THE EVALUATION OF THE FLUIDISED BED COMBUSTION PERFORMANCE OF SOUTH AFRICAN COALS IN THE PRESENCE OF SORBENTS

By

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ABSTRACT

The Fluidised Bed Combustion (FBC) technology has been widely used internationally for power generation. This technology has good fuel flexibility and reduced SO_2 emissions with dry sorbent (Limestone or Dolomite) addition. South Africa has large reserves of coals that are difficult to combust in conventional pulverised fuel fired boilers. These reserves could be potential feedstocks for new build FBC boilers. The chemical composition of these coals is site specific and could have an impact on the combustion performance of the fuel. This necessitates the need for FBC coal tests in the presence of a sorbent.

The objectives of this study were to investigate the changes in the production NO_x, SO_x, and the combustion efficiency of the three test coals under conditions of fluidised bed combustion, with the same sorbent. Tests with no sorbent were performed to evaluate the coals inherent calcium capabilities of capturing sulphur. Tests with varying ratios of sorbent were performed to evaluate the sorbent's capabilities for further levels of in-bed desulphurisation. The experimental equipment used in this investigation was the Eskom Fluidised Bed Test Facility (FBTF). This facility is a bubbling fluidised bed combustor/gasifier. The investigated bed temperature range was between 800 to 900°C, in intervals of 20°C. The operating pressure was 50kPa (gauge).

The three coals were compared at Ca/S molar ratio of 1. Carbon in ash has shown to decrease with an increase in bed temperature for Coal A, B and C. The best performing coal in terms of least quantity of remaining carbon in ash was Coal A. The NO emissions increased for an increase in bed temperature for Coal A, B and C. The greatest NO emissions were recorded during Coal B tests. The N_2O emissions decreased with an increase in bed temperature for Coal A and B tests. Higher N_2O emissions were observed for Coal B than Coal A tests. In terms of SO_2 retention Coal C performed the best. The optimal operating bed temperature for SO_2 retention observed for the three coals was in the region of 800-860°C.

DECLARATION

I Lesigen Moodley, student number 98195400, declare both the testing conducted on the FBTF and this dissertation is of my own original work. It is being submitted for the coursework Masters of Science degree to the University of Kwa-Zulu Natal, and has not been submitted for any degree at any other university.

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LIST OF SYMBOLS

A	Cross sectional area, m^2
W	The mass % of a chemical element or compound in coal
m	Mass, kg
$\dot{m}$	Mass flow rate, kg/hour
$\dot{v}$	Volumetric flow rate, m^3/hour
X	Concentration of flue gas species
T	Temperature, °C
P	Pressure, kPa, (gauge)
U	Retention, %

Subscripts

as received	Chemical analyses reported on an as received basis
air dried	Chemical analyses reported on an air dried basis
moist	Moisture
@6% O_2	The concentration has been normalised to a reactor exit of 6% O_2
max	Maximum produced
flue	Flue gas
actual flue	Actual flue gas corrected for pressure and temperature
sam	Flue gas sample point
reactor	Reactor vessel
emitted	Exiting the reactor

DEFINITIONS

Eskom	Is a South African electricity public utility company established in 1922.
Discard coal	Waste, non saleable coal
Sorbent	Limestone (sedimentary rock composed largely of mineral calcite) or dolomite (a rock consisting of largely magnesium carbonate)
Slag	Scrum formed by oxidation at the surface of molten metals
NOx	Oxides of nitrogen ($NO + NO_2$)
SO_2	Sulphur dioxide
N_2O	Nitrous oxide, Di-nitrogen monoxide

Chapter One

INTRODUCTION

The electricity demand in South Africa is increasing which can be attributed to increased consumption in industry, transport, agriculture, commerce and the residential sector (Digest of South African Energy Statistics, 2005). As a result of this increased consumption pattern, the electricity utility for South Africa Eskom is in the process of expanding its power generation capacity (Eskom News 2005). This utility provides more than 90% of South Africa's electricity requirements. Pulverised coal fired power plants currently make up 92.1% of this capacity (Eskom Annual Report 2005). The large reserve of cheap coal in South Africa still promotes the use of coal for future expansion of power generation (Eskom News 2005).

Annually 60 million tons of coal is discarded in South Africa from mining processes which have been estimated to be more than one billion tons (DME 2006). Fluidised Bed Combustion (FBC) is a clean coal technology which appears to be an attractive technology for future new build power plants. This technology has been proven internationally and has the ability to combust poor quality coals. The suitability of combustion of some of these coals in a fluidised bed is examined in this dissertation. These fluidised bed coal combustion tests were conducted by Eskom research, utilising its test facility (Figure 1-1).

The design of this test facility originates from India (BHEL), which combusts coals of similar properties to that utilised in the South African scenario. The purpose of the Fluidised Bed Test Facility (FBTF) is to evaluate the combustion or gasification properties of South African coals. The flexibility of the facility allows for in-bed desulphurisation, utilising locally available dolomite and limestone (sorbent). Additional capabilities of the FBTF are online feeding of coal, sorbent and online extraction of bottom and fly ash which allows the facility to be operated 24 hours a day, seven days a week if required to do so.

Figure 1-1 The Fluidised Bed Test Facility (FBTF),
(Area enclosed by the dotted rectangle)

Based on international experience FBC is an attractive technology for the combustion of poor quality coals. The inherent advantages of the FBC stems from the fact that it operates at a much lower temperature (800-900°C) than conventional pulverised fuel fired boilers. This means that the levels of thermal NO_x produced are of much lower concentrations. In-bed desulphurisation is possible through feeding of sorbent (limestone or dolomite) into the combustion zone. A heat exchanger that is submerged in the bed poses a very high heat transfer co-efficient, due to the interaction with the moving bed particles.

The fact that the current prices of oil and gas are rising (Wikipedia 2006), and the increased world demand for energy, interest in the return to use of coal as a source of energy has grown. This problem will place increased pressure on the world's coal reserves. Conventional pulverised fuel fired boilers require a certain quality of coal therefore problematic coals will be discarded. The FBC technology may prove to be an attractive power generation technology if discarded coals can be purchased at a much lower price.

The objectives of this study were to investigate the changes in the production of NO_x, SO_x, and the combustion efficiency of the three test coals under conditions of fluidised bed combustion, with the same sorbent. The results of these tests are intended to provide Eskom with strategic information on the possible coal and sorbent feedstock's for new build fluidised bed boilers for power generation.

Chapter Two

LITERATURE REVIEW

2.1 INTRODUCTION TO THE LITERATURE REVIEW

The expansion of South Africa's power generation capacity and the need for fluidised bed combustion tests were introduced in chapter one.

The objective of this chapter is to examine the following,

- The advantages and disadvantages of the FBC technology.
- The behaviour of a fluidised bed and the effects of process parameters.
- Coal combustion in a fluidised bed.
- Classification of fluidised bed boilers.
- A survey of fluidised bed boilers installed internationally.
- Other research on the tested sorbent

2.2 ADVANTAGES AND DISADVANTAGES OF THE FBC TECHNOLOGY

The power generation capacity by Eskom is largely supplied by pulverised coal fired power plants as discussed in chapter one. This means that there is a large knowledge base on this technology and the combustion behaviour of South African coals fired in these power plants. For the fluidised bed coal combustion technology to compete successfully, it must show benefits over current coal combustion technologies. This section intends to examine both the advantages and disadvantages of fluidised bed combustion against current coal combustion technology.

The advantages of FBC over conventional combustion systems, DTI (1999),

- The bed can be operated at temperatures typically between 800°C and 900°C above which ash may melt and form slag.
- The scrubbing action of the moving particles on the immersed water tubes increases the rate of heat transfer.
- The bed has a substantial thermal capacity which allows a variety of fuels to be burnt including low quality fuels with high levels of unwanted mineral matter or moisture content as well as mixed fuels.
- NO_x formation is reduced because the combustion takes place at relatively low temperatures and the system offers opportunities for air staging.
- Up to 90% of the SO_2 released during combustion can be captured in the bed by adding sorbent (limestone or dolomite).

The disadvantages of FBC technology, DTI (1999),

- Commercially proven only at relatively small scale compared to pulverised coal combustion.
- Relatively large amounts of solid residues generated (with the addition of sorbent), some of which requires special measures for disposal.
- Higher carbon in ash levels than those from pulverised coal combustion.
- Increased N_2O formation due to the lower combustion temperatures.

The extent of the latter two disadvantages (carbon in ash and N_2O emissions) will be examined for the test coals in this dissertation.

2.3 DESCRIPTION OF THE FLUIDISED BED PROCESS

The previous section expanded on advantages of FBC over current coal combustion technology, this section examines the differences in the fluidised bed processes and the effects of operational conditions. The current coal technology involves pulverised fuel being introduced into the boiler through a burner. The combustion of coal produces a flame jet. The fluidised bed combustion process involves having a mixture of particles suspended by an upward movement of air. These particles compose of ash, fuel and inert material which resemble a bed. A bed of particles will offer resistance when fluid flows through it. When the operating bed velocity increases, the drag force exerted on these particles increases. Howard (1983) explains that an upward flow through an unrestrained bed causes the particles to re-arrange themselves so that they offer less resistance to the fluid flow through it. This then causes the bed to expand. With a further increase in the upward fluid velocity, expansion continues until a stage is reached where the drag force exerted on the particles will be sufficient to support the weight of the particles in the bed. This paragraph summarises that the differences in the combustion zone of the FBC and current coal technology is that the FBC process involves having a bed of particles. The following subsections examine the fluidised bed in greater detail.

2.3.1. THE EFFECT OF FLUID VELOCITY ON PRESSURE DROP ACROSS THE BED

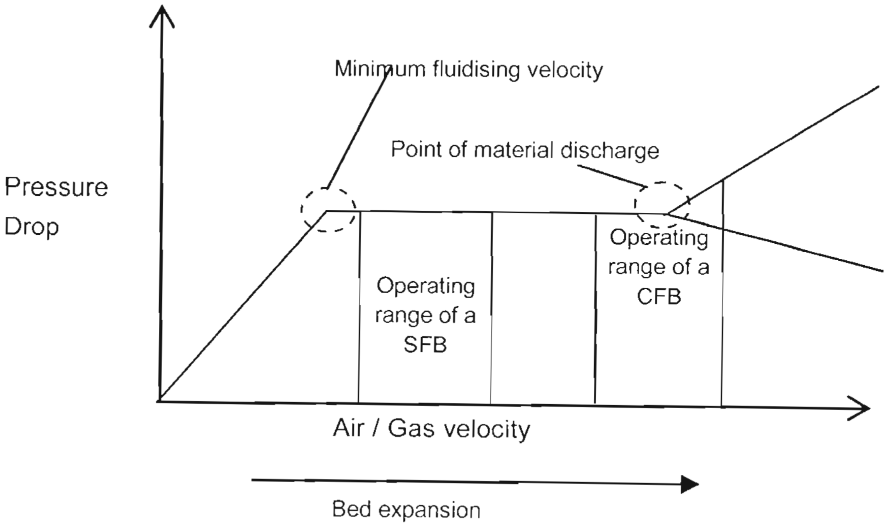

Figure 2 - 1 Fluid flow velocity versus bed pressure drop

The effect of bed fluid velocity on bed pressure drop is examined in this section and is depicted in Figure 2-1 (redrawn from Kakaras E et al., 2001). As the fluid velocity increases the pressure drop across the bed increases, until the minimum fluidising velocity is reached (U_{mf}). Beyond this point, the bed pressure drop remains constant. At this stage the pressure drop across the bed multiplied by the cross sectional area of the bed will be approximately equal to the weight of the bed per unit area. (Howard, 1983):

$$\Delta P_b A_b = \frac{m}{\rho_p \left(\rho_p - \rho_f \right) g} \qquad 2-1$$

The actual value of fluidising velocity depends on a compromise between the capital cost, the bed pressure drop and the efficiencies of combustion and sulphur retention, DTI (1999). This section details the operating velocity window for fluidised beds. To ensure consistency between tests, all tests must be conducted at the same velocity.

2.3.2. THE BEHAVIOUR OF BUBBLES IN A GAS FLUIDISED BED

Having examined the effect of fluidising velocity in the previous section, the behaviour of bubbles in the bed will be reviewed in this section. Bubble formation according to Laux et al. (2003), occurs as gas enters the distributor plate. These bubbles are formed at the orifices with small sizes. They rise towards the bed surface with growing size or converge into larger bubbles. At the bed surface some bubbles break up, leading to fluctuation of the bed surface. The shape of a bubble is either nearly spherical or a spherical cap with a rear indentation, Zhangfa (2003).

Howard (1983) concluded that a bubble within a chain or cloud of bubbles rises faster than one in isolation, except for those in beds of larger mean particle diameter of 1mm, when the rise rate tends to be slower. As bubbles rises through the bed they grow by collection of gas from the surrounding continuous phase and by coalescence. He explains that during coalescence a small bubble comes within influence of a faster rising one which overtakes it and draws it into its wake. The bubble behaviour is generally affected by the distributor design. Particles are also carried within the wake of a bubble. With deep beds of high aspect ratio and higher operational velocities, bubbles may grow until they occupy the whole cross sectional area of the bed, these then carry a slug of particles ahead of them until instability occurs and the solids collapse back into the bed (slugging).

2.3.3. THE EFFECT OF PARTICLE SIZE ON FLUIDISATION

This section reviews the effect of particle size (coal and sorbent) on the behaviour of a fluidised bed. This behaviour of different materials has been classified by Geldart (1973). Research by Geldart (1980) showed that the actual proportion of fines (classed as material <45μm) is of particular importance. Geldart also classified the significance of this effect in his material grouping categorisation (1973). Materials categorised in his group A, have densities less than 1400kg/m^3 and fall within the size range of 20 to 100μm. He also concluded that powders exhibit a pronounced degree of stable bed expansion when the minimum fluidisation velocity is first exceeded, and it may be possible to sustain uniform or particulate fluidisation until the minimum fluidisation velocity has been exceeded by a factor of two to three times.

Materials categorised by Gerldart's group B materials (1973), have a mean size range within 40 to 500μm and a density in the range of 1400 to 4000kg/m^3. These exhibit much less stable bed expansion, free bubbling (aggregative fluidisation) occurring at or a little above the minimum fluidisation velocity. Howard (1983) describes aggregative fluidisation as, when the gas velocity is increased further, a point will be reached when the bed will collapse back to a less expanded state approximating more closely to the degree of expansion under the minimum fluidising conditions and most of the excess gas will flow through the bed as the bubble phase.

Materials categorised by Gerldart's group C materials (1973), have smaller mean size (<30μm) and of lower density so that inter-particle forces have an effect greater than that of gravity. He describes such materials as very difficult to fluidise. Within these materials, channelling is prone to occur. That is the flow of fluid that opens up low resistance channels through the bed. Once the path has been created, it tends to enlarge with further increase in gas velocity so that the gas is not properly distributed into the bed which never becomes truly fluidised.

Materials categorised by Gerldart's group D materials (1973), are usually of mean diameter greater than 600μm or dense particles. For materials from this group, the interstitial flow rate of the gas through the continuous phase is necessary to fluidise the bed which is greater than the rising velocity of the bubbles, Howard (1983).

It is important to note that because of the difference in bubbling behaviour between different classes of materials there can be very different changes when scaling up from small experimental beds to full scale operational beds. This is because bubbles are constrained in size by the scale of the equipment in beds of small diameter but have the opportunity to grow very much larger in larger scale equipment.

The effects of solids in a bed according to Mogan et al, (2000) produced a much greater fluctuation in air velocity than air without solids. Without solids the flow velocity fluctuation was 0.5m/s and increased to 3m/s at a solid concentration of 0.924%.

This section summarises that particle size has an effect on bubbles in the bed, and the small diameters of experimental beds, constrains the size of the bubbles. The solid concentration in the bed does influence air velocity fluctuation. To ensure consistency between tests, the same bed height must be used for all tests.

2.3.4. THE EFFECT OF AIR DISTRIBUTION

The previous section explains that solid concentration in the bed influences air velocity fluctuations. Mogan, et al. (2000) also concludes that the air flow in cross section across the fluidised bed is not uniform. It is generally seen that the air velocity has a maximum at the bed centre and decreases gradually with an increasing distance from the centre. Air distribution will be a greater problem in fluidised beds with large cross sectional diameters.

2.4 COMBUSTION IN A FLUIDISED BED

The fluidised bed process was discussed in section 2.3. The process the coal undergoes in the FBC is explained in this section. The processes of combustion in a fluidised bed as indicated by Zhangfa (2003) are as follows:

- Drying (with or without shrinkage of the particles)
- Devolatilisation (with or without swelling and fragmentation)
- Combustion of the volatile matter
- Combustion of the residual char particles

Each process is examined in detail in the following subsections.

2.4.1 DRYING

Drying of the coal particles occurs once it enter the bed. During this process 2-20% of the fuel moisture is removed, according to Anthony et al, (1995). Their research concluded that it appears that evaporation of surface moisture does not affect the coal combustion process directly.

Low rank coals contain more than 40% inherent moisture and its evaporation may slow down the devolatilisation and ignition processes. The loss of water can result in significant morphological changes in low rank coals. Australian brown coal can experience a volume reduction of up to 40%, (Agarwal et al (1989)).

This section indicates drying of coal entering the bed can slow down devolatilisation and ignition of some coals. The coals once prepared must be stored in sealed drums to prevent changes in the surface moisture of coal.

2.4.2 DEVOLATILISATION

Devolatilisation (release of volatile matter) occurs by further heating of the coal particle after drying. This normally starts at 300°C, (Anthony (1995)). Devolatilisation is a pyrolytic process whereby the large and heavy molecules of organic substances in the coal break up according to Ross et al, (2000). The weaker carboxyl, hydroxyl, and aliphatic bonds break up at low temperatures, while at temperatures greater than 650°C the stronger heterocyclic components decompose.

The process produces various gaseous species such as low-molecular weight hydrocarbons, hydrogen, carbon monoxide, carbon dioxide, methane, and tars, (Kilpinen et al, (2002)). It finally leaves a solid residue (containing typically 98% carbon) which is called char. The amount of char varies widely for different fuels, and is typically about 45 wt% for bituminous coal, 20 wt% for peat and sewage sludge, and approximately 10 wt% for wood based fuels and various waste driven fuels.

During the process of devolatilisation the coal particles are surrounded by a characteristic luminous, plume like flame described by Urkan et al, (1994). The flame is formed almost immediately or a few seconds after introduction into a fluidised bed combustor and for most of the time remains near the surface of the bed throughout volatile evolution. The coal particles initially appear as dark spots inside the flame and eventually turn yellow towards extinction of this flame. The coal structure also changes and this can be a sequence of events of coal particle softening, re-solidification and contraction, (Strezov et al, (2002)).

Structural change is due to the hydrocarbon transformation from solid to gaseous phase explained by Daki et al, (2000). This process causes the volume of the hydrocarbons to increase by approximately an order of three in magnitude. This leads to an increase in pressure inside the coal particle and changes in coal structure such as porosity and appearance of the bubbles in the coal particle.

Structural change can vary from one coal to another according to Anthony et al, (1995). Larger particles of a bituminous coal which have a free swelling index of 5 to 5.5, caused the coal particle to swell, softened and developed a shell. Bituminous coals with a swelling index of 1 to 2 devolatilised with little swelling.

Zhangfa (2003) concluded that the rate of coal devolatilisation may be controlled by the following three mechanisms; chemical reaction kinetics, heat transfer to and within the particle and mass transfer of volatile products within the particle. Suggested models by Agarwal et al, (1989) described the rate of devolatilisation as follows:

- For very small particles less than 0.1mm, the devolatilisation time is independent of particle size, indicating chemical reaction kinetics is the rate controlling factor.
- As particles increases in size, both chemical reaction kinetics and heat transfer control the rate of devolatilisation.
- For particles larger than 1mm the rate is heat transfer controlled with devolatilisation time being proportional to the square of the particle diameter.

2.4.3 COMBUSTION OF VOLATILES

The volatiles released from devolatilisation of the coal particles burn in the presence of oxygen which comes mainly from the fluidising air but also from coal, (Zhangfa (2003)). Combustion of volatile species is a rapid process where CH_4 and H_2 oxidations are very fast, while CO oxidations is comparatively slow, as explained be Desroches et al, (2002). The combustion of volatiles can contribute to as much as 50% of the total heat released, (Zhangfa (2003)).

