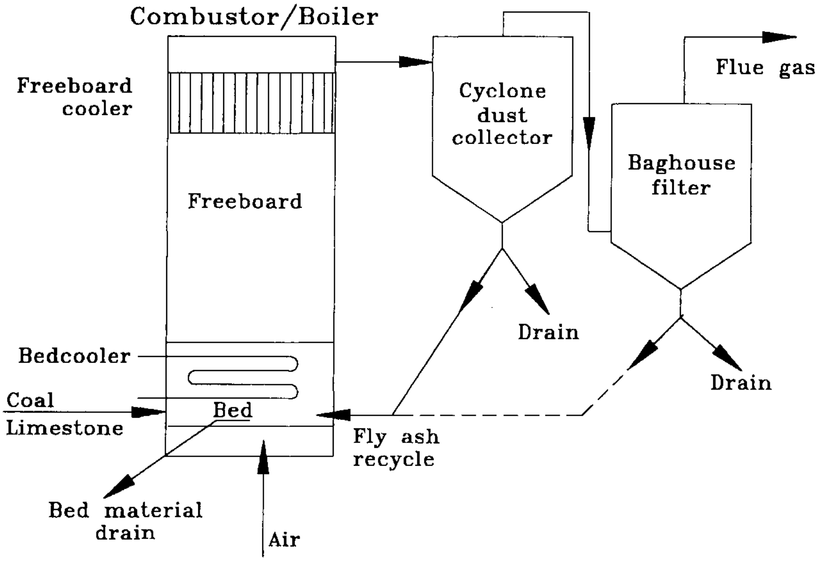


Figure 1.4.1 Flow diagram experimental facilities at TNO and TU

These experimental facilities have a system to recycle partly or totally the fly ash collected by the cyclone dust collector. Only the TNO unit has a system to recycle the fly ash from the baghouse filter. The ultimate difference of recycling both streams of fly ash compared to the recycle of only fly ash from the cyclone collector depends on the separation efficiency of the cyclone dust collector used in the latter system. For that reason these installations will be further simplified to a combustor, a cyclone dust collector and a fly ash recycle system, a flow diagram of which is presented in figure 1.4.2. In this scheme the solids mass flow rates are presented.

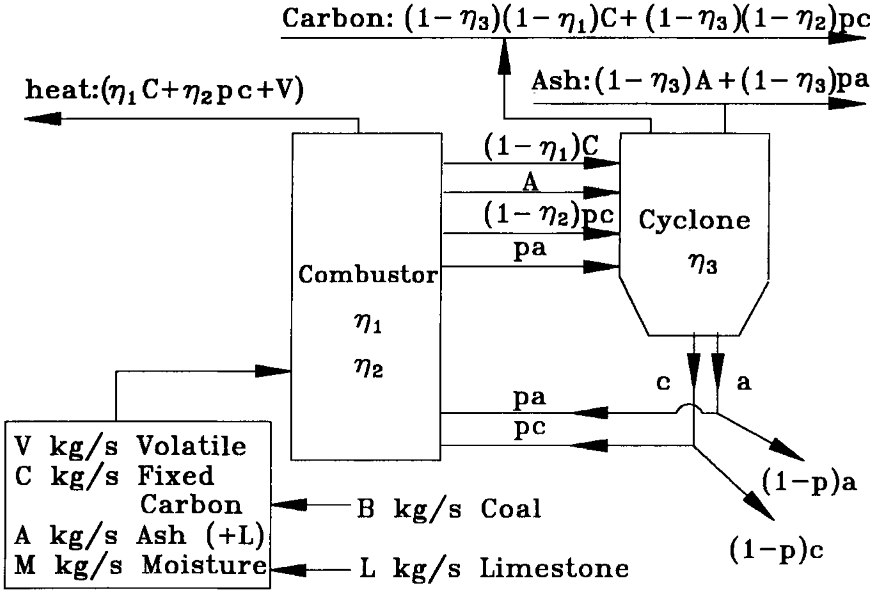


Figure 1.4.2. Scheme for balances of solids mass flow rates

The mass flow rate of coal, B kg/s, entering the system is divided into four separate mass flows, according to the proximate analysis of the coal. For the present no limestone is supplied. Heat is produced by the combustion of volatiles and fixed carbon. The ashes and unburned carbon comes out of the system, partly together with the flue gas and partly as a drain from the solids stream out of the cyclone separator. The mass flow of carbon collected by the cyclone separator is proposed to be c kg/s and that of the ash is a kg/s, both of which a fraction p is recycled into the combustor.

The following assumptions are made for the process to simplify the calculations:

- The volatiles burn completely. During the experiments the unburned gaseous components in the flue gases were sufficiently low to be neglected.
- The mass flow of ashes and unburned carbon drained from the bed is also neglected, as the changes in bed height were too small to establish a difference between the beginning and the end of an experimental run of some hours, beside the carbon content in the bed is only some percents.
- The once-through carbon conversion efficiency for the bed and the freeboard is η_1.
- The carbon conversion efficiency for the bed and the freeboard of recirculating solid carbon is η_2.
- The separation efficiency of the cyclone dust collector is η_3.