Combustion of volatiles are both homogenous (O_2, H_2, CH_4, C_2H_6, C_3H_8) and heterogeneous, (Liu et al (2002)). The combustion of HCN and NH_3 can be either homogenous or heterogeneous, (Desroches et al, (2002)). Some of the chemical reactions that occur in combustion of volatiles, detailed by Zhangfa (2003) are,

Carbon monoxide:

$$CO + \tfrac{1}{2}O_2 \rightarrow CO_2 \qquad 2 - 2$$

Hydrogen:

$$2H_2 + O_2 \rightarrow 2H_2O \qquad 2 - 3$$

Methane:

$$CH_4 + \tfrac{3}{2}O_2 \rightarrow CO + 2H_2O \qquad 2 - 4$$

C_2H_6:

$$C_2H_6 + \tfrac{5}{2}O_2 \rightarrow 2CO + 3H_2O \qquad 2 - 5$$

Propane:

$$C_3H_8 \rightarrow \tfrac{3}{2}C_2H_4 + H_2 \qquad 2 - 6$$

$$C_2H_4 + O_2 \rightarrow 2CO + 2H_2 \qquad 2 - 7$$

$$CO + \tfrac{1}{2}O_2 \rightarrow CO_2 \qquad 2 - 8$$

$$H_2 + \tfrac{1}{2}O_2 \rightarrow H_2O \qquad 2 - 9$$

HCN:

$$HCN + \frac{5}{4}O_2 \rightarrow NO + CO + \frac{1}{2}H_2O \qquad 2 - 10$$

$$HCN + \frac{3}{2}O_2 + NO \rightarrow N_2O + CO + \frac{1}{2}H_2O \qquad 2 - 11$$

$$HCN + \frac{5}{4}O_2 \rightarrow \frac{1}{2}N_2 + CO_2 + \frac{1}{2}H_2O \qquad 2 - 12$$

NH_3:

$$NH_3 + \frac{5}{4}O_2 \rightarrow NO + \frac{3}{2}H_2O \qquad 2 - 13$$

$$NH_3 + \frac{3}{4}O_2 \rightarrow \frac{1}{2}N_2 + \frac{3}{2}H_2O \qquad 2 - 14$$

Volatiles in bubbling fluidised beds may burn either in a solid free bubble phase or in the spaces between the solid particles called the emulsion phase explained by Bulewicz, (2002). Srinivansan et al, (1998) concluded that the phase in which combustion occurs depends on the following bed temperature,

- At low bed temperature, less than 700°C, the volatiles may not burn within the bed.
- At temperatures between 830 to 900°C, the volatiles are likely to burn only in the bubbles.
- At high bed temperature, above 900°C, combustion may occur within the bubbles and in the particulate phase of the bed.

The effect of particle size on the phase in which combustion occurs, for larger than 600μm and bed temperatures higher than 800°C, occurs in the emulsion phase (Stubington et al (1991)).

2.4.4 COMBUSTION OF THE CHAR PARTICLES

Char burning is the final stage of the combustion process and is much slower than the combustion of volatiles matter, (Zhangfa (2003)). Char is normally present in the bed less than 1% of the bed mass if crushed coal is used and less than 5% for large size grading, according to Metrick (1984). Metrick also concluded that the char particle is normally hotter than its surroundings and the particle surface temperature is up to 100°C higher than the bed temperature.

Zhangfa (2003) details the chemical reactions of carbon particles are,

$$CO + \tfrac{1}{2}O_2 \rightarrow CO \qquad 2 - 15$$

$$CO + \tfrac{1}{2}O_2 \rightarrow CO_2 \qquad 2 - 16$$

$$C + CO_2 \rightarrow 2CO \qquad 2 - 17$$

Basu (1999) concluded there are three mechanisms for the combustion of the char particle. Mechanism one assumes that oxygen diffuses to the char particle surface and oxidises it to CO. The CO then oxidises to form CO_2 in a gas phase reaction. As CO oxidises very close to the carbon surface, CO_2 may be considered as a primary combustion product. In mechanism two, oxygen diffuses to the char surface and produces both CO and CO_2. CO and CO_2 then diffuse away from the carbon surface. CO further reacts in a gas-phase with oxygen arriving from the bulk gas and forms CO_2. Mechanism three assumes that oxygen cannot reach the carbon surface and it reacts with CO in the gas phase reaction away from the carbon surface. Part of the CO_2 formed diffuses back to the carbon surface where it is reduced to CO.

In a combustion model explained by Anthony et al, (1989) the following processes occur,

- Char particle size decreases continuously while density remains constant (shrinking particle model).
- Char particle density decreases continuously while its size remains constant (progressive conversion model).
- Char particle density within a shrinking char core remains constant, with the appearance of an ash layer (shrinking core model).

2.4.5 COMBUSTION EFFICIENCY

Combustion efficiency in a fluidised bed defined by Zhangfa (2003) is the ratio of heat released to the energy supplied by the fuel. Generally circulating fluidised bed combustors have higher combustion efficiencies than bubbling fluidised bed combustors, due the high solid recycle rate. Fuel type plays a significant role in combustion efficiency. Volatile content allows for greater reactivity and greater char combustion rate, detailed by Armesto et al, (2001). A higher bed temperature produces higher reactions rates and higher combustion efficiencies, according to Adanez et al, (2001). Their results showed an increase in combustion efficiency from 94 to 96% for a temperature range of 800 to 900°C in a circulating fluidised bed.

An increase in gas velocity, decreases the combustion efficiency, which is explained by Zhangfa (2003). Armesto et al, (2001) showed a decrease in combustion efficiency from 98.4 to 97% by increasing the gas velocity from 1 to 1.2 m/s. Excess air increase means an increase in mean oxygen concentrations in the bed, therefore combustion efficiency increases when velocity is kept constant.

2.4.6 POLLUTANT FORMATION IN A FLUIDISED BED

The pollutants formed by a fluidised bed are; nitrogen oxides, nitrous oxide, sulphur oxides, particulates, carbon monoxide, and trace elements.

2.4.6.1 Nitrogen oxide and nitrous oxide

This section details the formation of nitrogen oxide and nitrous oxide from the coal char and volatile nitrogen components. Figure 2-2 shows gas specie formations during this process.

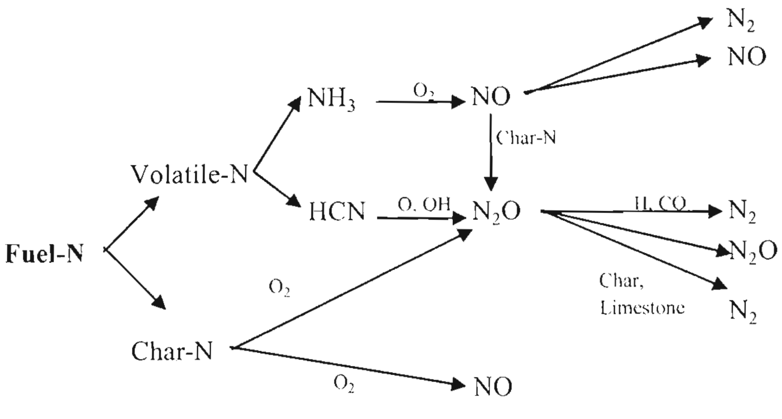

Figure 2 - 2 NO and N_2O formation and reduction (redrawn from Anthony et al, 1995)

During the process of fluidised bed combustion nitrogen in both the combustion air and fuel may converted to nitrogen oxides (NO and NO_2, together referred to as NO_x). The NO_x produced by combustion air is referred to as thermal NO_x. In a FBC at low temperatures the nitrogen in the combustion air does not oxidise significantly, thus thermal NOx is low, and therefore all the NO_x produced is from fuel nitrogen, as explained by Armesto, (2001). This implies NO_x emissions from FBC are much lower than conventional pulverised fuel fired combustors.

Although FBC produces lower NO_x emissions, it does produce significant quantities of nitrous oxide (N_2O) emissions. This is due to the lower combustion temperatures. The N_2O emissions range from 20 to 300ppm compared to 10ppm found in conventional pulverised fuel combustion, Armesto et al, (2001).

The formation of NO and N_2O from coal nitrogen is a complex process. This process involves devolatilisation of the coal, volatile species oxidation, char oxidation, and finally reduction to N_2, according to Zhangfa (2003). Anthony et al. (1995), explains that during the process of devolatilisation, organically bound nitrogen in the fuel is partitioned into volatile nitrogen (released as HCN and NH_3) and char nitrogen (bound in aromatic structures, see Figure 2-2). Desroches et al, (2002) concluded that volatile nitrogen compounds are oxidised through homogenous gas phase reactions or heterogeneous catalytic reactions over char and other bed solids such as ash. HCN forms N_2O.

Lui et al, (2002) explains that during char oxidation nitrogen oxidises mainly to NO, N_2O and small amounts of HCN and NH_3. The three main reactions concluded by them are,

$$Char - \mathrm{N} + \tfrac{1}{2}\mathrm{O_2} \rightarrow -\mathrm{CNO} \qquad 2\text{ - }18$$

$$-\mathrm{CNO} + \tfrac{1}{2}\mathrm{O_2} \rightarrow \mathrm{NO} + \mathrm{CO_2} \qquad 2\text{ - }19$$

$$-\mathrm{CNO} + \mathrm{NO} \rightarrow \mathrm{N_2O} + \mathrm{CO_2} \qquad 2\text{ - }20$$

The NO and N_2O formed are then partially reduced to N_2, as explained by Zhangfa (2003). NO can be reduced by NH_3, N_2O can be decomposed due to temperature, and also NO and N_2O can be reduced either directly by char or through reactions with CO catalysed char, (Zhangfa (2003)). The reduction reactions presented by Desroches et al, (2002) and Lan et al (2001), are:

NO reduction:

$$\mathrm{NO} + \mathrm{NH_3} + \tfrac{1}{4}\mathrm{O2} \rightarrow \mathrm{N_2} + \tfrac{3}{2}\mathrm{H_2O} \qquad 2\text{ - }21$$

$$\mathrm{NO} + \tfrac{2}{3}\mathrm{NH_3} \rightarrow \tfrac{5}{6}\mathrm{N_2} + \mathrm{H_2O} \qquad 2\text{ - }22$$

$$\mathrm{NO} + \mathrm{Char} \rightarrow \tfrac{1}{2}\mathrm{N_2} + \mathrm{CO} \qquad 2\text{ - }23$$

$$\mathrm{NO} + \mathrm{CO} \rightarrow \tfrac{1}{2}\mathrm{N_2} + \mathrm{CO_2} \qquad 2\text{ - }24$$

N_2O reduction:

$$N_2O \rightarrow N_2 + \frac{1}{2}O_2 \qquad 2 - 25$$

$$N_2O + \text{Char} \rightarrow N_2 + CO \qquad 2 - 26$$

$$N_2O + CO \rightarrow N_2 + CO_2 \qquad 2 - 27$$

The conversion of fuel nitrogen to NO_x and N_2O is affected by the fuel reactivity, explained by Anthony et al, (1995). They concluded that since wood is more reactive than coal, wood under conditions of combustion would experience a higher degree of fuel nitrogen conversion. With increasing content of fuel volatiles, the NO emissions usually increases but N_2O emissions decreases according to Basu (1999). The effect of temperature showed that as temperature was increased from 800 to 850°C, the conversion to NO_x increased from 30 to 50% while conversion to N_2O decreased from 11 to 8% according to Armesto et al, (2001). When the excess air is increased, it leads to an increase in both NO_x and N_2O, (Zhangfa (2003)). This is because increasing excess air increases the combustion rates, and enhances the oxidation of HCN and NH_3. Air staging substantially reduces both NO_x and N_2O, according to Abelha et al, (2003). This is achieved by injection of secondary air at different levels in the reactor. The addition of limestone influences both the formation and reduction of NO. It is a catalyst for NO formation from volatile NH_3 and the reduction of NO by CO, (Zhangfa (2003)). Anthony et al, (1995) indicated that in a circulating fluidised bed combustor, an increase in NO emissions where observed.

2.4.6.2 Sulphur oxides

Sulphur occurs mainly in coal as pyrite (Fe_2S) and as organic sulphur compounds. The sulphur content in coal ranges from 1 to 10%. The sulphur in coal oxidises to form mainly SO_2 (sulphur dioxide) and a small amount of SO_3 (sulphur trioxide) in the combustion process. By the process of the introduction of sorbent (limestone or dolomite) into the combustion zone, SO_2 can be captured within the bed as it is being produced. When limestone is first introduced, it calcines to calcium oxide (CaO) and dolomite to calcium oxide magnesium oxide (CaO.MgO). Podolski et al, (1995) and Anthony et al, (1995) describes the chemical reactions that take place as follows,

Limestone:

$$CaCO_3 \rightarrow CaO + CO_2 \qquad 2 - 28$$

$$2CaO + 2SO_2 + O_2 \rightarrow 2CaSO_4 \qquad 2 - 29$$

Dolomite:

$$CaCO_3.MgCO_3 \rightarrow CaO.MgO + 2CO_2 \qquad 2 - 30$$

$$2CaO.MgO + 2SO_2 + O_2 \rightarrow 2CaSO_4.MgO \qquad 2 - 31$$

In a fluidised bed under conditions of pressure, due to the increase in the partial pressure of CO_2, the calcination process is affected negatively. In this case the SO_2 reacts directly with $CaCO_3$ in the limestone or dolomite. Pololski et al. (1995) and Anthony et al. (1995), describes the chemical reactions that takes place for pressurised operation as follows,

Limestone:

$$2CaCO_3 + 2SO_2 + O_2 \rightarrow 2CaSO_4 + 2CO_2 \qquad 2 - 32$$

Dolomite:

$$2CaCO_3.MgCO_3 + 2SO_2 + O_2 \rightarrow 2CaSO_4.MgO + 4CO_2 \qquad 2 - 33$$

Typical sorbent utilisation researched by Anthony et al. (1995) in an FBC was 30-50%. The reactivity of the sorbent is influenced by the pores in the sorbent particles. The sulphation reaction may cover pores with $CaSO_4$ thus the core of CaO will not be available for further reaction with SO_2.

Research done by Zhangfa (2003) showed the maximum retention of SO_2 for the tested coals normally occurs at a bed temperature of 850°C. The effects of bed temperature on sulphur retention according to Anthony et al, (1995) are as follows:

- At higher temperatures enhances the reverse sulphation reaction, thus decreases sulphur retention.
- At higher temperatures enhances the sintering of sorbent particles, this reduces their porosity, thus reduces retention.
- The ratio of SO_2/SO_3, at higher temperatures reduces the availability of SO_3 for reaction with CaO.
- Oxygen depletion in the bed is due to increased volatile combustion at higher temperatures, inhibits the sulphation reaction above 900°C.

Podolski et al, (1995) showed that for a Ca/S mole ratio of 1 to 1.5 in a pressurised FBC, sulphur retention of 90% was observed, and for Ca/S mole ratio of 2 a sulphur retention of 95% is possible.

2.5 CLASSIFICATION OF FBC BOILERS

Section 2.4 examined the process of fluidised bed coal combustion and the formation of pollutants. This section examines classification of FBC boilers. Kakaras et al. (2001) concluded that FBC technologies can be distinguished based on fluidisation velocity and the combustion process pressure of the boiler. Based on the fluidisation velocity (see both Figure 2-1 and 2-3),

- The bed is said to be fixed if the operating velocity of the bed is below the minimum fluidisation velocity.
- If the operating velocity overcomes the minimum fluidisation velocity and goes up to a certain limit dependant on particle size (typically 5m/s), the bed is characterised as bubbling.
- If the operating velocity is increased, reaching 5 to 10m/s, there is a great elutriation of solid material in the range of 30 to 70 times the feed load. A cyclone is required as well as a system for re-circulation of flow (Figure 2-4 depicts a hot cyclone). The bed is then a Circulating Fluidised Bed (CFB).
- If the velocity is increased furthermore, the fluidisation disappears and there is pneumatic transport of the bed solids.

Based on the process pressure,

- Fluidised bed combustion may take place at atmospheric pressure (AFBC).
- Fluidised bed combustion may take place at pressures above atmospheric conditions, which is referred to as Pressurised Fluidised Bed Combustion (PFBC).

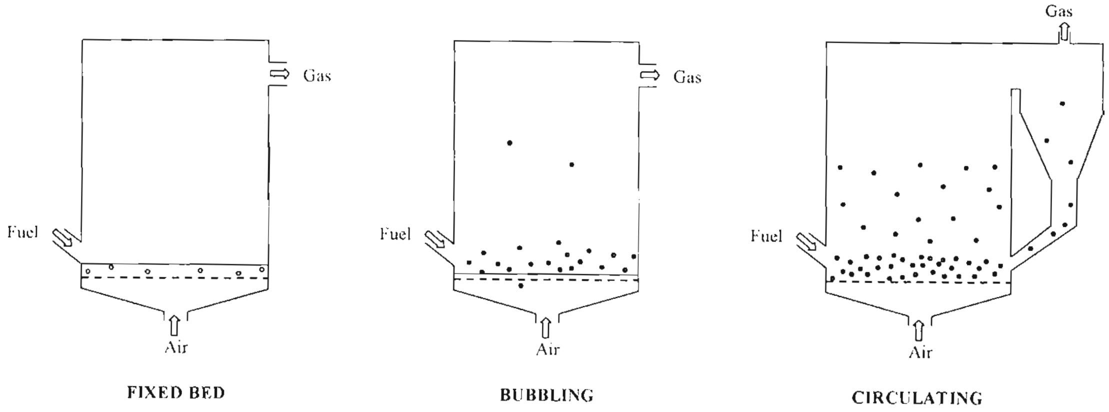

Figure 2 - 3 Fluidisation regimes, Kakaras et al, (2001)

2.5.1 CIRCULATING FLUIDISED BEDS

The circulating fluidised bed process was developed by Lurgi, a European engineering company (see Figure 2-4, functional schematic of Atmospheric CFBC). The first industrial application of their technology using coal was in Luenen, West Germany in 1982. This particular facility has a thermal capacity of 84MW, burning coal washery wastes with moisture content of up to 65% and 2% sulphur content. The plant generated process steam as well as electric power for a nearby aluminium oxide plant (Darling (1999), Bland et al. (1999)).

In the CFB of 250MW$_e$ nominal capacity, the flue gas moves at speeds of 3-6m/s carrying with it tons of silica sand, lime and ash up the roughly 40m high combustion chamber, together with coal. This material which is carried along with the gas stream is separated in a cyclone. Inside the cyclone the gas is forced into a swirling motion, throwing solids particles to the outside walls of the cone shaped collection unit. The solids are delivered to a hopper to be returned to the combustion chamber, (Hoy and Gill (1987)). The re-circulation of fly ash has a positive influence on the combustion efficiency (over 99% carbon burnout for all fuels), better utilisation of the limestone in absorbing sulphur dioxide and by cooling the fluidised bed by up to 50^{o}C. By this process of reducing the temperature, it helps to maintain optimum sulphur removal (Fuierer and Brokelmann (1997)).

The advantages of CFBC over BFBC presented by are:

- Achieved better rates of combustion than bubbling FBC, simply because circulating bed materials remain in the bed chamber longer and are burned up more efficiently, (Kakaras (2001)).
- Reduced levels of SO_2 and NO_x emissions compared to bubbling fluidised beds, (Kakaras (2001)).
- Combustion and sulphur retention efficiencies are improved, which results from the use of finer particles, turbulent gas-particle mixing and a high recycle rate, (Merrick (1984)).
- The use of high fluidising velocities makes a CFBC much smaller in bed area than a BFBC of the same output, (Merrick (1984)).
- The number of fuel feed points is reduced substantially due to the small combustor size and turbulent mixing conditions, (Merrick (1984)).
- Convective heat transfer co-efficient are increased because of the use of smaller particles. As a result, less heat transfer tubing would be required in the fluidising bed cooler than in the bed of a BFBC, (Merrick (1984)).

The disadvantages of CFBC over BFBC research by Merrick (1984) are:

- High auxiliary power requirement for circulation of bed material, (Kakaras (2001)).
- Erosion problems from ash carried along in the flue gas, (Kakaras (2001)).
- The pressure drop across a CFBC is generally greater than for BFBC. This results in increased fan power requirements, (Merrick (1984)).
- The large recycle rates require high efficiencies of the cyclone for the recovery of the bed solids from the gas stream, (Merrick (1984)).
- High gas velocities combined with the high particulate loadings may lead to erosion in the combustor, cyclone and associated ducting, (Merrick (1984)).

The types of CFBC have been detailed by Kakaras (2001) as follows,

- FBC with internals circulation: The materials are circulated in the combustion chamber. The air nozzle tray enables circulation at locally different levels of fluidisation. Peripheral circulation systems are not required such as, feedback cyclones, fluidised bed cooler.
- FBC with external circulation: In this process the materials are circulated at an operating point close to elutriation and require peripheral systems such as feedback cyclones and fluidised bed coolers.

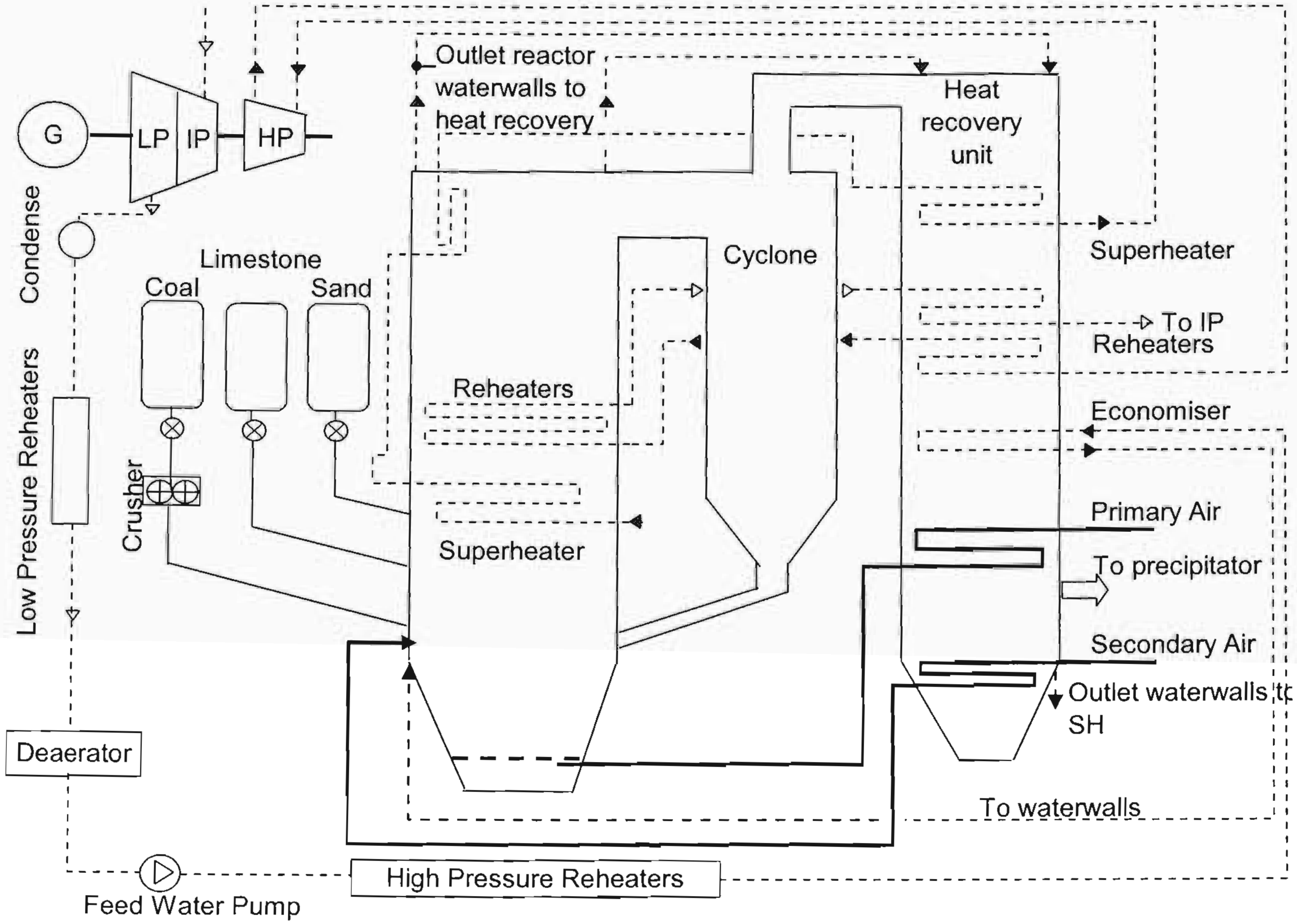

Figure 2-4 Functional schematic of Atmospheric CFBC (ACFBC), Kakaras 2001

2.5.2 BUBBLING FLUIDISED BEDS

The Atmospheric Fluidised Bed Combustor (AFBC) was an American idea which originated in the petroleum industry in 1942 where they investigated use of a bubbling bed (*Atmospheric Fluidised bed Combustion* (2006)). The environmental benefits of coal combustion in bubbling fluidised beds attracted worldwide interest in the mid 1960's, stimulated further by the oil embargo in 1973 and sustained by its fuel flexibility. In 1975, European companies installed the first major coal-fired bubbling AFBC in Renfrew, Scotland with a capacity of 18 t/h steam. The USA followed with a relatively large bubbling AFBC at Monongahela Power Co. in Rivesville, West Virginia. Development of bubbling AFBC continued in the USA with the construction of a 20 MWe pilot plant in 1982 by TVA and a 160 MWe demonstration plant in 1988.

Research by Zhangfa (2003) stated that in pressurised bubbling fluidised bed combustion, an increase in pressure results in smaller and more frequent bubbles being formed, which provides smoother fluidisation. This leads to better gas and solid contact (mixing), which then produces higher combustion efficiency. During the pressurised operation, coal and sorbent has to be fed across a pressure boundary and ash must be extracted. The lock hopper allows the pressurised operation.

Merrick (1984) showed PBFBC achieves considerably higher efficiencies of combustion and sulphur retention. He concluded the following are influencing factors,

- At elevated operating pressure fluidisation tends to become smoother, resulting in better gas-solid contacting.
- With elevated pressure operation, deeper beds of 2 to 4m can be used. This is because the additional pressure drop associated with a deep bed has less effect on the compression energy requirements than in the case of atmospheric pressure operation. For deep beds in atmospheric beds require high fan power requirements.
- As the throughput per unit of bed area increases with the pressure, adequately high throughputs can be obtained at moderate fluidising velocities of 1 to 2m/s.

2.6 CURRENT AND FUTURE DEVELOPMENTS IN FBC POWER PLANTS

Section 2.5 dealt with methods of classifying FBC boilers in terms of operating velocity (bubbling or circulating denoted as BFBC and CFBC respectively) and pressure (PBFBC and PCFBC). This section examines some of the current and future developments in FBC boilers. The flue gas emissions of some FBC boilers are also presented.

Table 2 - 1 Current and future developments in FBC boilers (Zhangfa (2006))

Overview of FBC technologies				
	BFBC	CFBC	PBFBC	PCFBC*
Development stage	well established, commercial	well proved, commercial	proved, commercial	pilot test
Application areas	primarily industrial boilers, also small-scale power generation industrial boilers	power generation,	power generation	power generation
Technology status				
Unit size (in operation)	up to 300 MWth	up to 300 MWe	up to 360 MWe	
Steam conditions	all subcritical	all subcritical with one supercritical being built	7 subcritical and 1 supercritical	
Plant efficiency (LHV)	typically 30% with industrial boilers‡	38–40% with subcritical 43% with supercritical	up to 44% with subcritical 44% with supercritical	
Plant availability, %	>90	90-98	not available	
NOx emissions, mg/m³	<400	<400	<400	
Sulphur retention efficiency†	lower than that of CFBC/PBFBC	typically 90% at a Ca/S ratio of 2	typically 90% at Ca/S ratios of 1.8–2.0	
Particulate emissions, mg/m³	<50 with ESP/bag filter	<50 with ESP/bag filter	3.5–76	
Future technological developments	continuous improvements	major developments	uncertain	uncertain
Market potential	favourable	strong	unfavourable	unfavourable

* potential better technical and environmental performances than other forms of FBC
† comparable to that of PCC+wet FGD but with lower sorbent utilisation
‡ potential higher efficiencies with power plant applications

The technological status of FBC has been summarised by Zhanfa 2006 (Table 2-1). The BFBC technology is well established, a major technological break through is not expected. There are major research and development programs for the CFBC technology.

Table 2 - 2 Atmospheric FBC Plants

Power Plant	**Specification**	**Emissions**
The 250MW$_e$ Provence Gardanne CFB boiler, France (1997), figure 2-5 is a schematic of the plant (Charet et al. (1997))	250MW$_e$, 700 t/hr, 567°C, 163bar	SO_2 400mg/Nm^3 @6%O_2, sulphur capture 97%, at Ca/S ratio <3 NOx 250mg/Nm^3 @6%O_2 Particulates 150mg/Nm^3 @ 6%O_2
La Pereda CFB in Oviedo Spain (Clean Coal Technologies Handbook (1995))	50MW$_e$, 185.2t/hr, 530°C	SO_2 emission 390mg/Nm^3 NO_x emissions 60mg/Nm^3 Particulates 15mg/Nm^3
The 125MWe Emile Huchet CFB boiler France (Clean Coal Technologies Handbook (1995))	125MW$_e$, 367t/hr, 134bar, 545°C	SO_2 emissions 150mg/Nm^3 @6%O_2, 95% sulphur capture with Ca/S =1.8 NO_x emission 100-110mg @6%O_2
CFBC Slough Heat and Power Ltd UK (Clean Coal Technologies Handbook (1995))	15MW$_e$, 94t/hr, 87bar, 509°C	SO_2 emission 1756mg/m^3 NOx emission 502mg/m^3 Particulate emission 51mg/m^3
CFBC Caledonian Paper Plc UK (Kakaras et al. ((2001))	42.6MW$_{th}$, 1806kg/s	SO_2 emissions 1650mg/Nm^3 NO_x emissions 650mg/Nm^3 Particulate emission 100mg/Nm^3
CFB Grenna co-generation plant (VGB (1999))	18MW$_e$, 29kg/s ,92bar, 505°C	SO_2 emission 100mg/MJ NO_x emission 150mg/MJ Particulate emission 50mg/Nm^3
The Czech retrofitting programme CFB K11 (CSFTA (2000))	150 MW$_{th}$, 29kg/s, 92bar, 505°C	SO_2 emission 400mg/m^3 NO_x emission 280mg/m^3 Particulate emission 9mg/m^3
Alholma CFB boiler, Finland, (Arhippainen et al. (2000))	240MW$_e$, 194kg/s ,165bar, 545°C	SO_2 emission 140mg/MJ NO_x emission 50mg/MJ Particulate emission 30mg/Nm^3

The emissions of some of the atmospheric FBC's in service have been detailed in Table 2-2. A schematic of the Provence Gardanne CFBC is presented in Figure 2-5. Examining both Table 2-1 and 2-2 the following conclusions can be drawn for the FBC boilers in operation:

- NO_x emissions <400mg/m^3.
- Typical sulphur retention is 90% for CFBC and PBFBC.

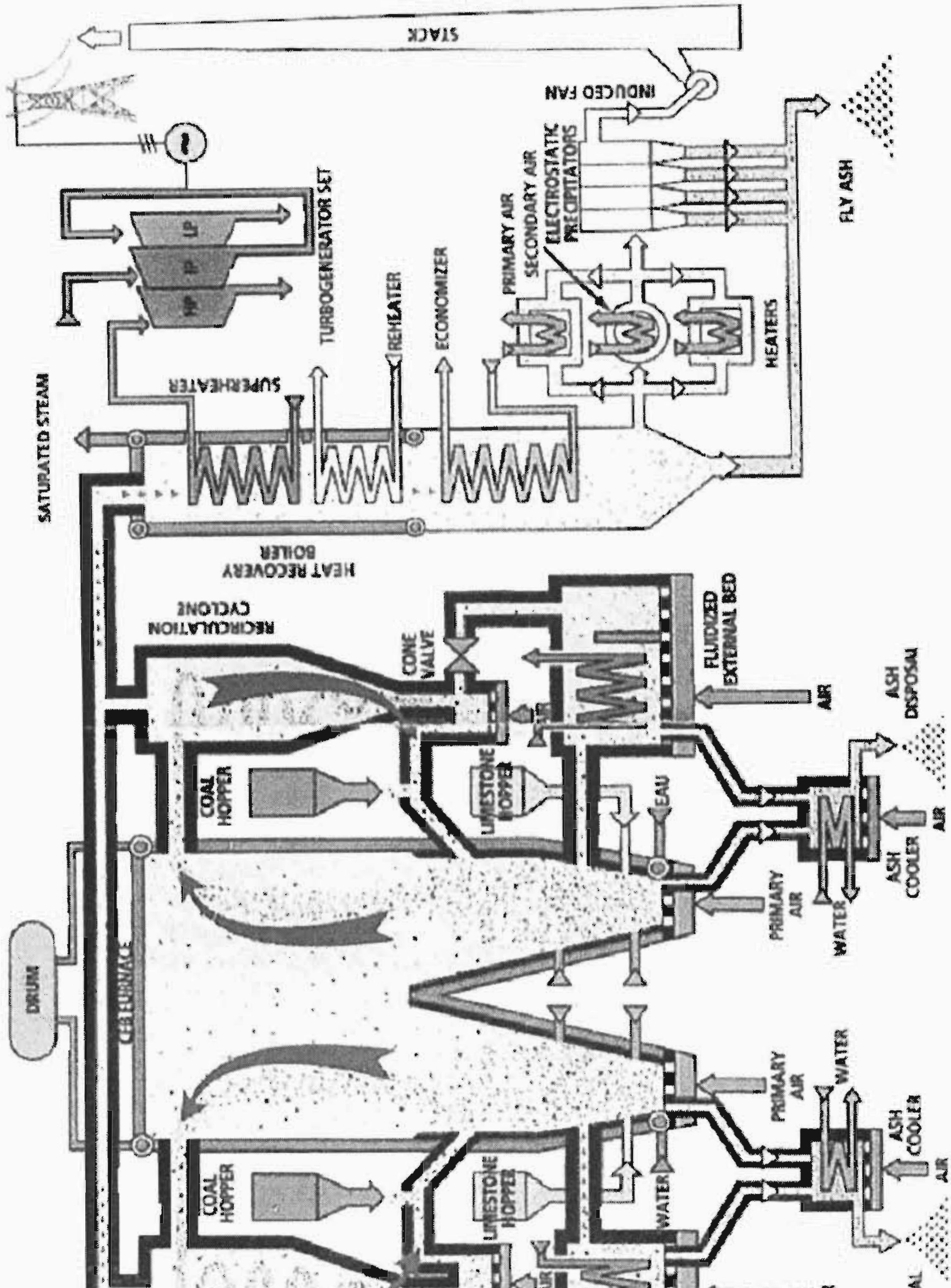

Figure 2 - 5 The 250MWe Provence / Gardanne ACFBC Power Plants (Charet et al. (1997))

2.7 FLUIDISED BED RESEARCH ON LOCAL SORBENTS

The reactivity of only the sorbent can be determined with laboratory scale equipment. This type of experimentation involves using a bench scale reactor which fluidises a bed of sorbent particles. The fluidising gas is composed of a known concentration of SO_2. Measurement of the reactor exit gas concentration of SO_2, allows the experimenter to calculate retention of SO_2 based difference of the concentrations SO_2 into and exiting the reactor. Local sorbents have been characterised using this method by Govender K (2006). The effect of combustion on the reactivity of the sorbent can be examined through a comparison of the bench scale test data and the FBTF results.

2.8 SUMMARY OF LITERATURE REVIEW

- The FBC technology produces higher carbon in ash and N_2O emissions than current coal fired power plants. The extent of this will be evaluated for the test coals.
- Fluidising velocity has an effect on bed pressure drop. To ensure consistency between tests, all tests must be conducted at the same velocity.
- Particle size has an effect on bubbles in the bed, and the small diameters of experimental beds, constrains the size of the bubbles. The solid concentration in the bed does influence air velocity fluctuation.
- Air flow in cross section across the fluidised bed is not uniform. It is generally seen that the air velocity has a maximum at the bed centre and decreases gradually with an increasing distance from the centre.
- The drying of coal entering the bed can slow down devolatilisation and ignition of some coals. The coals once prepared must be stored in sealed drums to prevent changes in the surface moisture of coal.
- The rate of coal devolatilisation is controlled by three mechanisms, chemical reaction kinetics, heat transfer to and within the particle, and mass transfer of volatile products within the particle.
- Volatiles in bubbling fluidised beds may burn either in a solid free bubble phase or in the spaces between the solid particles. The phase in which combustion occurs depends on the bed temperature.
- The char particle is hotter than its surroundings and the particle surface temperature is up to 100°C higher than the bed temperature. This is important since if the bed particles exceed 1000°C, there is a possibility of producing agglomerates.
- Generally circulating fluidised bed combustors have higher combustion efficiencies than bubbling fluidised bed combustors, due the high solid recycle rate.
- Volatile content allows for greater reactivity and greater char combustion rate, higher combustion efficiency.
- A higher bed temperature produces higher reactions rates and higher combustion efficiencies.
- An increase in fluidising velocity decreases combustion efficiency.
- Excess air increase means an increase in mean oxygen concentrations in the bed, therefore combustion efficiency increases.

- Although FBC produces lower NO_x emissions (<400mg/m^3), it does produce significant quantities of nitrous oxide (N_2O) emissions. This is due to the lower combustion temperatures. The N_2O emissions range from 20 to 300ppm compared to 10ppm found in conventional pulverised fuel combustion.
- The effect of temperature showed that as temperature was increased from 800 to 850°C, the conversion to NO_x increased from 30 to 50% while conversion to N_2O decreased from 11 to 8%.
- As excess air is increased, it leads to an increase in both NO_x and N_2O.
- Air staging substantially reduces both NO_x and N_2O.
- The addition of limestone influences both the formation and reduction of NO. In a circulating fluidised bed combustor, an increase in NO emissions where observed.
- The reactivity of the sorbent is influenced by the pores in the sorbent particles.
- The maximum retention of SO_2 for the coals tested by Zhangfa (2003) occurred at a bed temperature of 850°C.

Chapter Three

EXPERIMENTAL

3.1 INTRODUCTION TO THE EXPERIMENTAL INVESTIGATION

A literature review of the FBC technology was discussed in chapter two. Some of these findings such as fuel preparation, operational conditions will be implemented in the experimental investigation. The need for fluidised bed coal combustion tests was also introduced in chapter one. This chapter concentrates on the experimental equipment, the operating test conditions, the test matrix, and the chemical analyses of the tests coals and sorbent.

3.2 DESCRIPTION OF EXPERIMENTAL EQUIPMENT

The Fluidised Bed Test Facility (FBTF) is a bubbling FBC which can be operated at either atmospheric or pressurised conditions up to 3 bar (gauge). A bubbling bed is differentiated from that of a circulating bed by its operational velocity, which was expanded on in section 2.5. The purpose of this facility is to examine the fluidised bed combustion or gasification properties of South African coals utilising locally available sorbents. A schematic of FBTF is provided in Figure 3-1. This schematic indicates the point of entry of the test coal and sorbent, fluidising air and the exit of points of the products of fluidised bed combustion (flue gas, fly ash and course ash). The key processes of this facility are as follows:

- Fuel and sorbent preparation.
- Chemical analysis of coal and sorbent.
- Reactor pre-heating.
- Combustion tests of coal and sorbent.
- Calculations

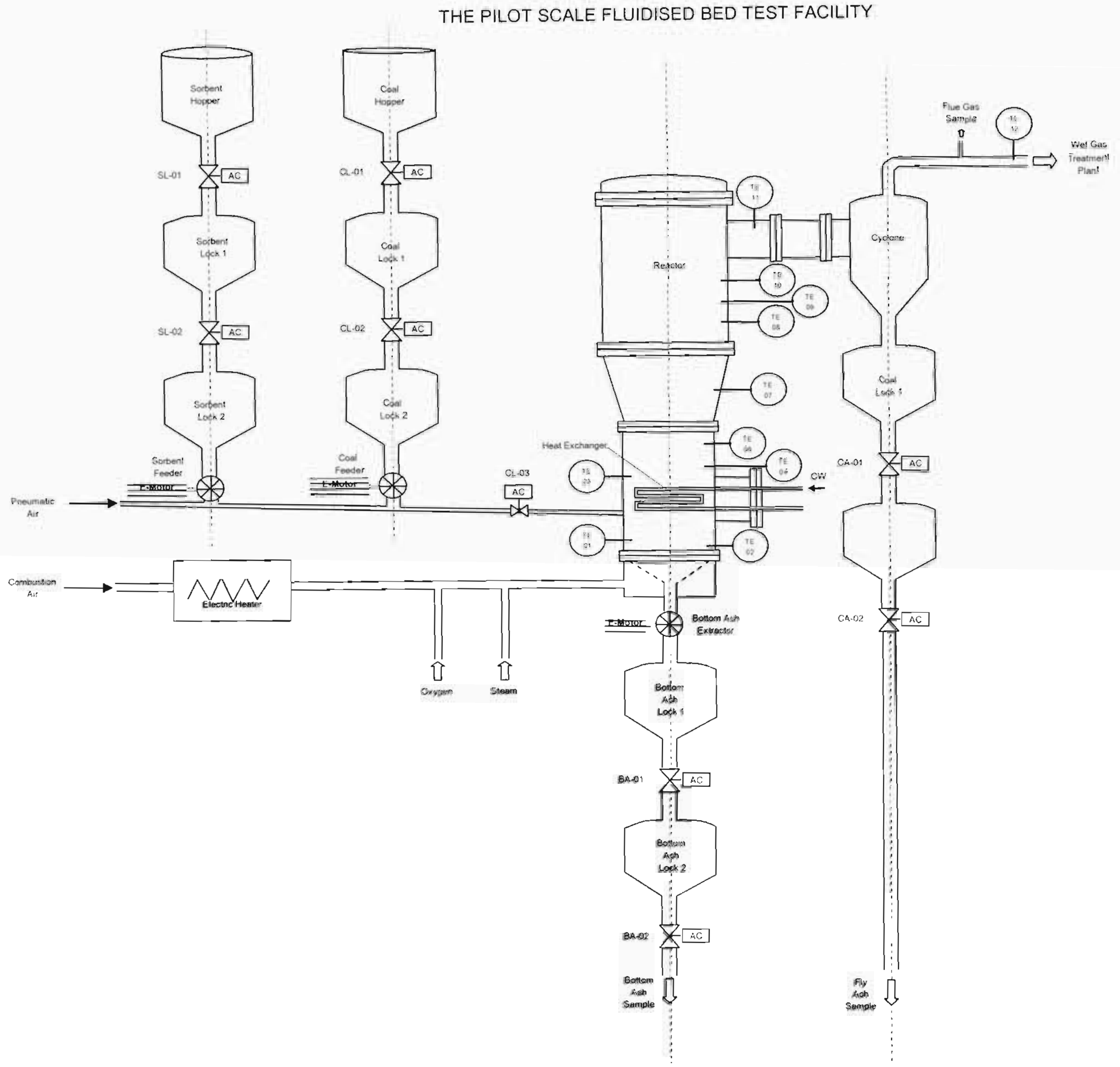

Figure 3 - 1 The Fluidised Bed Test Facility

3.2.1 THE FUEL AND SORBENT PREPARATION

In the literature review it was identified that coal surface moisture content does affect the evaporation rate of moisture and slows devolatilisation and the process of ignition (section 2.4.1). Mitigation to prevent coal moisture increasing through exposure to rainfall or being dehydrated was to store the coal in sealed drums. The surface moisture content of coal and sorbent also affects the preparation process. A wet sorbent is difficult to crush and forms a paste which clogs the crusher screens. The process of preparation begins when the raw coal or sorbent is loaded into the crusher hopper. The volumetric feedrate into the crusher is controlled. The size fraction of the coal or sorbent post crushing is less than 10mm (screen size, Figure 3-2). The particle size at this stage is too large for fluidisation or pneumatic transportation. The particles must be re-crushed to decrease the larger size fractions. Traditionally the larger size fractions were separated out manually by Eskom staff with a screen for the coal and sorbent, 2.2mm and 1mm respectively. This was an extremely labour intensive process. An electrically driven screen was designed and fabricated during the process of this dissertation which reduced the time required for coal and sorbent preparation (Figure 3-3).