The carbon balance for this system is:

$$C = \eta_1 C + \eta_2 pc + (1-p)c + (1-\eta_3)(1-\eta_1)C + (1-\eta_3)(1-\eta_2)pc$$

or

$$c = \frac{\eta_3(1-\eta_1)C}{1-\eta_3(1-\eta_2)p} \tag{1.4.1}$$

And the ash balance is:

$$A = (1-\eta_3)A + (1-\eta_3)pa + (1-p)a$$

or

$$a = \frac{\eta_3 A}{(1-\eta_3 p)} \tag{1.4.2}$$

The combustion efficiency, defined in section 1, is for this system (see figure 1.4.2):

$$\eta_c = \frac{H_o B - H_c\{(1-\eta_3)(1-\eta_1)C + (1-\eta_3)(1-\eta_2)pc + (1-p)c\}}{H_o B} \tag{1.4.3}$$

And for the case when there is no recycling of fly ash, this formula can be reduced to:

$$\eta_c^o = \frac{H_o B - H_c(1-\eta_1)C}{H_o B}$$

When this η_c^o is known from experiments the value of η_1 can be calculated:

$$\eta_1 = 1 - \frac{H_o B}{H_c C}(1-\eta_c^o) \tag{1.4.4}$$

This once-through carbon conversion η_1 and also the carbon conversion of the recirculating solids η_2 will depend on the recycle ratio and the process conditions in the combustor e.g. oxygen concentration and the freeboard temperature are different for various recycle ratios. However within the context of this calculations these two carbon conversion efficiencies are assumed to have a constant value for one coal type under specified process conditions.

The recycle ratio is, according the definition mentioned in section 1:

$$R = \frac{p(a+c)}{B} \qquad (1.4.5)$$

1.4.2 Simulations and analyses

The equations (1.4.1) to (1.4.5) will be used to simulate the combustion efficiency η_c for various values of η_c^o , η_2 , η_3 , and p for a specific coal type, and also to compare different coal types with relation to the combustion efficiencies.

For the simulations described below a Polish coal is used, of which the proximate analysis is:

(as received, weight %)	
volatile matter	V = 28.8
fixed carbon	C = 61.8
ash	A = 8.0
moisture	M = 1.4
Gross calorific value	H_o = 30.71 MJ/kg
Calorific value of carbon	H_c = 34.0 MJ/kg

The amount of coal can have any given value and is set at B = 100 kg/s and thus A = 8.0 kg/s, etc.

The separation efficiency of the cyclone dust collector is set at η_3 = 0.95.

The carbon conversion efficiency of the recycled fly ash is supposed to be $\eta_2 = 0.4*\eta_1$ assuming such a relation between these two parameters for a certain coal.

The first simulation is made for three different values of the combustion efficiency of the system without recycling: η_c^o = 0.85; 0.90 and 0.95. For each of these values η_1 is calculated with formula (1.4.4).

1.4.2.1 Coal type, recycle ratio

For the process conditions, set above, the combustion efficiencies as a function of the recycle ratio R are simulated in figure 1.4.3. (The points on the curves indicate the calculated values for p = 0 - 0.5 with steps of 0.1 and for p = 0.5 - 1 with step of 0.05).

The curves in this figure show that the combustion efficiency increases with the increase in recycle ratio R as expected and known from experimental results described in previous sections.

Curves with a lower once-through efficiency (η_c^o) show a steeper increase than those with a higher value of this. That is mainly caused by the higher carbon content in the recycled fly ash flow, as this amount of carbon depends on η_1 and thus on η_c^o, see formulas (1.4.1), (1.4.2) and (1.4.4), while the amount of ash in this recycled flow is independent on the once-through efficiency η_c^o. For these once-through efficiencies the amount of carbon c for the lowest curve is about three times the amount of carbon for the upper curve.

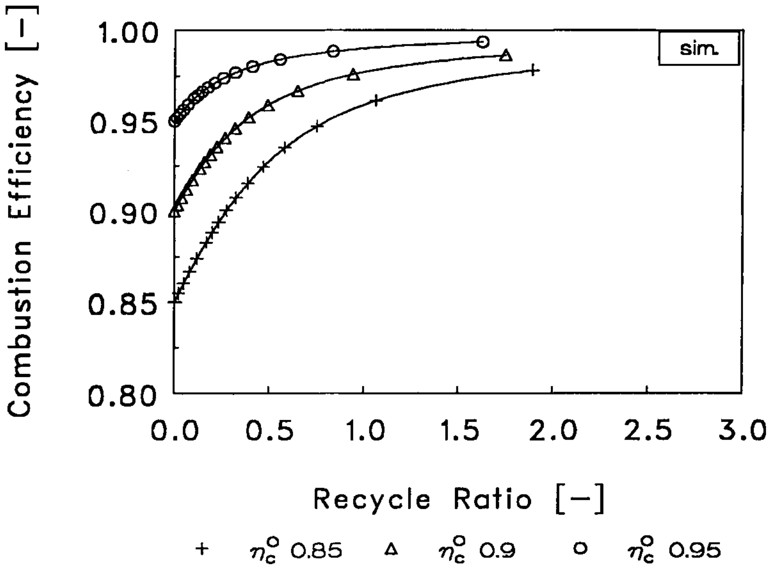


Figure 1.4.3 Simulation of combustion efficiency for three once-through efficiencies

This means that for coals with lower once-through efficiencies recycling fly ash can be a very effective option to obtain a sufficiently high combustion efficiency.

The end of each curve is the condition for which no fly ash is discharged from the cyclone flow, so p = 1 and the recycle flow rate has a maximum value for the given cyclone separation efficiency. The combustion efficiency which can be obtained in a given combustor depends on the carbon conversion efficiencies η_1 and η_2 related to coal properties.

1.4.2.2 Combustor design

With relation to this simulation some more comments can be given. The three curves in the figure can be seen as curves of three different coal types, as discussed above, but also as curves of one coal type for three different combustors. For this second situation the differences in the curves are related with differences in design and/or operation conditions. The carbon conversion efficiency η_1 depends on the properties and behaviour of the bed, like bed height, bed temperature, particle size bed material, bubble size and amount of coal supply points per square meter. In a combustor the carbon conversion efficiency η_2 (and to some extend also η_1) depends on freeboard properties like height, temperature and degree of gas mixing.

To illustrate the relative influence of η_2, the next simulation step is the calculation of the combustion efficiencies for the same coal, but for three different carbon conversion efficiencies of the recycled solids: $\eta_2 = 0.2*\eta_1$; $0.3*\eta_1$ and $0.4*\eta_1$.

The other parameter which must be set to a fixed value are: the cyclone separation eficiency $\eta_3 = 0.95$ and the combustion efficiency without recycling $\eta_c^o = 0.9$. Thus η_1 has a fixed value, calculated from η_c^o so the influence of η_2 the carbon conversion efficiency of recycled fly ash is simulated and shown in figure 1.4.4. In this situation doubling the recycle ratio has almost the same effect as doubling the carbon conversion efficiency η_2.

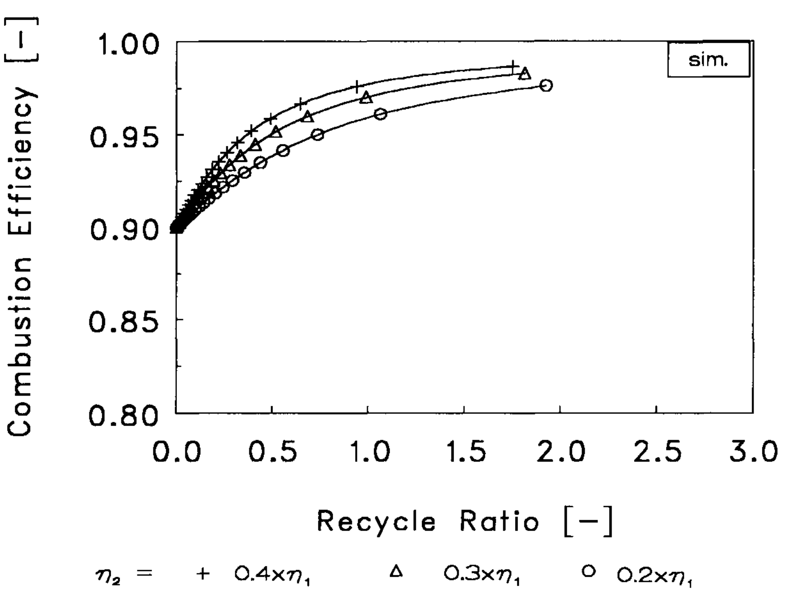


Figure 1.4.4. Simulation of the combustion efficiency for three different values of the carbon conversion efficiency of recycle fly ash.

1.4.2.3 Ash content in coal, limestone

The third simulation is related to the influence of the amount incombustible solids, like ash from coal and limestone for sulphur retention, supplied to the combustor. For this simulation the amount of coal is kept equal to that in the previous simulations: B = 100 kg/s, likewise η_3 = 0.95, η_2 = 0.4*η_1, and η_c° = 0.90. Now a simulation will be made for which also limestone is supplied to the combustor: L = 0; 7 and 12 kg/s so for A = 8 kg/s (equal to the previous simulations), A = 8 + 7 = 15 kg/s and A = 8 + 12 = 20 kg/s the process is simulated. For this it is assumed that the behaviour of limestone is equal to the behaviour of the fly ash from the coal.

The three curves are presented in figure 1.4.5. From this simulation can be concluded that equal ultimate combustion efficiencies can be reached, however at large differences in recycle ratios. That is caused by the amount of recycled ash and limestone which increases, while the amount of recycled carbon c stays at the same value for a certain recycle ratio. Thus limestone supply has a negative influence on the combustion efficiency, this influence can be reduced in a real combustor by the application of a soft and fine limestone mostly collected by the baghouse filter and not by the cyclone separator so that it is not recycled or a hard and coarse limestone type most of which remains in the bed. The influence on sulphur retention for these options is not considered here.