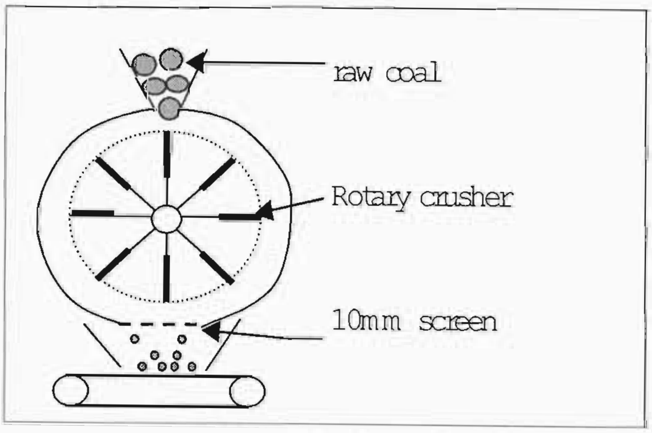

Figure 3 - 2 Crusher Schematic

Figure 3-3 shows the collection drums for the oversize and undersize product. The drums are positioned below the siever. The oversize product is then re-crushed.

Figure 3 - 3 The siever

On completion of re-crushing the oversize product, must now be mixed with the undersize product. This is because the crushing and sieving process has segregated the dense and less dense material. The method employed for mixing is cone and quartering. A brief explanation of this method is that the coal or sorbent is first unloaded on a clean concrete surface. The product is thoroughly mixed and arranged in the shape of a cone, by manual labour using a spade (Figure 3-4). Thereafter a quarter of the cone is removed and placed aside. The remaining three quarters are re-arranged in the shape of a quarter and the process re-starts. Each time the process continues a different quarter is taken. The process of cone and quartering continues many times until the remaining cone can be placed in sample bags (50kg bag). This is a representative sample and can be analysed for its chemical composition.

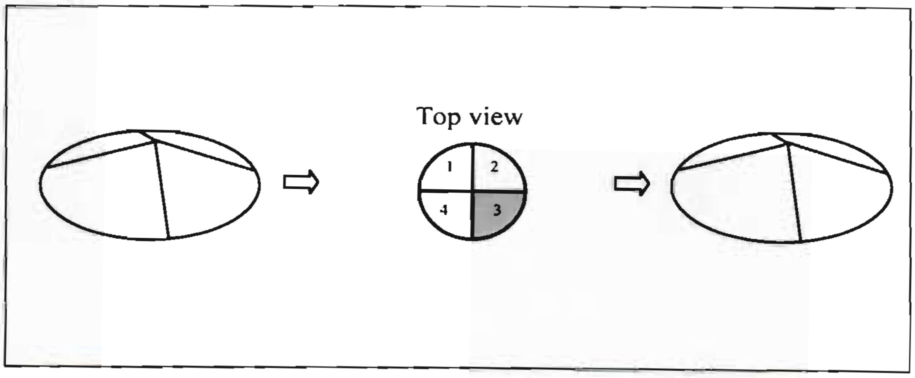

Figure 3 - 4 Stepwise process for cone and quartering

3.2.2 COAL AND SORBENT CHEMICAL ANALYSES

Preparation of representative samples for coal and sorbent was discussed in the previous section. This section details the results of the chemical analyses performed on the representative samples for coal and sorbent. Three test coals and one sorbent was analysed. The following chemical analyses were performed on the test coals samples at the Eskom Coal and X-ray laboratory:

1. Ultimate analysis
2. Proximate analysis
3. Ash elemental
4. Surface moisture
5. Particle size distribution
6. Gross CV
7. Ash fusion temperatures

The chemical results for Coal A, B, and C are presented in Tables 3-1. The sorbent chemical analyses are presented in Table 3-2.

Table 3 - 1 Chemical analyses results for Coal A, B and C

PROXIMATE AND ULTIMATE ANALYSIS			
	Coal A (%)	Coal B (%)	Coal C (%)
Ash (Total)	42.40	41.80	38.20
Volatile matter	21.00	20.00	20.20
Fixed carbon (by difference)	33.80	36.00	38.30
Carbon	41.37	43.37	46.42
Nitrogen	1.03	1.05	0.99
Hydrogen	2.72	2.27	1.69
Total sulphur	0.98	3.65	1.82
Carbonate (as CO_3)	3.17	2.93	1.53
Oxygen (difference)	5.53	2.69	6.05
Inherent moisture	3.00	2.20	3.00
Surface moisture	1.80	3.00	1.80
Total moisture	4.80	5.10	4.80
ASH ELEMENTAL			
Silicon (as SiO_2)	52.90	50.5	47.3
Aluminium (as Al_2O_3)	33.90	24.4	26.5
Iron (as Fe_2O_3)	3.33	11.7	8.2
Titanium (as TiO_2)	-	1.2	1.3
Phosphorus (as P_2O_5)	0.47	0.47	0.58
Calcium (as CaO)	3.51	6.1	7.5
Magnesium (as MgO)	0.98	1.7	1.8
Sodium (as Na_2O)	0.18	0.01	0.01
Potassium (as K_2O)	0.97	0.8	0.6
Sulphur (as SO_3)	1.72	3.1	7.1
Manganese (as MnO)	-	0.06	0.05
Gross Calorific Value (MJ/kg)	16.85	16.47	17.13

Table 3 - 2 Chemical analyses results for Sorbent A

SORBENT A		
CALCIUM ANALYSIS		
Ca	36.78	%
$CaCO_3$	91.96	%

3.2.3 REACTOR HEATING

On completion of the coal and sorbent preparation (previous section), the heating of the reactor can now take place. The reactor is the heart of the plant where the coal combustion reactions take place. This pressure vessel is 7m high and is refractory lined (Figure 3-5 shows the reactor bed zone region). The bed diameter is 230mm and extends to a diameter of 360mm in the free board region. There are six thermocouples located on the reactor wall (see Figure 3-1, TE01-06) in the bed zone region. There are also a number of thermocouples located on the wall above the bed region (TE07-10). A removable submerged heat exchanger (bed coil) is inserted for bed temperature control for higher feedrate of coal. Alternatively a refractory line plug can be inserted for lower feedrate of coals. The bed level is controlled by a variable volumetric type bottom ash extractor at the base of the reactor.

The process begins by preheating the reactor (electric air heaters depicted in Figure 3-1) with the combustion air until a bed zone temperature of 450°C is attained. The combustion and transport air is reduced to maintain a bed velocity of 1m/s. At this velocity sufficient mixing occurs of the bed, but extensive entrainment of bed particles does not occur (bubbling regime, discussed in section 2.3.1). A bed in reactor must now be created. To achieve this, inert material (13kg of bottom ash, usually obtained from previous tests) is fed into the reactor, and heated with the combustion air. The bed material is heated until a temperature greater than 350°C is attained. Pre-crushed charcoal with a fineness of less than 10mm is introduced into the bed (charcoal has a lower density than coal, if fine char particles are used, entrainment will occur). On ignition of the charcoal, coal is fed a kilogram at a time and the changes in bed temperature are monitored. Once ignition of the coal occurs, a stable feedrate is chosen to allow for gradual heating of the reactor. The bed temperature is controlled by varying the flow rate of cooling water to the bed coil or if the refractory plug is used, the coal feedrate is used as the measure of control. There are look-out glasses located on the reactor outer wall (Figure 3-5, shows four of them vertically, Figure 3-6 is a view inside one of them) which allows the bed height to be monitored visually.

Due to the fact that the reactor is a pressurised system, a pressure barrier exists between the reactor and the atmospheric air. To ensure introduction of coal and sorbent into the reactor, the facility is designed with a double air lock system. The coal and sorbent feeding systems are both identical, therefore only the coal feed system will be discussed. The introduction of coal from coal lock 2 (Figure 3-1) is controlled by a variable speed volumetric feeder.

Conveyance of coal into the reactor is done pneumatically, and the mass flow rate of air used is regulated with a control valve and flow rate measured.

Figure 3-5 depicts the reactor bed zone region.

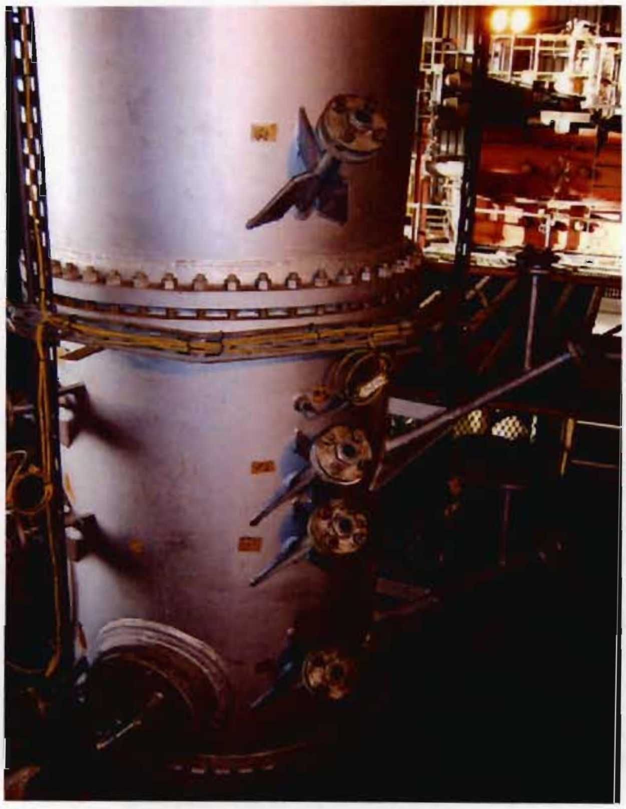

Figure 3 - 5 The Reactor

Figure 3-6 is a view through one of the look-out glasses. This allows the operator to examine the bed mixing and height of the bed.

Figure 3 - 6 The Reactor bed as seen through the top port

3.2.4 COAL AND SORBENT COMBUSTION TESTS

The process of preheating the reactor has been discussed (section 3.2.2). This section discusses the coal and sorbent combustion tests. The literature review discussed the need for consistency between different tests for fluidising velocity (section 2.3.1), operating pressure and bed height (section 2.3.3). For the purpose of coal and sorbent fluidised bed combustion tests, the reactor was stabilised to the operating conditions detailed in Table 3-3.

Table 3 - 3 Operating conditions of the FBTF

Temperature range	800-900°C
Reactor pressure	50kPa (gauge)
Reactor gas velocity	1m/s
Bed height	1m
Ca/S mole ratio range	0-1.7

Section 3.2.2 explained that coal combustion reactions takes place in the reactor. The product of combustions (hot flue gas and particulates) exits the reactor and enters the cyclone through a connecting pipe. The cyclone essentially serves to remove particulates. These particulates are referred to as fly ash. Fly ash samples are extracted during tests by this system, to test for combustion efficiency and sulphur retained. After the flue gas passes through the cyclone, online gas sampling takes place. The sample point is located just before TE12 (Figure 3-1). Although a large portion of the fly ash has been removed, a sintered filter of 20μm size must still be used to remove very small particles. The gas sample must be conditioned before the analyser can accept this sample. The reasoning behind this is that the moisture in the sample gas may condense before it reaches the analyser. This will then cause soluble gases such as SO_2 and NO_x to dissolve. This results in a much lower SO_2 and NO_x measurement being recorded. The moisture must be first removed. A Peltier Cooler is used to achieve this.

The refractory plug was used during the tests of Coal A, B and C. The bed temperature was controlled through the coal feedrate. Due to the fact that the bed velocity is maintained at 1m/s, changes to the air flow cannot be made. This does affect excess air setting (a mixture were the air quantity is greater than stoichiometric requirements for that fuel). This section discussed fluidised bed combustion tests but also the FBTF can be operated as a gasifier.

The objectives of this study were to investigate the changes in the production of NO_x, SO_x, and the combustion efficiency of the three test coals under conditions of fluidised bed combustion, with the same sorbent. Table 3-4, Table 3-5 and Table 3-6 details the test matrix for Coal A, B and C respectively. Tests with no sorbent were performed to evaluate the coals inherent calcium capabilities of capturing sulphur. Tests with varying ratios of sorbents were performed to evaluate the sorbents capabilities for further levels of in-bed desulphurisation.

Table 3 - 4 Test matrix Coal A

Test number	**Bed Temperature**	**Ca/S**
1	800°C	0
2	820°C	0
3	840°C	0
4	860°C	0
5	880°C	0
6	900°C	0
7	800°C	0.5
8	820°C	0.5
9	840°C	0.5
10	860°C	0.5
11	880°C	0.5
12	900°C	0.5
13	800°C	1
14	820°C	1
15	840°C	1
16	860°C	1
17	880°C	1
18	900°C	1

Table 3 - 5 Test matrix Coal B

Test number	**Bed Temperature**	**Ca/S**
19	800°C	0
20	820°C	0
21	840°C	0
22	860°C	0
23	880°C	0
24	900°C	0
25	900°C	1
26	880°C	1
27	860°C	1
28	840°C	1
29	820°C	1
30	800°C	1
31	800°C	1.5
32	820°C	1.5
33	840°C	1.5
34	860°C	1.5
35	880°C	1.5
36	900°C	1.5

Table 3 - 6 Test matrix Coal C

Test number	**Bed Temperature**	**Ca/S**
37	800°C	0
38	820°C	0
39	840°C	0
40	860°C	0
41	880°C	0
42	900°C	0
43	800°C	1.1
44	820°C	1.1
45	840°C	1.1
46	860°C	1.1
47	880°C	1.1
48	900°C	1.1
49	800°C	1.4
50	820°C	1.4
51	840°C	1.4
52	860°C	1.4
53	880°C	1.4
54	900°C	1.4
55	800°C	1.7
56	820°C	1.7
57	840°C	1.7
58	860°C	1.7
59	880°C	1.7
60	900°C	1.7

3.2.5 FBTF CALCULATIONS

The process of FBTF combustion testing has been discussed (section 3.2.3), where the production of NO_x, SO_x, and the combustion efficiency of the test coals and sorbent were recorded. Also recorded during testing are process parameters such as, O_2, CO, CO_2, reactor pressure, air flowrate, coal and sorbent feedrate. These recorded values, the coal and sorbent chemical analysis (section 3.2.2) will be used in the FBTF calculations, which are discussed in this section. The process of FBTF calculations can be summarised as follows:

- Conversion of chemical composition for coal from air dried to a dry basis.
- Normalising the gas species measurements to 6% O_2 and converting to mg/Nm^3 @6% O_2.
- SO_2 retention.

3.2.5.1 The conversion of the chemical composition for coal from air dried to as received

The percentages of carbon, nitrogen, hydrogen, total sulphur, oxygen and ash in coal must be converted from an air dried basis (the standard reported form of Eskom coal laboratories) to an as received basis (Equation 3-1 shows the conversion for the carbon component in coal). These chemical analyses are used in flue gas calculations.

$$W_{\text{as received},C} = W_{\text{air dried},C} \times \frac{\left(100 - W_{moist,\text{total}}\right)}{\left(100 - W_{moist,\text{inherent}}\right)} \qquad 3\text{ - }1$$

3.2.5.2 Normalising the gas species measurements to 6% O_2 and converting to mg/Nm^3 @6% O_2

The preferred method of reporting the concentrations of flue gas species measurements in the power industry is to normalise the measurement to 6% O_2. This is because different combustion technologies operate at different excess air settings. Equation 3-2 details the method for converting the measured concentration of CO_{ppm} to a normalised value at 6% O_2.

$$X_{CO,@6\%O_2,ppm} = \frac{X_{CO,ppm}}{\left[1 + \frac{\left(6 - O_{2,\%}\right)}{\left(20.95 - 6\right)}\right]} \qquad 3\text{ - }2$$

Similar calculations for CO_2 were done for the converting from ppm to ppm @6% O_2.

Another method of reporting flue gas emissions is in mg/Nm^3 @6% O_2. The normalised ppm value of the gas measurement is multiplied by the gas density at STP (101.3kPa, 273.15K). Equation 3-4 details the calculation for mg/Nm^3 of CO @ 6% O_2.

$$\rho_{CO,STP} = 1.25 \qquad 3-3$$

$$X_{CO,mg/Nm^3} = X_{CO,@6\%O_2,ppm} \times \rho_{CO,STP} \qquad 3-4$$

Similar calculations for CO_2 is done for converting from ppm @6% O_2 to mg/Nm^3 @6% O_2. The density of CO_2 is presented in Equations 3-5.

$$\rho_{CO_2} = 1.96 \qquad 3-5$$

3.2.5.3 SO_2 retention

This section details the procedure for calculating SO_2 retention. The sulphur into the reactor is calculated from the mass flow of coal and the fraction of sulphur in coal. (Equation 3-6).

$$\text{Mass flow rate of sulphur into reactor } \dot{m}_S = \dot{m}_{coal} \times \frac{W_S}{100} \qquad 3-6$$

The maximum quantity of SO_2 per hour that can be produced assumes all the sulphur will be converted to form SO_2 gas (Equation 3-7). The molar mass of SO_2 is twice that of molar mass of sulphur, therefore the mass flow of SO_2 is twice the mass flow of sulphur entering the reactor.

$$\dot{m}_{max,SO_2} = \dot{m}_S \times 2 \qquad 3-7$$

Due to corrosion of the flue gas mass flow meter, the mass flow of flue gas exiting the reactor could not be measured and is calculated according to products of combustion spreadsheet developed by Keir (1991). Having calculated the mass flow of flue gas, volumetric flowrate of the flue gas is calculated according to equation 3-9.

$$\rho_{flue,STP} = 1.32 \qquad 3-8$$

$$\dot{V}_{flue,Nm^3/hour} = \frac{\dot{m}_{flue}}{\rho_{flue,STP}} \qquad 3-9$$

The volumetric flowrate of the flue gas must be corrected for temperature and pressure at the reactor sample point (Equation 3-10).

$$\dot{v}_{\text{actual flue}, m^3/hour} = \dot{v}_{\text{flue}, Nm^3/hour} \times \left[\frac{(T_{\text{sam}} + 273) \times 101.3}{(P_{\text{reactor}} + 101.3) \times 273} \right] \qquad 3-10$$

The volumetric flow rate and mass flow of SO_2 emissions is calculated according to equations 3-11 and 3-12 respectively.

$$\dot{v}_{SO_2, \text{emitted}, m^3/hour} = \left(\frac{X_{SO_2, ppm}}{1000000} \right) \times \dot{v}_{\text{actual flue}, m^3/hour} \qquad 3-11$$

$$\dot{m}_{SO_2, \text{emitted}} = \dot{v}_{SO_2, \text{emitted}, m^3/hour} \times \left\{ 2.86 \times \left[\frac{(P_{\text{reactor}} + 101.3) \times 273}{(T_{\text{sam}} + 273) \times 101.3} \right] \right\} \qquad 3-12$$

SO_2 retention is calculated from the maximum quantity of SO_2 that can be produced and the SO_2 that was emitted.

$$U_{SO_2} = \left(1 - \frac{\dot{m}_{SO_2, \text{emitted}}}{\dot{m}_{\max, SO_2}} \right) \times 100 \qquad 3-13$$

Chapter Four

RESULTS AND DISCUSSION

4.1 INTRODUCTION TO THE RESULTS AND DISCUSSION

The process of FBTF calculations have been discussed in section 3.2.4. The results of these calculations are discussed in this chapter (results of 60 tests). Three coals were tested on the same sorbent. The gaseous emissions were also normalised to 6% O_2 and also converted to mg/Nm^3 @6% O_2. This is the preferred way of reporting gaseous emissions in the power industry. To reduce the volume of results presented in this chapter, only the summary of each coal tested is discussed. Appendix A, B and C presents a detailed graphical analysis for Ca/S molar ratio for the coals tested. The effect of bed temperature on the products of combustion was evaluated for a temperature range of 800 to 900°C, in increments of 20°C.