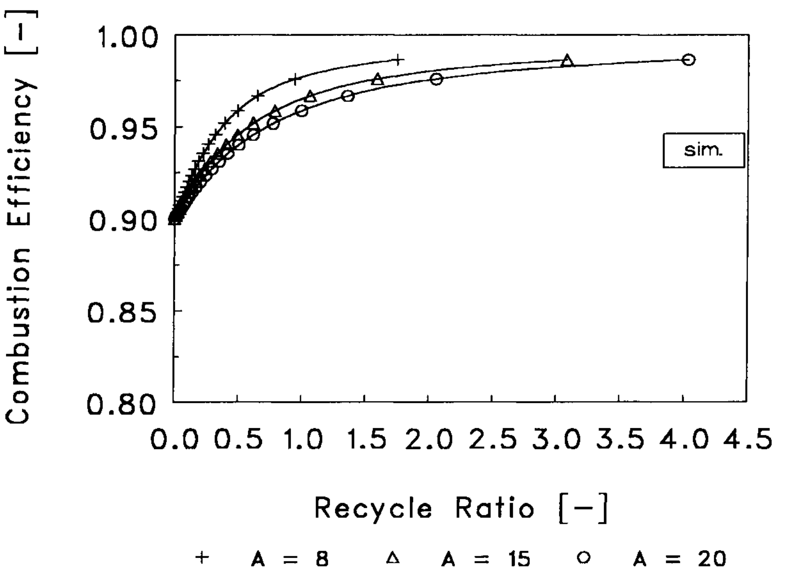


Figure 1.4.5 Simulation of the combustion efficiency for three different amounts of solids/ashes supplied to the combustor

An equivalent behaviour will be found when coals with different ash content is compared. In order to reach a sufficiently high combustion efficiency for the combustion of coals with high ash content two options are available.
Firstly a high recycle ratio can be chosen with the need of a large recycle system and high power consumption by the draft fan of the combustor. However in a fluidized bed combustor with fly ash recycling the fly ash coming out of the combustor is cooled at the end of the freeboard together with the flue gases. The temperature of the reinjected fly ash is much lower then the temperature of the bed. Thus, depending on the amount of fly ash, heat is rejected from the bed and transported to the freeboard by the fly ash particles, resulting in a change in heat balance of the boiler.
Secondly the carbon conversion efficiencies can be improved e.g. by the choice of a higher freeboard so that the residence time in the combustor of the fly ash particles is increased or by applying a higher freeboard temperature.

A simulation can be made for a real combustor for which various process parameters are given fitted values. Doing this for the Polish coal, investigated in the TNO facility of which the results are presented in figure 1.2.3 of that section, reveals that the curve in that figure can be simulated with the following values of the process parameters: $\eta_c^o = 0.9$; $\eta_2 = 0.5*\eta_1$; $\eta_3 = 0.9$ and A = 15 kg/s (ash and limestone)..

1.4.2.4 Cyclone separation efficiency

For all previous simulations the efficiency of the cyclone separator was set on $\eta_3 = 0.95$, the value of which is close to the quality of the cyclone separators of the experimental facilities, described in previous sections.
To indicate the influence of this separation quality a second value for $\eta_3 = 0.85$ is simulated. For this simulation $\eta_c^o = 0.9$, $\eta_2 = 0.4*\eta_1$, and A = 8 kg/s, the results of which are presented in figure 1.4.6.

As could be expected, the maximum of the combustion efficiency and the maximum of the recycle ratio which can be obtained are related to the cyclone separation efficiency of each curve as those maximum points depend on the amount of fly ash leaving the recycle system at the top of the cyclone together with the flue gases. The two curves coincide as for a certain value of the recycle ratio the amount of fly ash and unburned carbon leaving the system are for both processes equivalent. The next two simulations will give more insight in these relations, for which the remaining process conditions have the same values.

The maximum recycle ratio that can be obtained for various separation efficiencies of the cyclone dust collector is simulated and presented in figure 1.4.7.
And likewise the maximum combustion efficiency as a function of the separation efficiency is shown in figure 1.4.8. For these simulations $\eta_c^o = 0,9$, $\eta_2 = 0,4*\eta_1$, and the values of A is given in the figures.

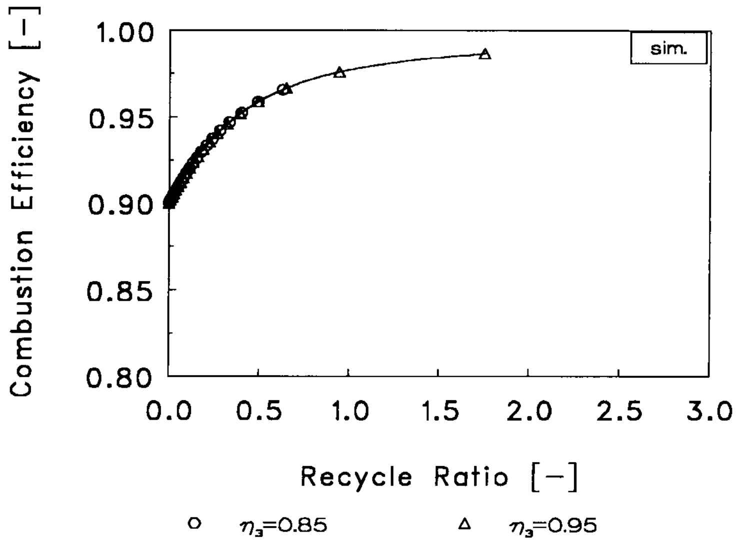


Figure 1.4.6. Simulation of the influence of the cyclone separation efficiency g_3 on the combustion efficiency

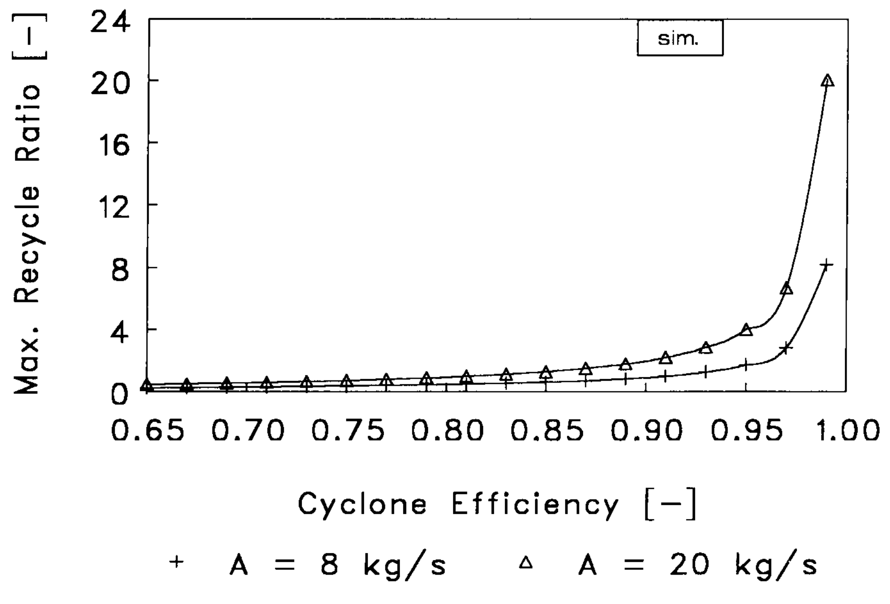


Figure 1.4.7. Values of the maximum recycle ratios as a function of cyclone efficiency for two values of ash/limestone supply

Figure 1.4.7 shows that the recycle ratio and thus the recycle flow rate can reach high values for higher cyclone efficiencies. Furthermore the level of this curve depends on the ash content of the coal and the amount of other solids (limestone) entering the combuster as already discussed at figure 1.4.5.

The level of the curve in figure 1.4.8 depends on coal properties and on the design of the combustor, the influence of these are demonstrated in figures 1.4.3. and 1.4.4.
The quality of the cyclone separator is important for the combustion efficiency when only cyclone fly ash is recycled. Recycled fly ash from the baghouse filter is less effective when the carbon content of this fly ash is lower than that of cyclone fly ash as appers from experiments in the previous section.

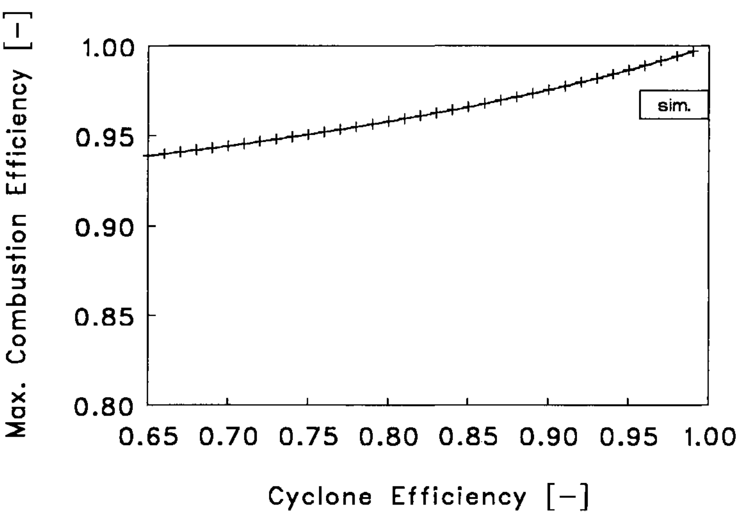


Figure 1.4.8. Maximum combustion efficiency attained as a function of the cyclone efficiency, A = 8 kg/s.

1.4.2.5 Carbon content in fly ash

Now another quantity of the simulated process will be illustrated, the carbon content of fly ashes. From the scheme of the process (figure 1.4.2) it can be seen that the carbon content in the recycled fly ash is:

$$C_R = \frac{c}{a + c}$$

As for the simulations demonstrated in this section the carbon concentration of the mass flows entering or leaving the cyclone is supposed to be equivalent to each other, this formula applies also for the fly ash leaving the top of the cyclone and consequently entering and leaving the bag house filter. That means also that the values simulated below must be seen as mean values of both ash flows leaving the system.

Simulated percentages of carbon concentrations as a function of the recycle ratio are presented in figure 1.4.9. In this simulation the following conditions are taken:
$\eta_1 = 0.9$, $\eta_2 = 0.4*\eta_1$, $\eta_3 = 0.95$ for A = 8 and A = 20 kg/s.

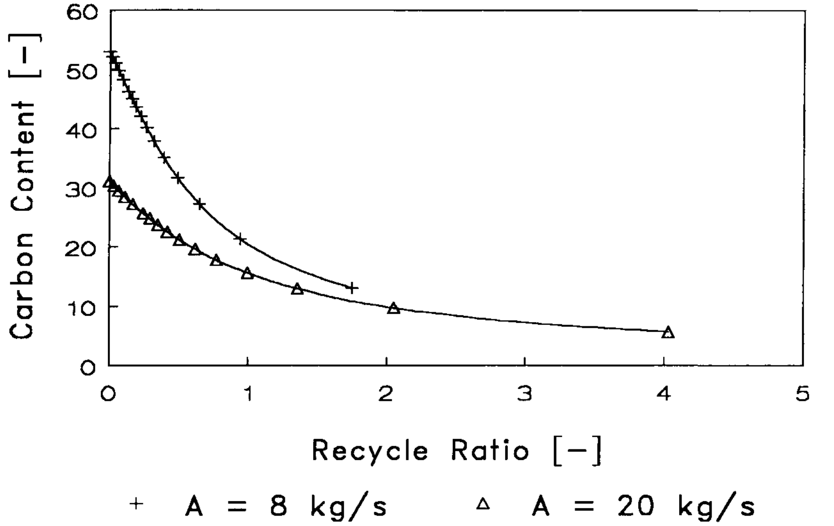


Figure 1.4.9. Simulation of carbon content of fly ash for two values of ash/limestone supply as a function of recycle ratio

This simulation demonstrates that the percentage of carbon in fly ash is lowered by recycling. Higher amouts of ash in coal or other inerts like limestone results in lower carbon contents.