4.2 COAL A AND SORBENT A TESTS

This section presents the results of Coal A and Sorbent A tests (Table 4-1). During Tests 1 to 6 no sorbent was introduced into the bed which enabled inherent sulphur capture to be evaluated. For Tests 7 to 12 and 13 to 18 Limestone (sorbent) was added at a Ca/S mole ratio of 0.5 and 1 respectively. The purpose of these tests was to investigate the sorbent capabilities of in-bed desulphurisation. The effect of bed temperature on products of combustion was evaluated for a temperature range of 800 to 900°C, in increments of 20°C.

Table 4 - 1 Results Coal A and Sorbent A

	Coal federate (kg/h)	Ave bed temperature (°C)	Total air (kg/h)	Fluidising air velocity (m/s)	Carbon cyclone ash (%)	Ca/S molar ratio	Operating pressure (kPa)	Sulphur cyclone ash (%)	O_2 (%)	CO (ppm)	CO (ppm @6% O_2)	CO (mg/Nm^3 @6% O_2)	CO_2 (%)	CO_2 (% @6% O_2)	CO_2 (mg/Nm^3 @6% O_2)	NO (ppm)	N_2O (ppm)	SO_2 (ppm)	SO_2 retention (%)
Test 1	4.14	794	64.7	1.00	5.4	0	51.5	1.48	13.1	105	199	248	8.19	15.5	30.4	162	101	247	65
Test 2	4.66	818	65.5	1.00	4.2	0	57.1	1.6	12.4	79.6	140	175	9.24	16.2	31.8	188	94.3	253	66
Test 3	4.71	838	65.2	0.99	4.0	0	55.5	1.72	11.9	65.7	109	136	9.75	16.1	31.6	164	85.7	256	68
Test 4	5.14	850	65.3	1.00	3.2	0	55.5	1.44	11.6	54.4	87.2	109	10.1	16.2	31.8	216	77.5	272	67
Test 5	5.89	885	65.8	1.02	2.6	0	57.7	1.32	10.2	45.3	63.2	78.9	11.5	16.0	31.4	230	63.1	444	52
Test 6	5.61	895	66.0	1.02	3.0	0	59.5	1.32	10.2	41.9	58.4	73	11.6	16.1	31.6	228	55.2	441	52
Test 7	3.18	798	57.9	0.90	2.7	0.5	47.6	1.28	14.3	72.4	163	203	7.49	16.9	33.0	182	75.0	173	71
Test 8	4.03	820	57.6	0.91	3.1	0.5	47.6	1.52	12.4	76.6	134	168	9.52	16.7	32.6	210	91.3	228	70
Test 9	4.68	842	57.6	0.93	2.8	0.5	47.9	1.48	11.0	67.1	101	126	10.6	15.9	31.2	225	88.3	272	69
Test 10	4.96	864	58.4	0.93	3.2	0.5	52.1	1.48	10.7	55.9	81.4	102	11.1	16.2	31.7	225	79.0	308	66
Test 11	5.34	882	57.9	0.95	2.4	0.5	49.8	1.28	9.79	50.3	67.4	84.2	12.2	16.3	31.9	229	74.6	384	60
Test 12	5.81	913	58.2	0.97	2.3	0.5	51.8	1.28	8.76	39.1	47.9	59.9	13.0	15.9	31.3	236	50.4	533	49
Test 13	3.78	801	57.0	0.89	2.3	1	48.1	1.32	13.5	61.7	124	155	8.39	16.9	33.1	196	72.1	139	79
Test 14	4.15	821	57.3	0.90	2.8	1	47.7	1.64	12.5	60.4	107	134	9.48	16.8	32.9	211	76.6	154	79
Test 15	4.79	844	57.1	0.93	2.6	1	47.4	1.76	10.8	53.5	78.6	98.3	11.0	16.1	31.6	216	75.1	188	79
Test 16	5.25	867	57.9	0.94	2.5	1	50.2	1.64	9.87	46.6	62.9	78.6	11.9	16.1	31.6	214	70.1	230	76
Test 17	5.44	884	58.1	0.95	2.3	1	51.1	1.52	9.03	40.2	50.4	63	12.6	15.8	31	209	56.9	307	70
Test 18	5.98	912	59.1	0.97	2.3	1	53.7	1.48	7.98	35.1	40.4	50.5	14.0	16.1	31.6	214	45.5	538	51

4.2.1 Discussion of Coal A and Sorbent A tests (Ca/S=0, 0.5, 1)

4.2.1.1 Carbon in ash

Carbon in ash decreased with an increase in bed temperature for Ca/S = 0 (Figure 4-1). The introduction of sorbent resulted in a constant carbon in ash trend. The increase Ca/S molar ratio from 0.5 to 1 had little impact on carbon in ash.

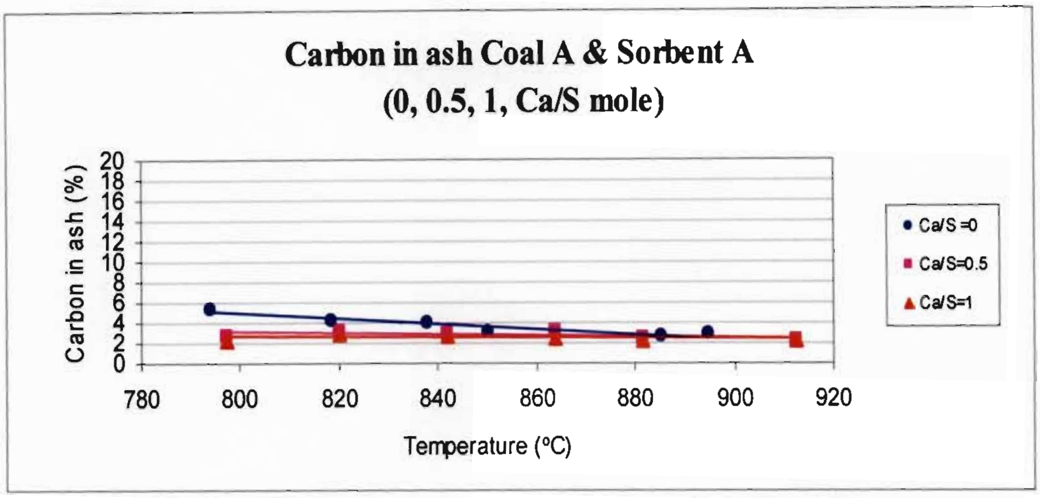

Figure 4 - 1 Carbon in ash for Coal A tests

4.2.1.2 Nitrogen oxide emissions

The NO emissions increased with an increase in bed temperature for Ca/S = 0, 0.5, 1 (Figure 4-2). The introduction of sorbent resulted in an increase in NO emissions for a bed temperature of 800 to 840°C. An increase in Ca/S molar ratio from 0.5 to 1 had little impact on NO emissions.

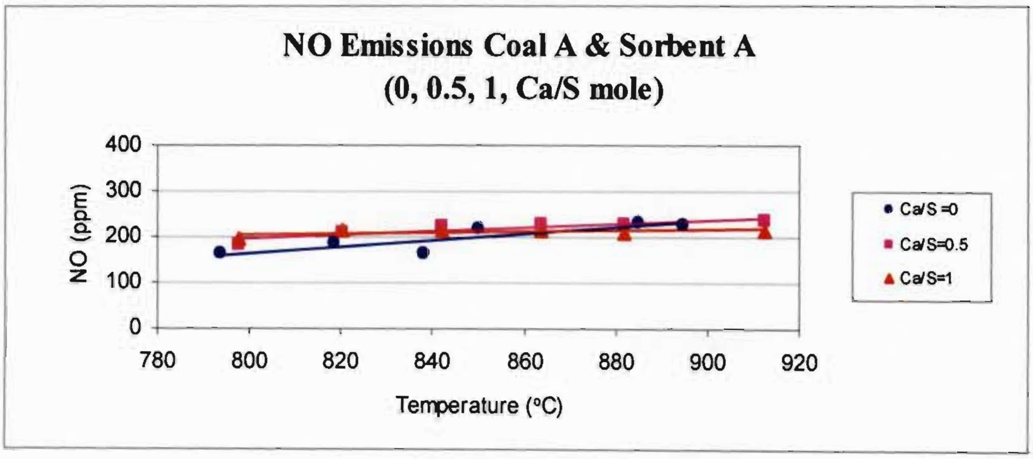

Figure 4 - 2 NO Emissions for Coal A tests

4.2.1.3 N_2O emissions

The N_2O emissions decreased with an increase in bed temperature for Ca/S = 0, 0.5, 1 (see Figure 4-3). The introduction of sorbent resulted in a decrease in N_2O emissions for a bed temperature of 800°C. An increase in Ca/S molar ratio from 0.5 to 1 resulted in a slight decrease in N_2O emissions.

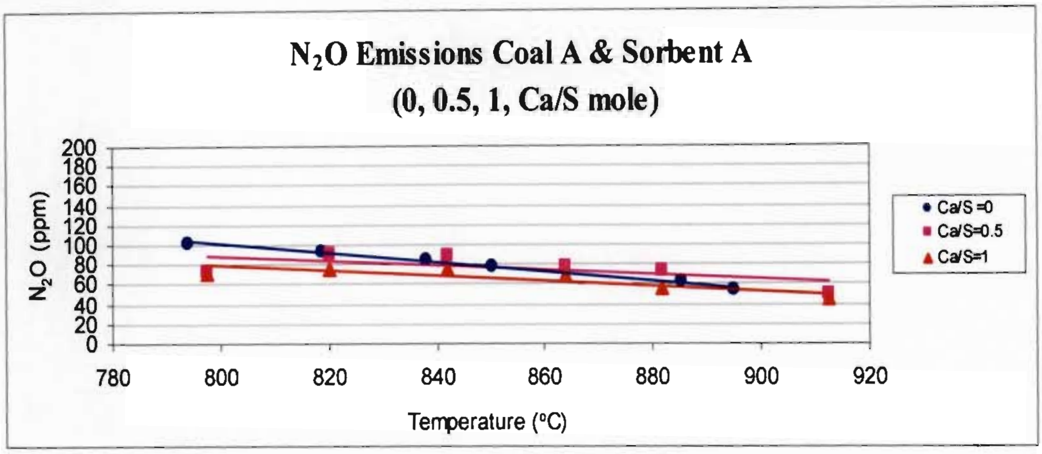

Figure 4 - 3 N_2O Emissions for Coal A tests

4.2.1.4 SO_2 Retention

A similar SO_2 retention trend was observed for Ca/S = 0, 0.5, 1 where the maximum SO_2 retention occurred in the 800 to 840°C bed temperature range (Figure 4-4). The introduction of sorbent at Ca/S = 0.5 showed little improvement in SO_2 retention. An increase in Ca/S molar ratio from 0.5 to 1 resulted in further increase in SO_2 retention.

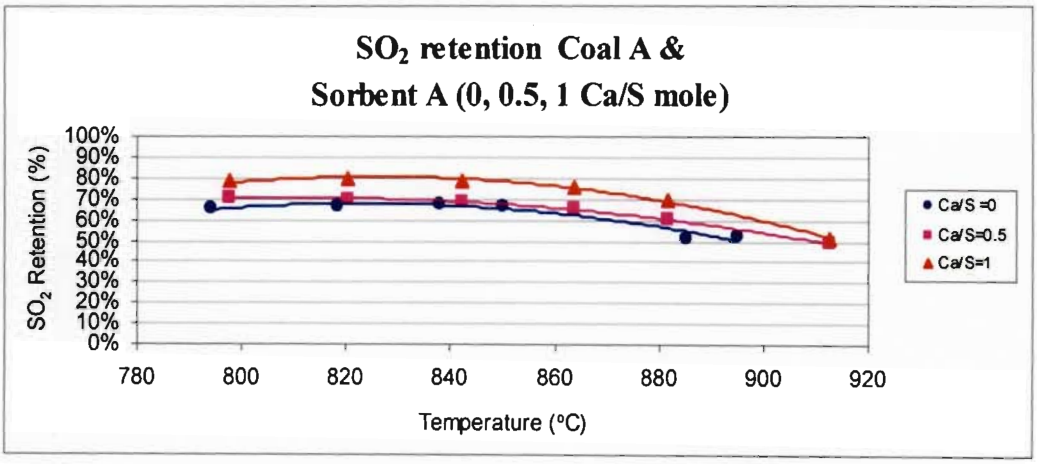

Figure 4 - 4 SO_2 retention for Coal A tests

4.2.1.5 Comparison of the SO_2 retention of sorbent A under bench scale tests of fluidisation with SO_2 gas only and under combustion conditions on the FBTF with Coal A

Table 4 - 2 Comparison of the SO_2 retention of sorbent A, with bench scale tests of fluidisation with SO_2 gas and the FBTF results of Coal A.

Bed temperature	**Maximum SO_2 retention (bench scale tests, fluidisation with SO_2 gas only Ca/S=2.85)**	**Maximum SO_2 retention (FBTF Ca/S=0)**	**Maximum SO_2 retention (FBTF Ca/S=0.5)**	**Maximum SO_2 retention (FBTF Ca/S=1)**
800°C	64.78%	61%	69%	79%
850-860°C	71.61%	65%	63%	74%
900°C	68.76%	47%	46%	46%

This comparison intends to examine if there are any similarities in the reactivity of the sorbent under conditions of fluidised bed combustion and fluidisation with SO_2 gas only. The bench scale test data for sorbent A were extracted from Govender (2006). Section 2.7 briefly explains the method bench scale testing. The bench tests details higher SO_2 retention in the 850-860°C temperature range. On the FBTF for a Ca/S = 0.5 to 1 the maximum SO_2 retention was observed in the 800-840°C temperature range.

4.3 COAL B AND SORBENT A

This section presents the results of Coal B and Sorbent A tests (Table 4-3). During Tests 19 to 24 no sorbent was introduced into the bed which enabled inherent sulphur capture to be evaluated. For Tests 25 to 30 and 31 to 36 Limestone (sorbent) was added at a Ca/S mole ratio of 1 and 1.5 respectively. The purpose of these tests was to investigate the sorbent capabilities of in-bed desulphurisation. The effect of bed temperature on products of combustion was evaluated for a temperature range of 800 to 900°C, in increments of 20°C.

Table 4 - 3 Results Coal B and Sorbent A

	Coal federate (kg/h)	Ave bed temperature (°C)	Total air (kg/h)	Fluidising air velocity (m/s)	Carbon cyclone ash (%)	Ca/S molar ratio	Operating pressure (kPa)	O_2 (%)	CO (ppm)	CO (ppm @6% O_2)	CO (mg/Nm^3 @6% O_2)	CO_2 (%)	CO_2 (% @6% O_2)	CO_2 (mg/Nm^3 @6% O_2)	NO (ppm)	N_2O (ppm)	SO_2 (ppm)	SO_2 retention (%)
Test 19	4.18	798	65.2	0.97	5.8	0	61.1	11.1	155	235	294	8.35	12.7	24.8	188	106	2323	27
Test 20	4.84	820	64.9	0.99	7.6	0	60.4	9.78	142	190	238	9.13	12.2	24.0	202	106	2571	29
Test 21	5.31	840	67	1.09	7.2	0	54.4	8.49	137	164	205	10.5	12.6	24.7	307	114	2763	32
Test 22	5.79	864	63.1	1.07	7.4	0	52.1	7.27	124	136	169	11.7	12.8	25.1	333	108	3129	29
Test 23	6.17	890	65.8	1.05	6.6	0	62.5	6.84	97	103	128	12.1	12.8	25.1	342	87	3244	28
Test 24	6.55	912	65.2	1.08	6.2	0	61.0	6.25	84	86	107	12.6	12.8	25.1	358	72	3484	25
Test 25	5.65	908	65.6	1.07	5.6	1	62.6	6.57	81	84	105	13.1	13.6	26.7	333	76	3114	31
Test 26	5.40	894	65.5	1.05	5.2	1	63.4	7.45	80	89	111	12.1	13.4	26.3	333	82	2775	35
Test 27	4.81	867	66.2	1.04	4.6	1	64.1	8.57	90	109	136	11.1	13.4	26.3	362	96	2548	35
Test 28	4.22	842	65.2	1.01	5.0	1	62.6	9.58	102	134	168	9.68	12.7	24.9	306	96	2346	35
Test 29	3.89	813	65.3	0.98	5.9	1	63.7	10.7	128	187	233	8.50	12.4	24.3	285	101	2115	36
Test 30	3.42	797	65.0	0.97	6.9	1	62.9	11.5	138	218	273	7.41	11.7	23	227	88	1928	38
Test 31	3.45	792	64.9	0.96	8.1	1.5	63.3	11.4	151	236	295	7.71	12.1	23.7	257	101	1992	37
Test 32	4.39	816	63.2	0.97	8.4	1.5	63.2	9.16	163	207	258	10.2	12.9	25.4	296	119	2406	38
Test 33	4.95	836	64.7	1.01	8.4	1.5	61.8	8.18	139	163	203	11.1	13.0	25.5	296	111	2354	44
Test 34	5.35	859	64.6	1.03	8.6	1.5	62	6.78	112	118	148	12.3	13.0	25.4	308	101	2730	41
Test 35	5.74	879	64.4	1.05	7.2	1.5	61.5	6.49	95	98	122	12.5	12.9	25.3	214	88	2831	39
Test 36	5.83	902	65.2	1.04	9.3	1.5	67.8	3.56	100	86	107	15.7	13.5	26.5	250	87	3420	38

4.3.1 Discussion of Coal B and Sorbent A tests (Ca/S=0, 1, 1.5)

4.3.1.1 Carbon in ash

A constant carbon in ash trend was observed for an increase in bed temperature for Ca/S = 0, 1 and 1.5 (Figure 4-5).

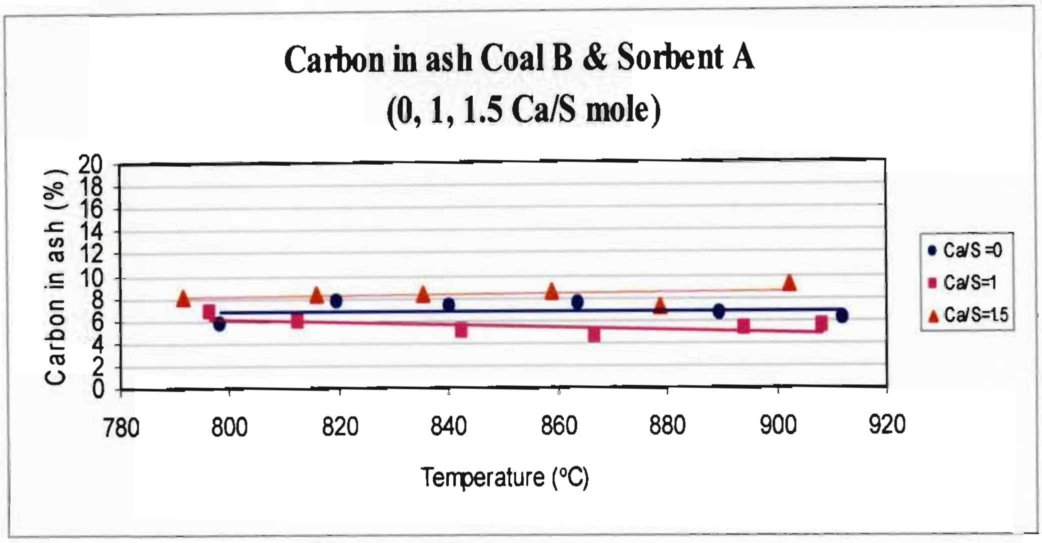

Figure 4 - 5 Carbon in Ash for Coal B tests

4.3.1.2 Nitrogen oxide emissions

The NO emissions increased with an increase in bed temperature for Ca/S = 0 and 0.5 (Figure 4-6). The NO emissions remained relatively constant for an increase in bed temperature for Ca/S = 1.5. The introduction of sorbent resulted in higher NO emissions for a bed temperature of 800 to 820°C (Figure 4-6).

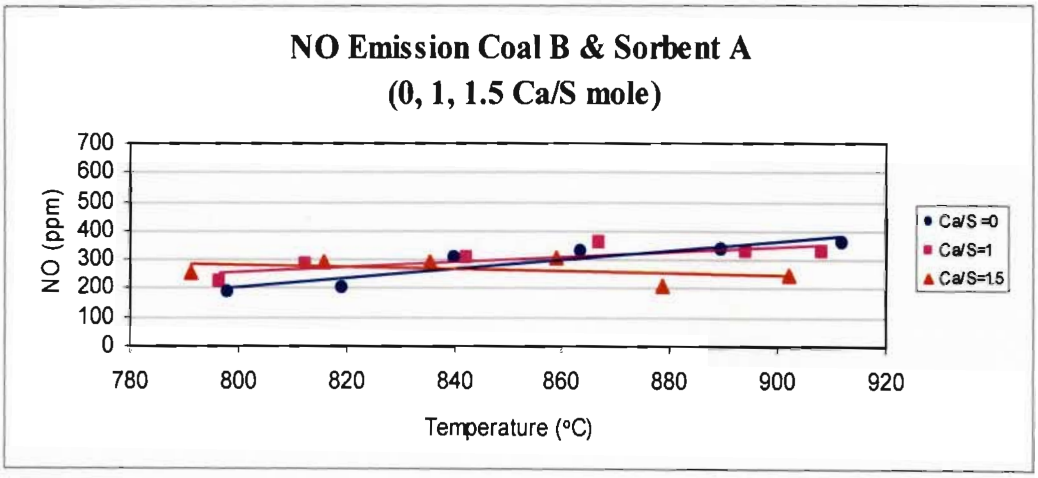

Figure 4 - 6 NO Emissions for Coal B tests

4.3.1.3 N_2O emissions

The N_2O emissions decreased with an increase in bed temperature for Ca/S = 0, 1 and 1.5 (Figure 4-7). The introduction of sorbent had little impact on N_2O emissions.

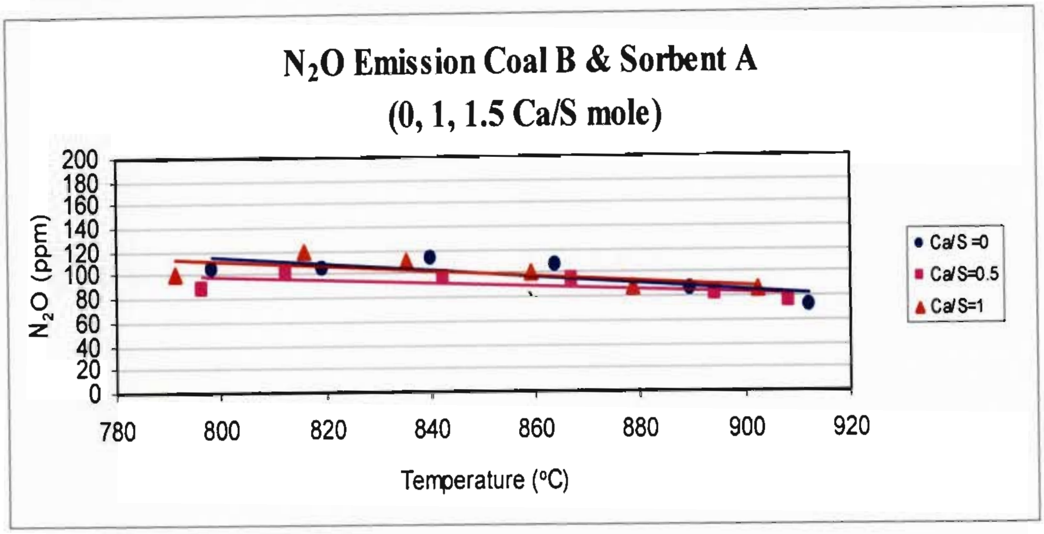

Figure 4 - 7 N_2O Emissions for Coal B tests

4.3.1.4 SO_2 Retention

A relatively constant SO_2 retention trend was observed for Ca/S = 0, 1, 1.5 (Figure 4-8). The maximum SO_2 retention was observed in the 800 to 840°C bed temperature range.

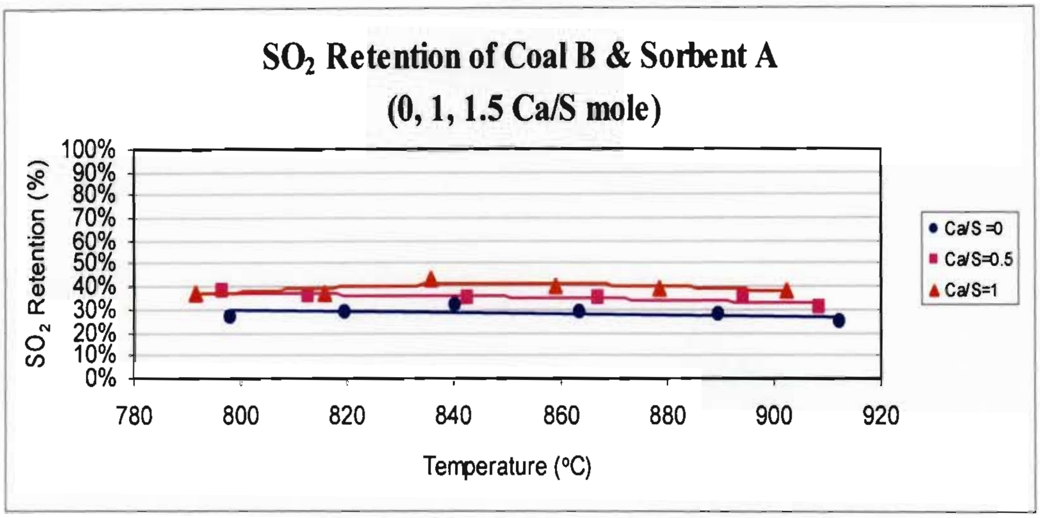

Figure 4 - 8 SO_2 Emissions for Coal B tests

4.3.1.5 Comparison of the SO_2 retention of sorbent A under bench scale tests of fluidisation with SO_2 gas only and under combustion conditions on the FBTF with Coal B

Table 4 - 4 Comparison of the SO_2 retention of sorbent A, under bench scale tests of fluidisation with SO_2 gas and the FBTF results of Coal B.

Bed temperature	**Max SO_2 retention (bench scale tests, fluidisation with SO_2 gas only Ca/S=2.85)**	**Max SO_2 retention (FBTF Ca/S=0)**	**Max SO_2 retention (FBTF Ca/S=1)**	**Max SO_2 retention (FBTF Ca/S=1.5)**
800°C	64.78%	27%	38%	37%
850-860°C	71.61%	29%	35%	41%
900°C	68.76%	25%	31%	38%

This comparison intends to examine if there any similarities in the reactivity of the sorbent under conditions of fluidised bed combustion and fluidisation with SO_2. The bench scale test data for sorbent A were extracted from Govender (2006). Section 2.7 briefly explains the method bench scale testing. A similar retention profile was obtained for the bench tests and a Ca/S = 1.5 on the FBTF.

4.4 COAL C AND SORBENT A

This section presents the results of Coal C and Sorbent A tests (Table 4-5). During Tests 37 to 42 no sorbent was introduced into the bed which enabled inherent sulphur capture to be evaluated. For Tests 43 to 48, 49 to 54 and 55 to 60 Limestone (sorbent) was added at a Ca/S mole ratio of 1.1, 1.4 and 1.7 respectively. The purpose of these tests was to investigate the sorbent capabilities of in-bed desulphurisation. The effect of bed temperature on products of combustion was evaluated for a temperature range of 800 to 900°C, in increments of 20°C.

Table 4 - 5 Results Coal C and Sorbent A

	Coal federate (kg/h)	Ave bed temperature (°C)	Total air (kg/h)	Fluidising air velocity (m/s)	Carbon cyclone ash (%)	Ca/S molar ratio	Operating pressure (kPa)	O_2 (%)	CO (ppm)	CO (ppm @6% O_2)	CO (mg/Nm^3 @6% O_2)	CO_2 (%)	CO_2 (% @6% O_2)	CO_2 (mg/Nm^3 @6% O_2)	NO (ppm)	SO_2 (ppm)	SO_2 retention (%)
Test 37	6.60	802	87.1	0.97	12.2	0	52	11.69	201	324	405	8.22	13.3	16.6	208	542	55
Test 38	6.90	826	87.2	1.00	11.2	0	52	10.92	163	243	304	8.88	13.2	16.5	209	562	56
Test 39	7.11	842	87.3	1.02	9.6	0	52.4	10.63	152	219	274	9.16	13.3	16.6	217	516	60
Test 40	7.20	862	87.3	1.03	8.2	0	53.5	10.30	134	188	235	9.47	13.3	16.6	236	530	60
Test 41	7.25	876	87.3	1.04	8.6	0	52.5	9.97	114	156	195	9.75	13.3	16.6	258	554	58
Test 42	7.20	903	86.9	1.07	7.0	0	50.8	9.46	100	130	163	10.21	13.3	16.6	285	604	54
Test 43	6.30	800	85.8	0.98	11.1	1.1	48.3	11.89	133	219	274	8.10	13.4	16.7	178	162	86
Test 44	6.10	822	91.7	1.04	9.1	1.1	53.9	11.51	108	171	213	8.36	13.2	16.6	219	122	89
Test 45	6.20	841	91.2	1.05	8.1	1.1	54	10.97	94	141	177	8.88	13.3	16.6	234	132	88
Test 46	6.20	858	89.9	1.05	7.4	1.1	55.3	10.76	80	117	147	8.90	13.1	16.3	239	136	88
Test 47	6.41	879	88.3	1.10	6.6	1.1	48	10.52	78	112	140	9.26	13.3	16.6	298	146	88
Test 48	7.10	904	85.7	1.05	7.1	1.1	52.9	9.38	74	96	120	10.27	13.3	16.6	301	245	81
Test 49	6.40	796	87.0	0.98	11.5	1.4	50.5	11.86	142	234	292	8.05	13.2	16.6	156	201	83
Test 50	6.10	814	90.7	1.02	9.6	1.4	53.8	11.58	116	184	230	8.28	13.2	16.5	205	117	89
Test 51	6.30	840	91.2	1.05	8.2	1.4	54.1	10.93	98	146	183	8.86	13.2	16.5	237	124	89
Test 52	6.30	856	87	1.04	6.9	1.4	50.9	10.44	88	125	156	9.29	13.2	16.5	227	115	90
Test 53	6.50	881	88.3	1.09	6.4	1.4	48.3	10.35	82	116	145	9.54	13.5	16.8	304	145	88
Test 54	7.02	906	86.1	1.01	6.5	1.4	53.7	9.28	83	107	133	10.40	13.3	16.7	320	176	87
Test 55	6.60	795	86.9	0.98	11.9	1.7	49.8	11.64	146	234	293	8.23	13.2	16.5	149	220	82
Test 56	6.14	816	87.5	0.99	10.2	1.7	52.5	10.81	119	175	219	8.98	13.2	16.6	221	113	90
Test 57	6.40	839	91.9	1.06	8.6	1.7	54.3	10.85	99	147	184	8.95	13.2	16.6	237	107	90
Test 58	6.50	861	87.1	1.04	6.6	1.7	50.9	10.26	83	115	144	9.39	13.1	16.4	236	100	92
Test 59	6.60	882	89.2	1.05	6.2	1.7	57.3	10.22	81	112	140	9.45	13.2	16.5	294	127	89
Test 60	6.90	906	86.3	1.04	6.2	1.7	54.8	9.45	73	95	118	10.13	13.2	16.5	321	137	89

4.4.1 Discussion of Coal C and Sorbent A tests (Ca/S=0, 1.1, 1.4, 1.7)

4.4.1.1 Carbon in ash decreased for an increase in bed temperature from 800 to 900°C for Ca/S molar ratio = 0, 1.1, 1.4 and 1.7. The addition of sorbent resulted in a decrease in carbon in ash (Figure 4-9).

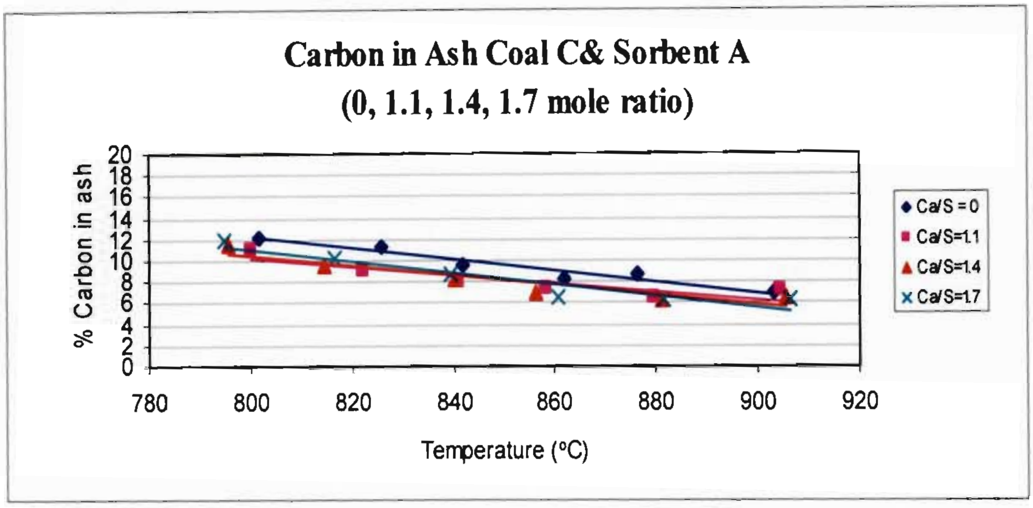

Figure 4 - 9 Carbon in ash for Coal C tests

4.4.1.2 Nitrogen oxide emissions

Nitrogen oxide emissions increased with an increase in temperature for Ca/S mole ratios of 0, 1.1, 1.4, 1.7 (Figure 4-9). An increase in Ca/S mole ratio showed a slight increase in emissions of NO for a bed temperature of 840 to 900°C.

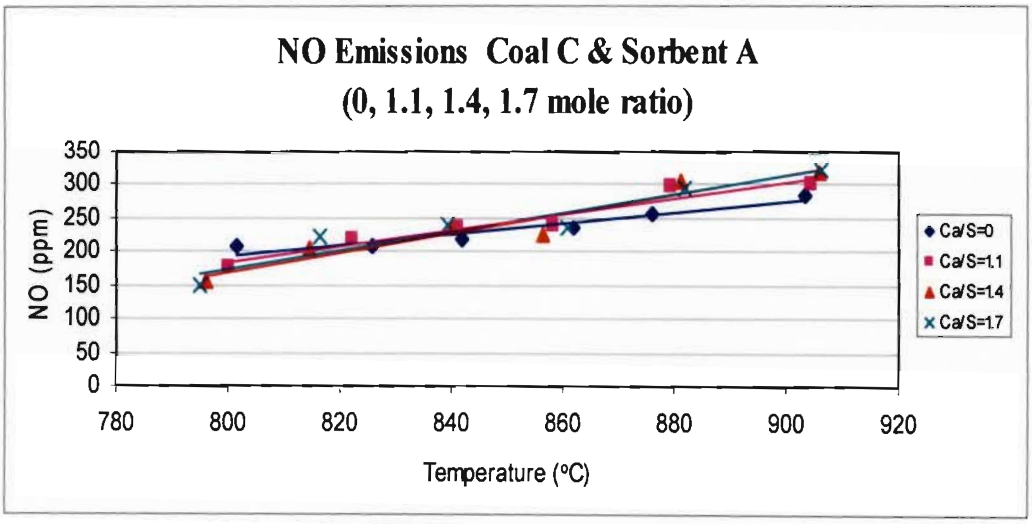

Figure 4 - 10 NO Emissions for Coal C tests

4.4.1.3 SO_2 retention

A similar SO_2 retention profile was observed for an increase in bed temperature for Ca/S 0, 1.1, 1.4 and 1.7, where the maximum retention occurred in the 820 to 860°C bed temperature range (Figure 4-11). The increase in Ca/S from 1.1 to 1.7 results in little improvement in SO_2 retention.

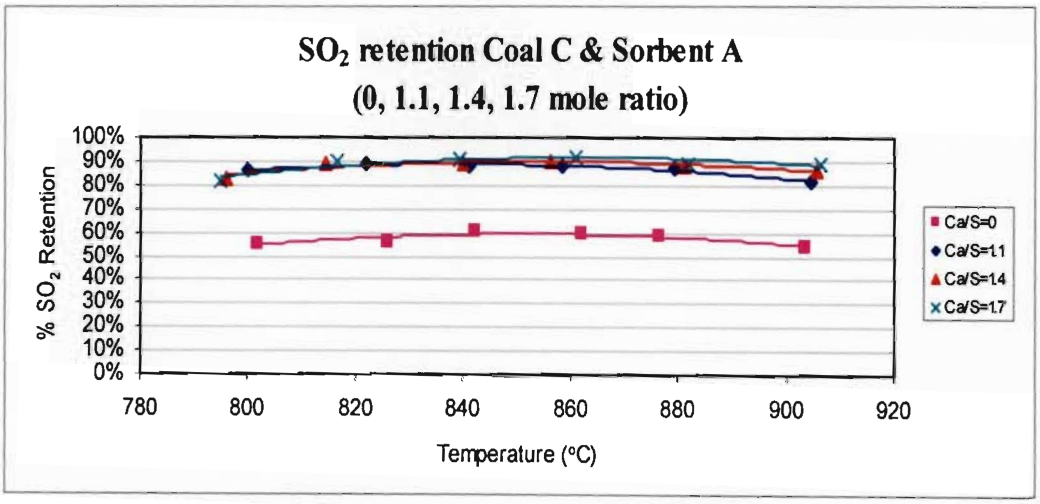

Figure 4 - 11 SO_2 retention for Coal C tests

4.4.1.4 Comparison of the SO_2 retention of sorbent A under bench scale tests of fluidisation with SO_2 gas only and under combustion conditions on the FBTF with Coal C

Table 4 - 6 Comparison of the SO_2 retention of sorbent A, under bench scale tests of fluidisation with SO_2 gas and the FBTF results of Coal C.

Bed temperature	**Maximum SO_2 retention (bench scale tests, fluidisation with SO_2 gas only Ca/S=2.85)**	**Maximum SO_2 Retention (FBTF Ca/S=0)**	**Maximum SO_2 Retention (FBTF Ca/S=1.1)**	**Maximum SO_2 Retention (FBTF Ca/S=1.4)**	**Maximum SO_2 Retention (FBTF Ca/S=1.7)**
800°C	64.78%	55%	86%	83%	82%
850-860°C	71.61%	60%	88%	90%	92%
900°C	68.76%	54%	81%	87%	89%

This comparison intends to examine if there any similarities in the reactivity of the sorbent under conditions of fluidised bed combustion and fluidisation with SO_2 gas only. The bench scale test data for sorbent A were extracted from Govender (2006). Section 2.7 briefly explains the method bench scale testing. There were similarity in the SO_2 trends of the fluidised bed combustion tests of Coal C with Sorbent A (Ca/S= 1.4, 1.7) and the bench scale tests with SO_2 gas (Table 4-6). In both experiments the peak retention observed was in the 850 to 860°C bed temperature range.

4.5 AN OVERALL COMPARISON OF COAL A, B AND C WITH SORBENT A AT Ca/S=1

The results for Coal A, B and C at various Ca/S molar ratios has been discussed in sections 4.2, 4.3 and 4.4 respectively. In-order to compare the combustion performance of these three coals with Sorbent A, a comparison must be done on the same Ca/S molar ratio. A common Ca/S mole ratio of 1 was tested for the three coals. Table 4-7 details the data extracted for the three coals at this mole ratio.

Table 4 - 7 Overall comparison of Coal A, B and C with Sorbent A at Ca/S=1

	Carbon in cyclone ash	NO (ppm)	N_2O (ppm)	SO_2 (ppm)	SO_2 retention (%)
Coal A	2.8-2.3	196-216	76-45	139-538	79% @ 800-820°C
Coal B	6.9-4.6	227-362	101-76	1928-3114	38% @797°C
Coal C	11.1-6.6	178-301	-	122-245	89% @822°C

4.5.1 Carbon in ash

Carbon in ash has shown to decrease with an increase in bed temperature. Adanez et al, (2001) concluded the same for the coals they tested. The least quantity of remaining carbon in ash was observed during Coal A tests.

4.5.2 NO emissions

The highest NO emissions observed were for Coal B tests, where a peak NO reading of 362ppm was observed.

4.5.3 N_2O emissions

N_2O emissions have shown to decrease with an increase in temperature. This is consistent with research done by Armesto et al, (2001). No N_2O emission measurements could be obtained for Coal C tests. Higher N_2O emissions were observed for Coal B than Coal A tests, where a peak N_2O reading of 101ppm was observed.

4.5.4 SO_2 emissions

The peak SO_2 retention for Coal A and C occurred at a bed temperature of 820°C, where the retention was 79 and 89% respectively. The Coal B tests showed maximum SO_2 retention of 38% was observed at a bed temperature of 797°C. Research done by Zhangfa (2003) showed the maximum retention of SO_2 for the coals they tested normally occurs at a bed temperature of 850°C. Podolski et al, (1995) showed that for a Ca/S mole ratio of 1 to 1.5 in a pressurised FBC SO_2 retention of 90% was observed for the coals they tested. In terms of SO_2 retention Coal C performed the best (Figure 4-12).