Comparing this simulation with experimental results shown in figures 1.3.8 and 1.3.9 of a previous section, the same trend is seen in figure 1.3.8 for the cyclone fly ash, not for the fly ash from the baghouse filter. For real combustors the carbon content depends not only on the combustion process itself and the amount of inert solids entering and leaving the system but e.g. also on the particle size distributions of carbon, ashes and limestone.

1.5. HEAT TRANSFER

Due to the gas-solid interaction, typical for fluidized beds, very high heat transfer rates are achieved compared to heat transfer in the gas phase alone [11].
Heat transfer in a fluidized bed combustor consists of particle convection, gas convection and radiation.
The most important process parameters influencing the heat transfer in fluidized beds, are:

- bed particle size and particle size distribution;
- bed temperature;
- bed voidage and fluidizing velocity.

Experiments have been carried out to determine the correlation between these parameters and heat transfer. The results are compared to calculations based on theoretical models.

1.5.1 Particle size and particle size distribution.

The mechanism of heat transfer from particles to a wall is based on heat transport via the contact surface between the particles and tube wall and, even more important, via the narrow layer of flue gas between the particles and tube wall.
The resistance for heat transfer is a function of distance between particles and tube wall and the distance itself depends on the particle diameter. And so, a decreasing particle diameter brings about an increase in heat transfer coefficient.
In calculations involving heat and mass transfer the equivalent particle diameter (also called the "Sauter" mean diameter) is used:

$$d_{eq} = \frac{1}{\sum \frac{f_i}{d_i}}$$

in which f_i is the fraction of particles with diameter d_i.

In figure 1.5.1 the heat transfer coefficient is shown as a function of the equivalent bed-particle diameter, as measured in the 4 MW_{th} test facility (see section 1.2). Also indicated in the figure are calculated values. The calculations are based on theoretical models developed by Martin [12] concerning the particle convective and radiative heat transfer, and Vortmeyer [13] with respect to the gas radiative heat transfer, of which theories an overview is given by Görmar and Renz [14].
The decrease in heat transfer due to increasing particle diameter is considerable. For example, a drop in particle size from 0.9 to 0.6 mm means a 20 % increase in heat transfer. This is also shown in fig. 1.5.2, illustrating experiments with different sulfur sorbents causing a variation in particle size, and related steam production of the in-bed tube bundle.

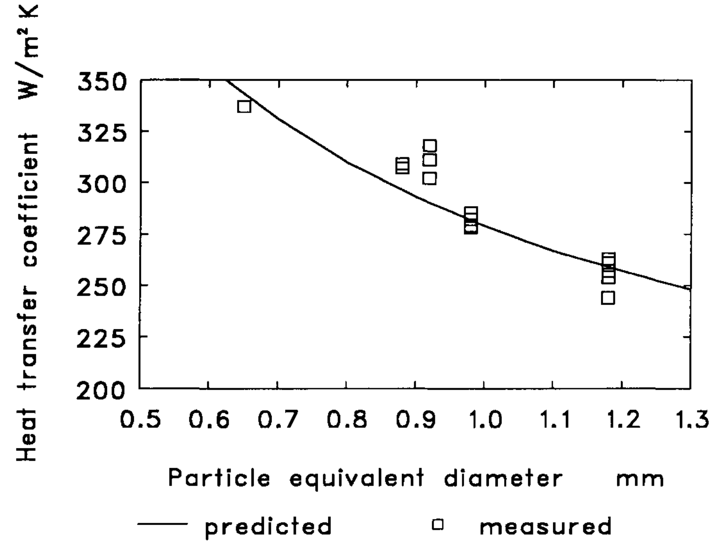


Figure 1.5.1. Heat transfer coefficient vs. bed particle equivalent diameter. T_{bed} = 850 °C, bed height 1.05 m, fluidizing velocity 2.2 m/s.

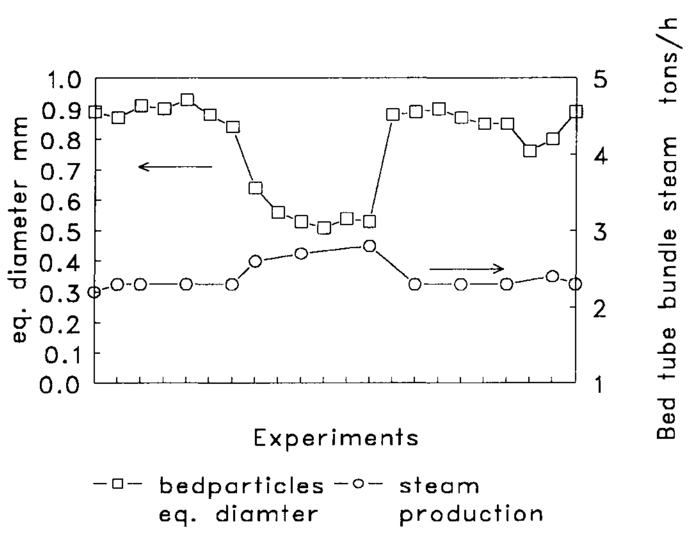


Figure 1.5.2. Steam production of in-bed tube bundle vs. bed particle diameter.