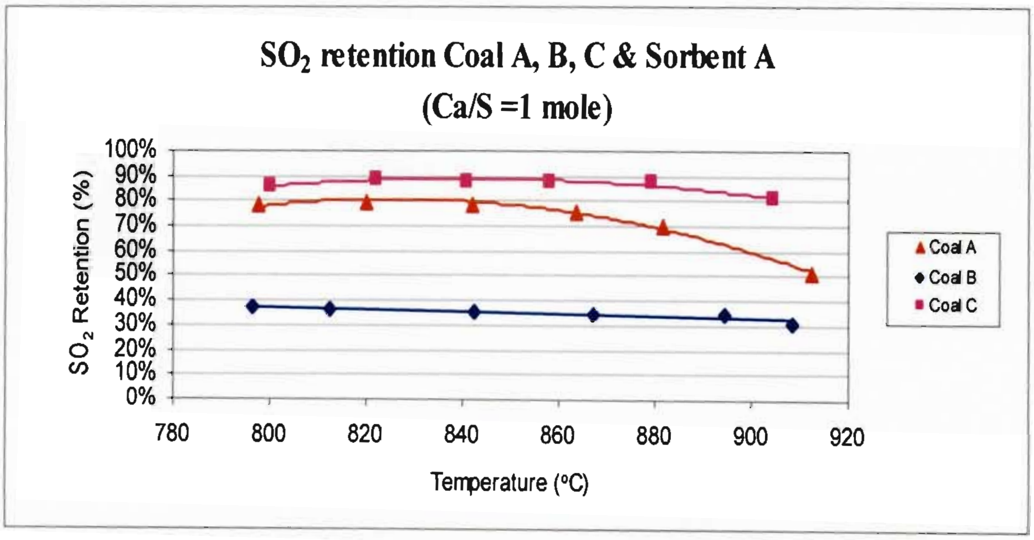

Figure 4 - 12 Comparison of SO_2 retention Coal A, B, C and Sorbent A Ca/S=1

Chapter Five

CONCLUSIONS AND RECOMMENDATIONS

5.1 CONCLUSIONS

This chapter details with the conclusions and recommendations that can be drawn from the 60 tests conducted, which comprised of three different coals being tested on the same sorbent. A common Ca/S molar ratio of 1 was tested for the three coals, therefore comparisons between these coals can only be examined at this particular ratio. The following parameters have been chosen to be discussed further carbon in ash, NO emissions, N_2O emissions, and SO_2 retention.

5.1.1 Carbon in ash

- The carbon in ash follows the expected descending trend with an increase in bed temperature for Coals A, B and C.
- The best performing coal in terms of the least quantity of remaining carbon in ash was Coal A at a Ca/S molar ratio of 1.

5.1.2 NO Emissions

- The NO emissions increased for an increase in bed temperature for Coal A, B and C.
- The greatest NO emissions were recorded during Coal B tests at a Ca/S molar ratio of 1.

5.1.3 N_2O Emissions

- The N_2O emissions decreased with an increase in bed temperature for Coal A and B tests.
- Higher N_2O emissions were observed for Coal B than Coal A tests.

5.1.3 SO_2 Retention

- The optimal operating bed temperature for SO_2 retention observed for the three coals was in the 800-860°C.
- In terms of SO_2 retention Coal C performed the best.

5.2 RECOMMENDATIONS

- If a new build fluidised bed combustor is considered for power generation in the future, the possibility of firing with Coal C should be further examined with the view point of the high sulphur retention with Sorbent A.
- Although a high carbon in ash was observed for Coal C tests, a re-circulating fluidised bed would have to be considered to reduce carbon in ash content. Fuierer and Brokelmann (1997) showed that over 99% carbon burnout is possible with re-circulation of fly ash.
- Future research on the three coals tested should examine the mechanisms behind sulphur capture. This would involve examining current two phase fluidised bed combustion models, and determining their applicability in predicting the combustion performance of these coals.