Results from the TVA Demonstration AFBC [16] showed a drop in heat transfer from 350 to 300 W/m^{2}K with a particle diameter increase from 0.5 to 1.5 mm.

1.5.2 Bed temperature

The bed temperature (i.e. particle and flue gas temperature) may considerably change heat transfer through the physical properties of flue gas and solids, i.e. density of gases, specific heat, viscosity and thermal conductivity.
The radiative heat transfer coefficient increases with the third power of temperature difference between bed and tube wall. The particle convective heat transfer coefficient also increases with increasing temperature, mainly because of the increase in specific heat conductivity of the flue gas.

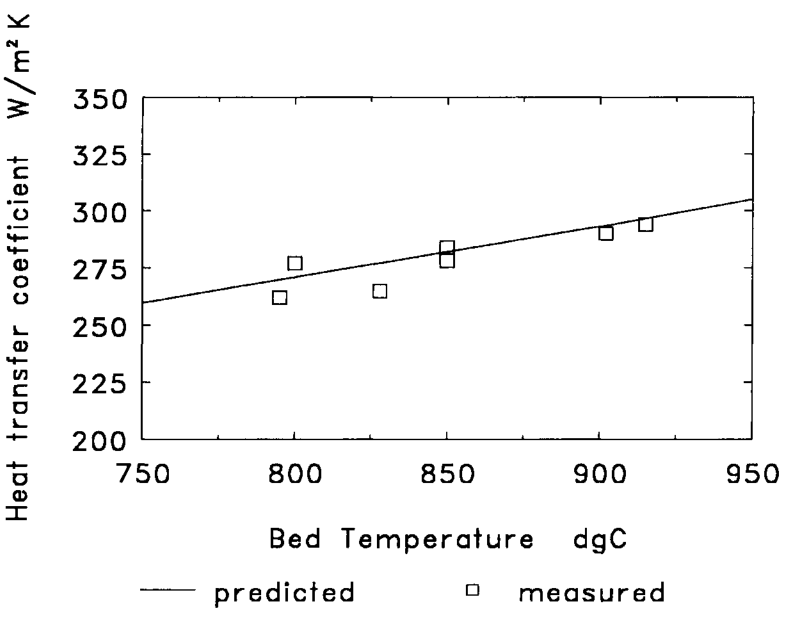


Figure 1.5.3. Heat transfer coefficient vs. bed temperature. Fluidizing velocity 2.4 m/s. Bed height 1.1 m; particles 0.92 mm.

1.5.3 Fluidizing velocity.

A higher bed voidage due to increased fluidizing velocity means a lower concentration of bed particles around the tubes, and so a shorter mean contact time between particles and tube wall. This results in a lower particle convective heat transfer coefficient.
Increasing bed voidage also results in lower particle densities in the bed section, causing

a decrease in the radiative heat transfer coefficient due to less radiating particles being present around the tubes.

Summarizing, increasing the fluidizing velocity leads to an increase in bed voidage resulting in a decrease of heat transfer.

On the other hand, the gas convective heat transfer increases due to an increased gas velocity around the tubes.

The effect of lower particle convective heat transfer however is predominant and this results in a lower total heat transfer coefficient with increasing superficial gas velocity. With normal operating conditions of the AFBC boiler, i.e. 850 °C bed temperature, 1.05 m bed height, 2 m/s fluidizing velocity and 0.9 mm equivalent particle diameter, the heat transfer coefficient amounts to about 300 W/m^{2}K.

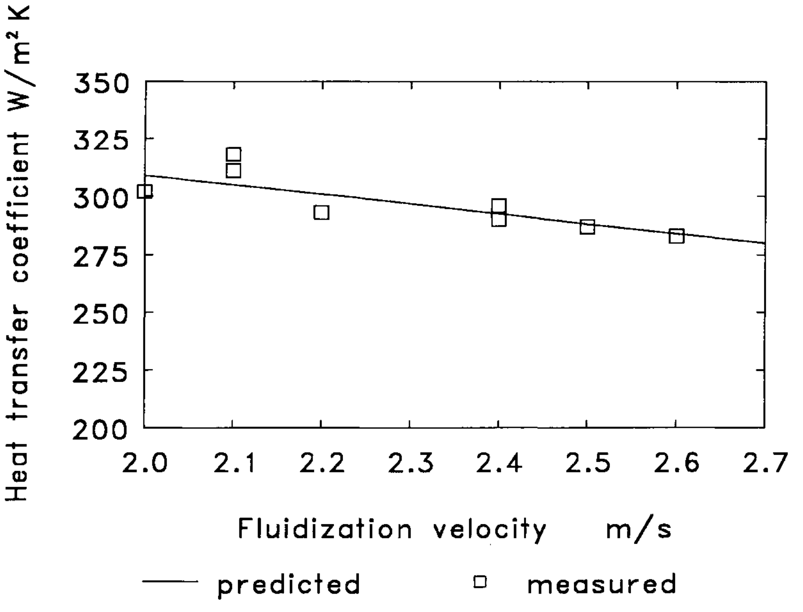


Figure 1.5.4. Heat transfer coefficient vs. fluidizing velocity. Bed temperature 850 °C; bed height 1.2 m; bed particle eq. diameter 0.92 mm

1.6 CONCLUDING REMARKS

The experimental research presented gives quantitative information about the difference in behaveour of various coal types and about the influence of process parameters on each coal type.

Simulation of the combustion efficiency by a simple model reveals insight in the influence of various combustion conditions on the combustion process and the sensitiveness of some specific process parameters related to the design and construction of fluidized bed combustors/boilers.

1.7 REFERENCES

1 van der Honing G.,
"Volatile and char combustion in large scale fluidized bed coal combustors",
PhD-Thesis,Twente University of Technology, Enschede, The Netherlands, (1990).

2 Jüntgen H.,
"Coal characterization in relation to coal combustion",
Erdöl und Kohle-Erdgas-Petrochemie, Bd.40, pp. 153-165, (1987).

3 van Engelen A.B., van der Honing G.,
"Char characterization based on combustion experiments and mathematical modelling ",
11 th International Conference on Fluidized Bed Combustion.
Montreal, vol. 1, 13-18, (1991).

4 Brem G.,
"Mathematical Modelling of Coal Conversion Processes",
PhD-Thesis,Twente University of Technology, Enschede, The Netherlands (1990).

5 van Engelen A.B., Valk M.,
"Coal characterization applied to overall mathematical modelling of fluidized bed combustion",
Report March 1991 (in Dutch), Twente University , Enschede, The Netherlands (1990).

6 Arena U., D'Amore M., Massimilla L.,
AIChE J, 29,1 pp. 40-49, (1983).

7 Daw C.S., Rowley D.R., Perna M.A., Stallings, J.W., Divilio, R.J.,
"Characterazation of fuels for atmospheric fluidized bed combustion",
11 th International Conference on fluidized bed combustion.
Montreal, vol. 1, 157-165, (1991).

8 Rappoldt L.M., van Gasselt M.L.G.,
"Improved fluid-bed combustor efficiencies through freeboard internals",
8 th International Conference on Fluidized Bed Combustion.
Houston, vol. 3, 647-654, (1985).

9 Wormgoor J.W., Bijvoet U.H.C., Gerrits B.J., van Gasselt M.L.G.,
"Enhanced environmental and economic performance of atmospheric fluidised bed boilers",
Proceedings of the 11 th International Conference on Fluidized Bed Combustion, Montreal, vol. 2, 665-676, (1991).

10 Bijvoet U.H.C., Wormgoor J.W., Tossaint H.H.J.,
"The characterization of coal and staged combustion in the TNO 4 MW_{th} AFFB research facility", Proceedings of the 10 th International Conference on Fluidized Bed Combustion, San Francisco, vol. 2, pp. 667-673, (1989).

11 Wein W.
"Stand und zukünftige Entwicklung der Wirbelschichtsysteme",
VGB Kraftwerkstechnik 68, Heft 12, (1988).

12 Martin H.,
"Heat transfer between gas fluidized beds of solid particles and surfaces of immersed heat exchanger elements, Part I + II", Chem. Eng.Proc. 18, 157-223, (1984).

13 Vortmeyer D., Börner C.J.,
"Die Strahlungsdurchlaßzahl in Schüttungen",
Chem. Ing. Tech., 38, 1077-1079, (1966).

14 Görmar H., Renz U.,
"Experimental and theoretical study on heat transfer in pressurized fluidized beds",
11 th International Conference on Fluidized Bed Combustion.
Montreal, vol. 1, 19-26, (1991).

15 Merrick, Highly,
"Particle size reduction and elutriation in a fluidized bed process",
AIChE Symposium Series No. 137, vol. 70, p. 366-378, (1974).

16 Carson R., Wheeldon J., Castleman J., Hansen P.,
"TVA 160 MW_e AFBC Demonstration plant process performance"
11 th International Conference on Fluidized Bed Combustion.
Montreal, vol. 1, p. 391-401, (1991).

17 Rappoldt L.M.,
"Summaries of TNO reports and publications on coal technology 1985-1990"
intended for NOVEM, TNO, Apeldoorn, The Netherlamds, (1991).

18 Valk M., Radovanović M., Alblas B.P.,
"Effect of fly ash recycle ratio on combustion efficiency for various coal types"
Proceedings of the Eighth International Conference on Fluidized-Bed Combustion, Houston, (1985), volume III, pp. 1408-1417.

19 Valk M., Bramer E.A., Tossaint H.H.J.,
"Effect of staged combustion of coal on emission levels of NOX and SO2 in a fluidized bed"
Proceedings of the 9th International Conference on fluidized bed combustion, Boston, (1987), volume 2, pp. 784-792.

20 Valk M. Bramer E.A., Tossaint H.H.J.,
"Optimal staged combustion conditions in a fluidised bed for simultanious low NOX and SO2 emission levels"
Proceedings of the 10th international conference on fluidized bed combustion, San Francisco, (1989), volume 2, pp. 995-1001.