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APPENDIX A

COAL A AND SORBENT A TESTS

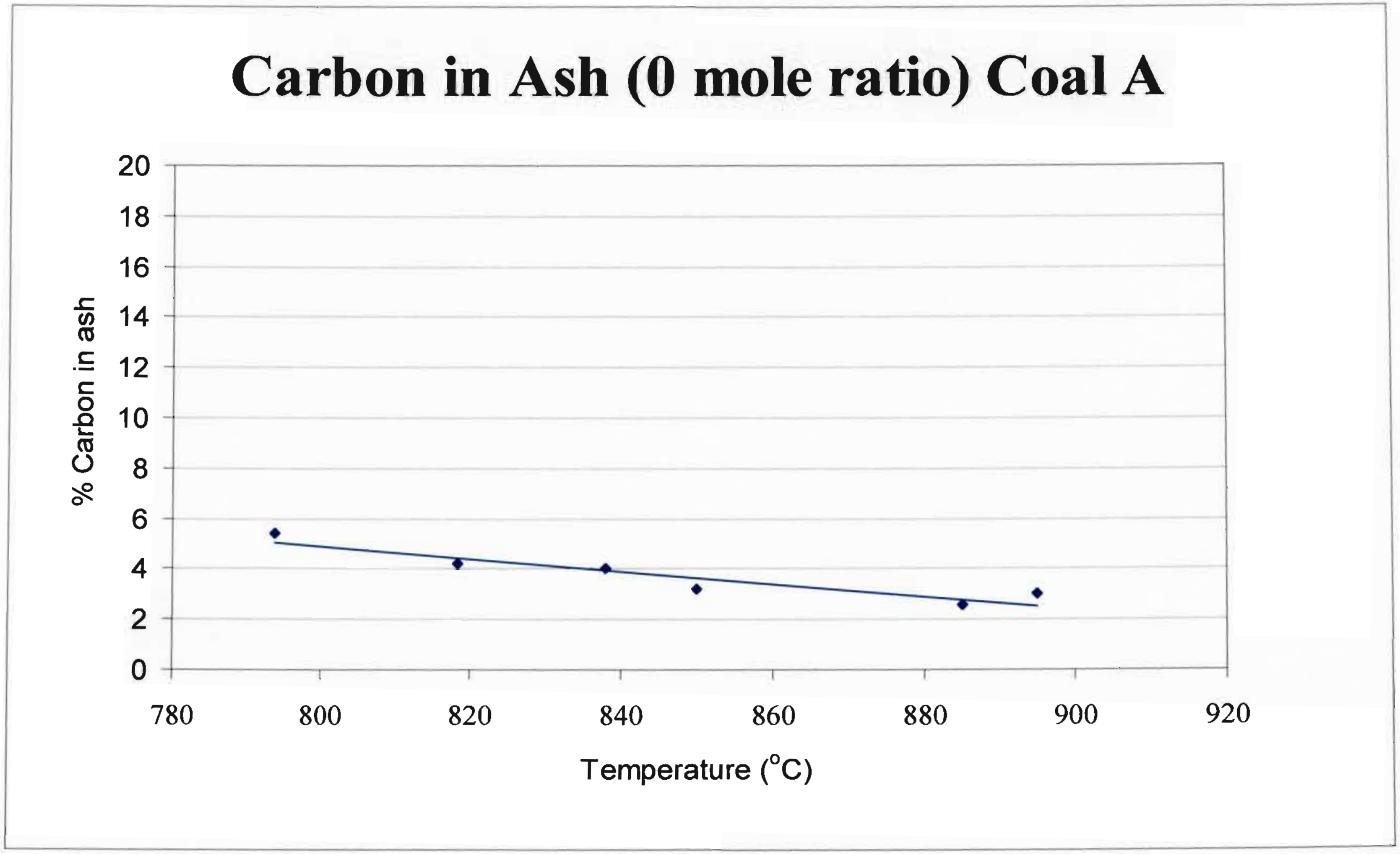


Figure A - 1 Carbon in ash for Coal A Ca/S=0

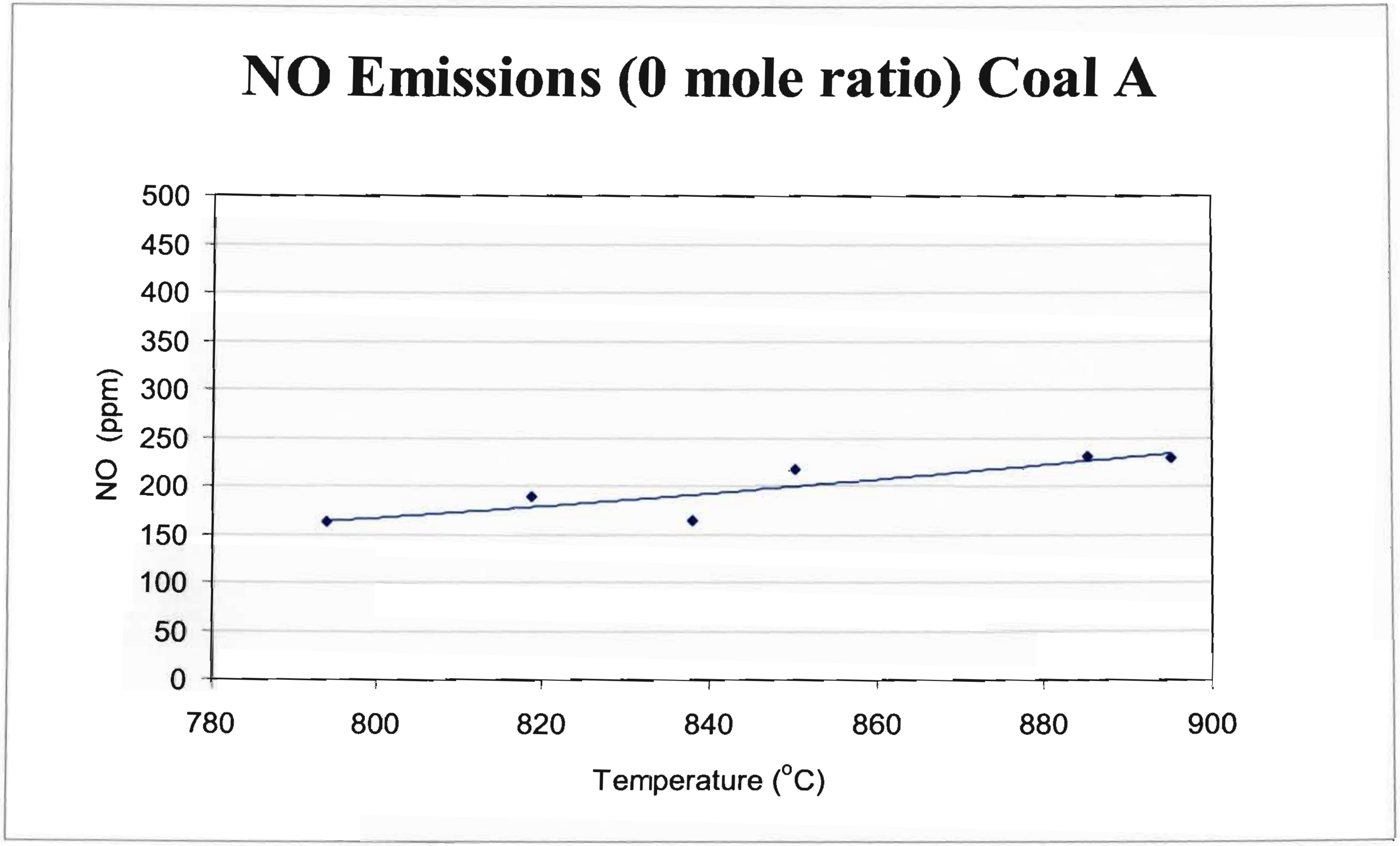


Figure A - 2 NO Emissions for Coal A Ca/S=0

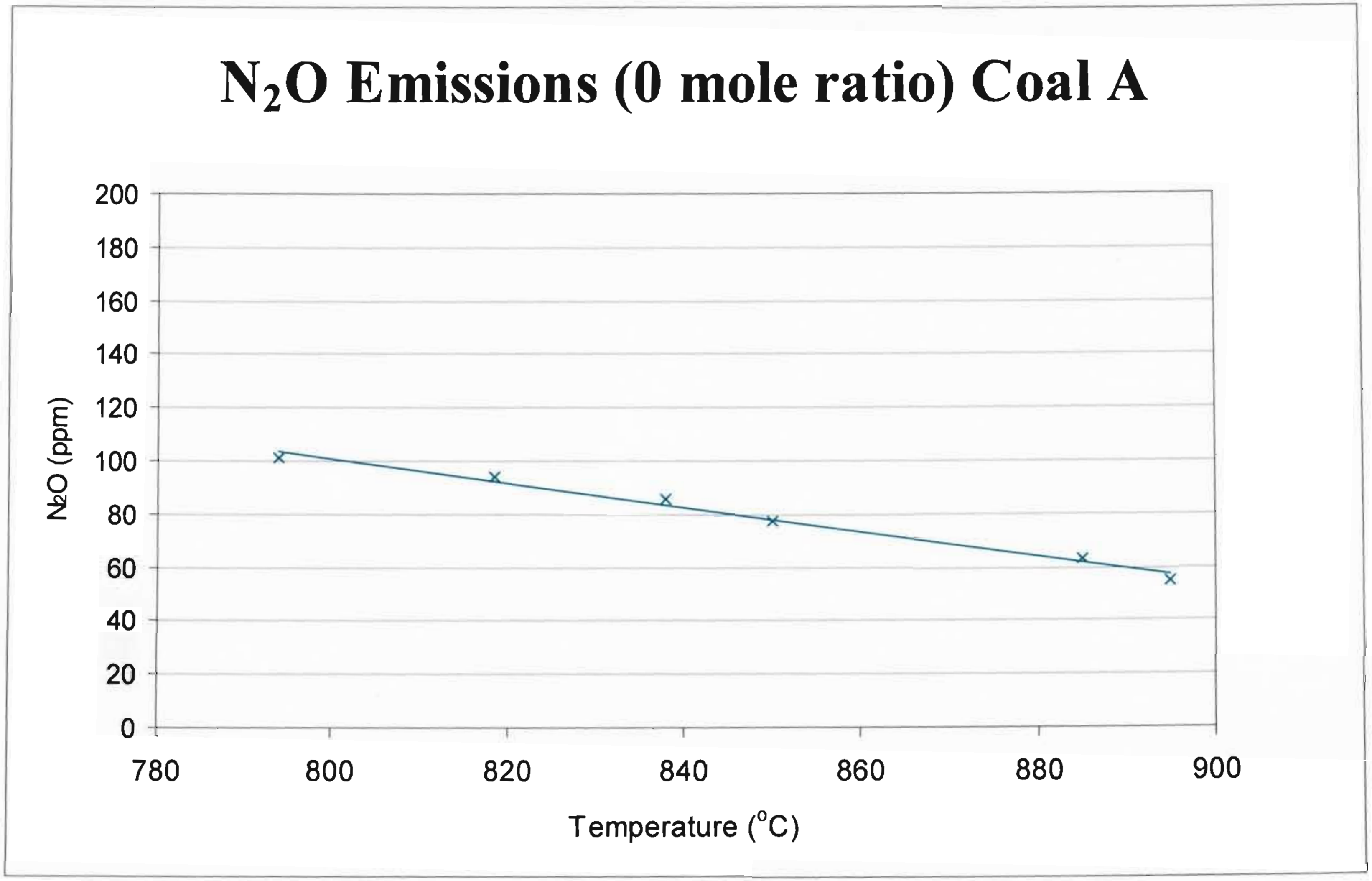


Figure A - 3 N_2O Emissions for Coal A Ca/S=0

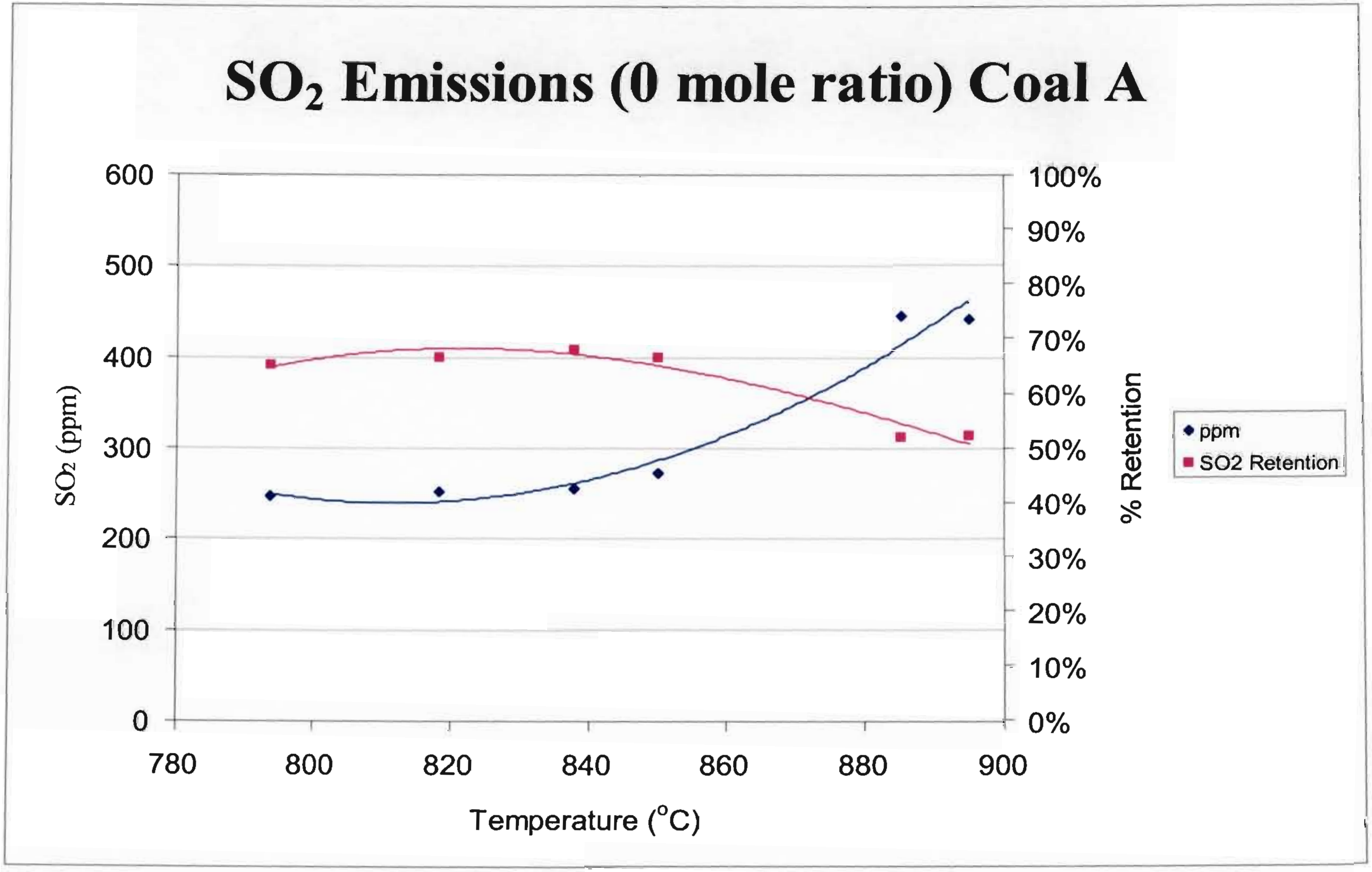


Figure A - 4 SO$_2$ Emissions for Coal A Ca/S=0

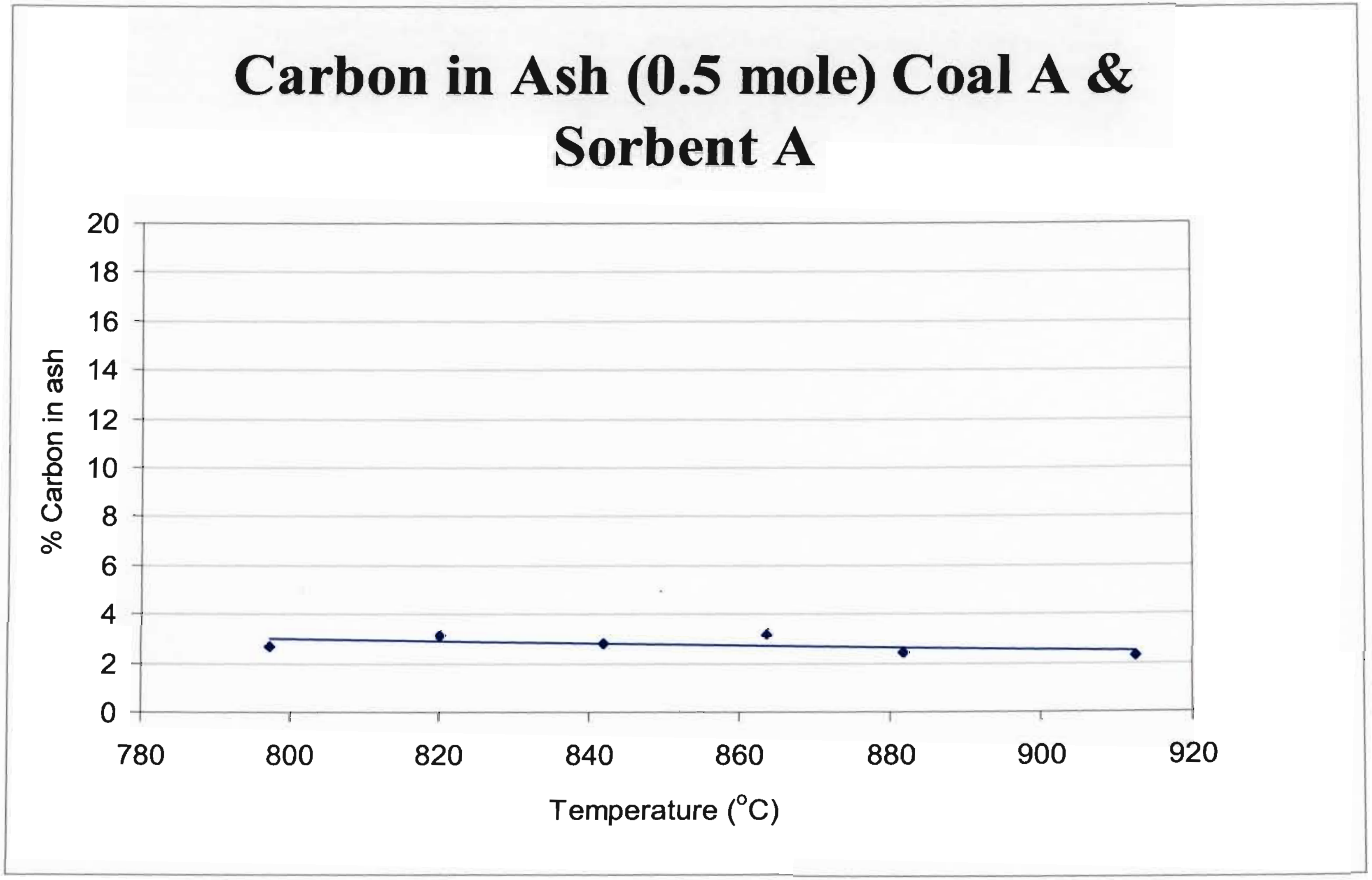


Figure A - 5 Carbon in ash for Coal A Ca/S=0.5

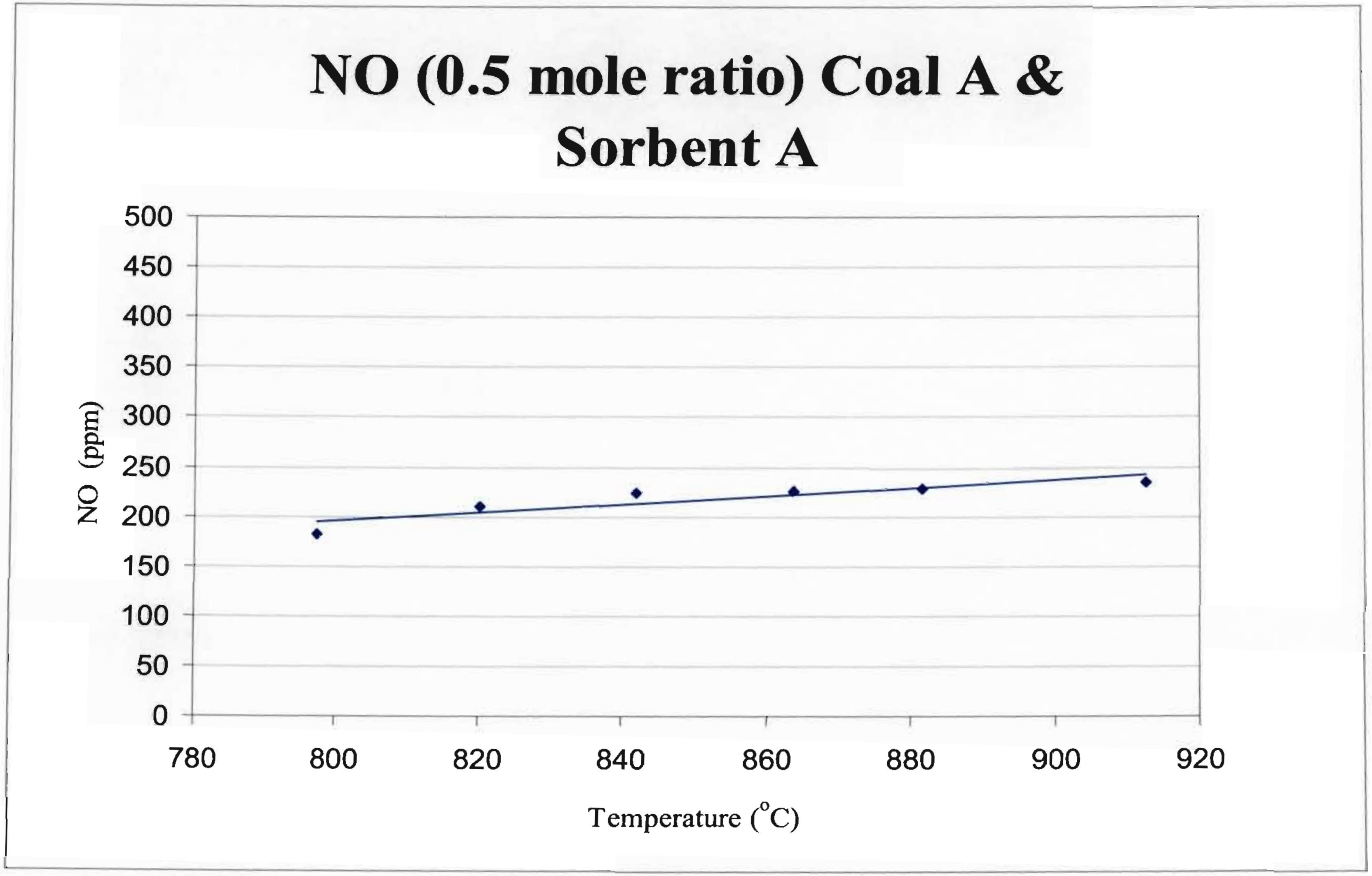


Figure A - 6 NO Emissions for Coal A Ca/S=0.5

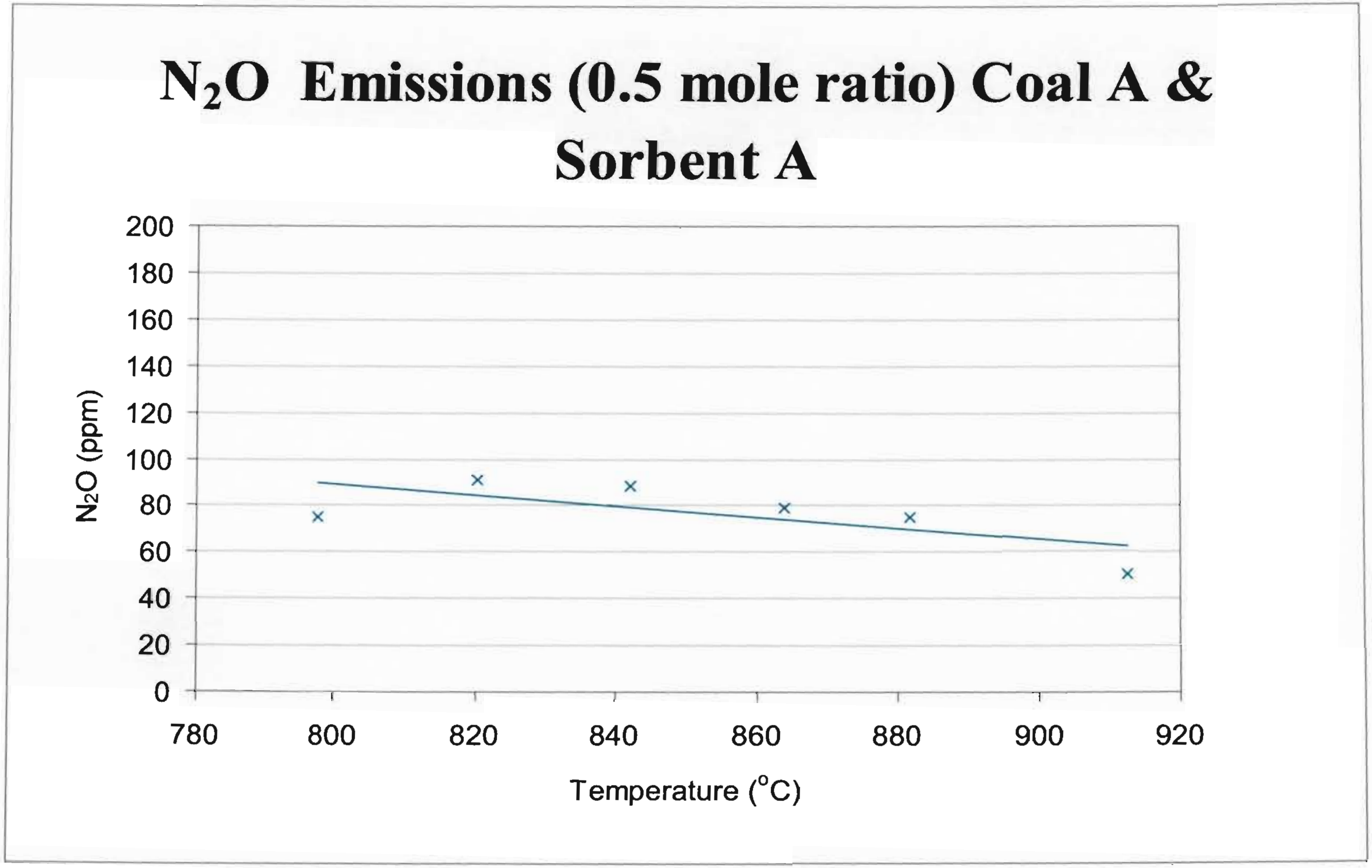


Figure A - 7 N_2O Emissions for Coal A Ca/S=0.5

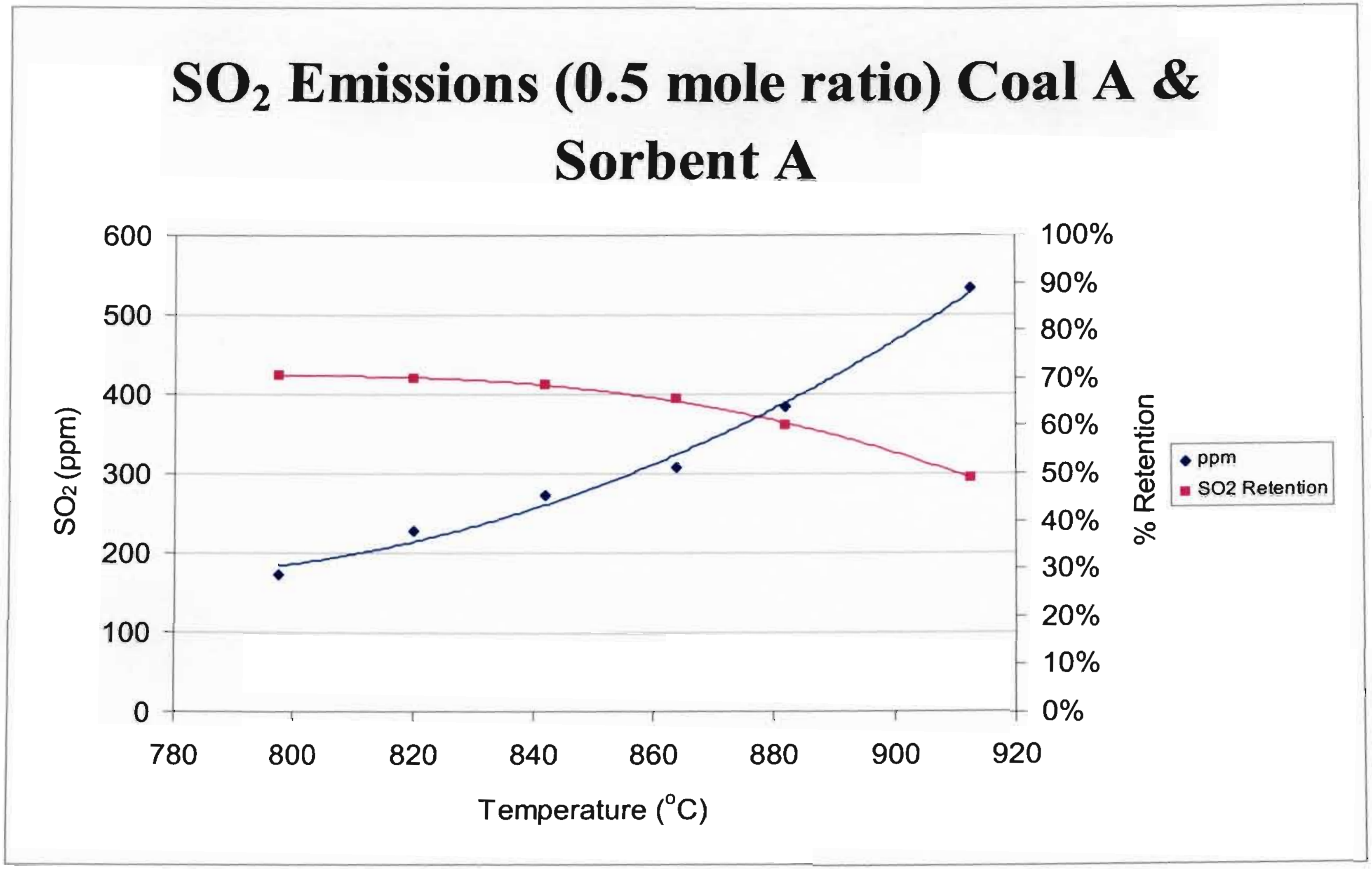


Figure A - 8 SO_2 Emissions for Coal A Ca/S=0.5

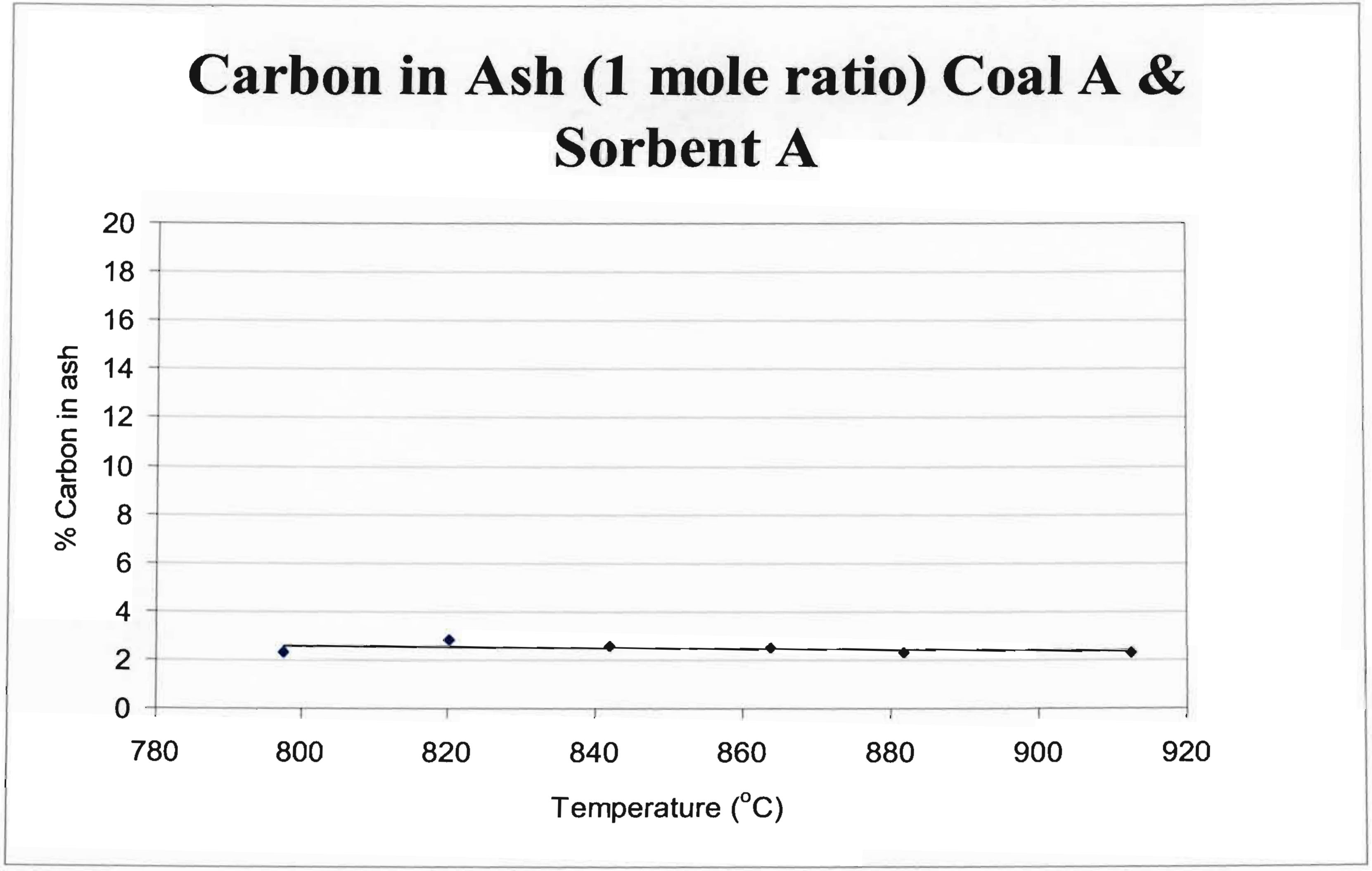


Figure A - 9 Carbon in ash for Coal A Ca/S=1

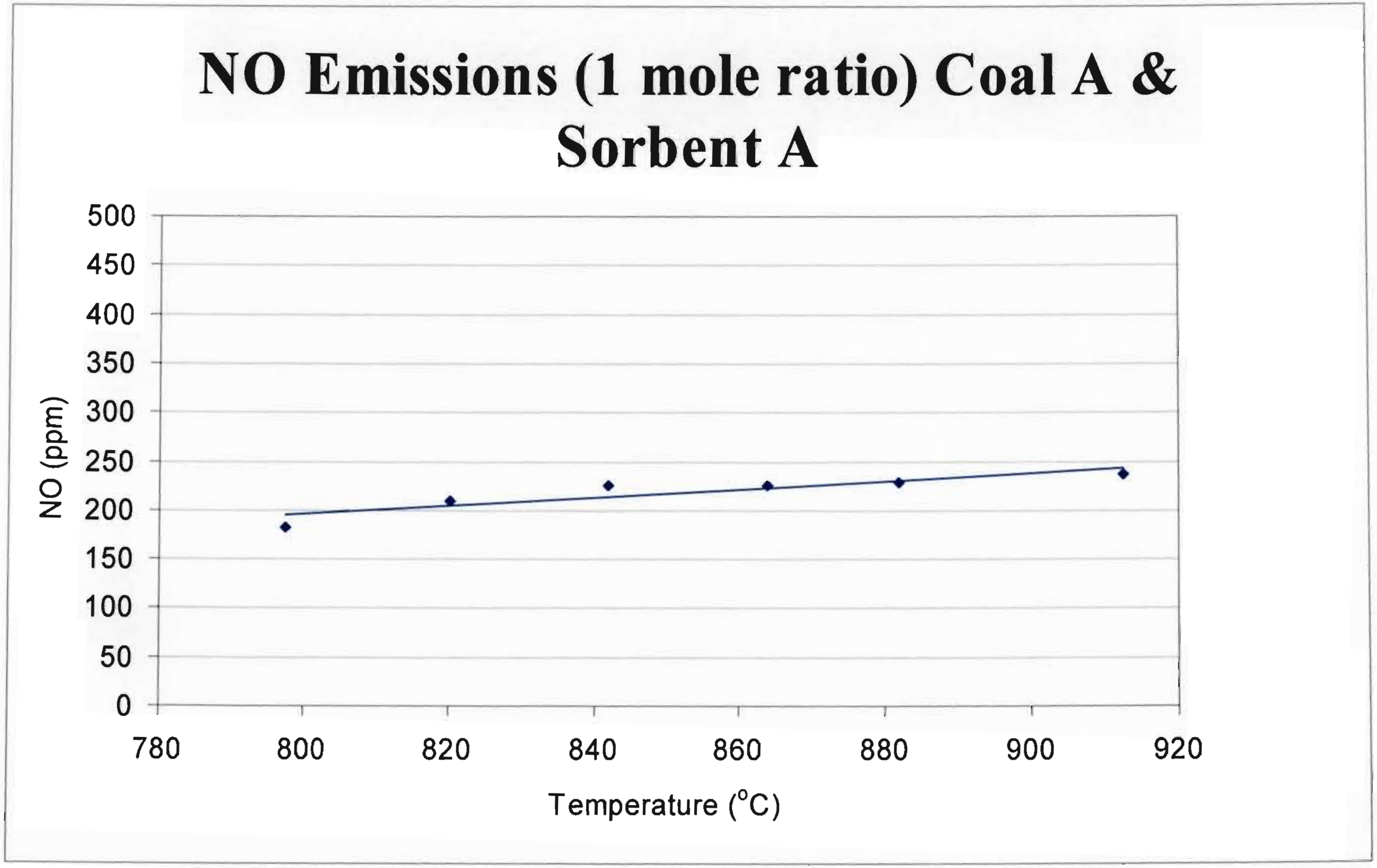


Figure A - 10 NO Emissions for Coal A Ca/S=1

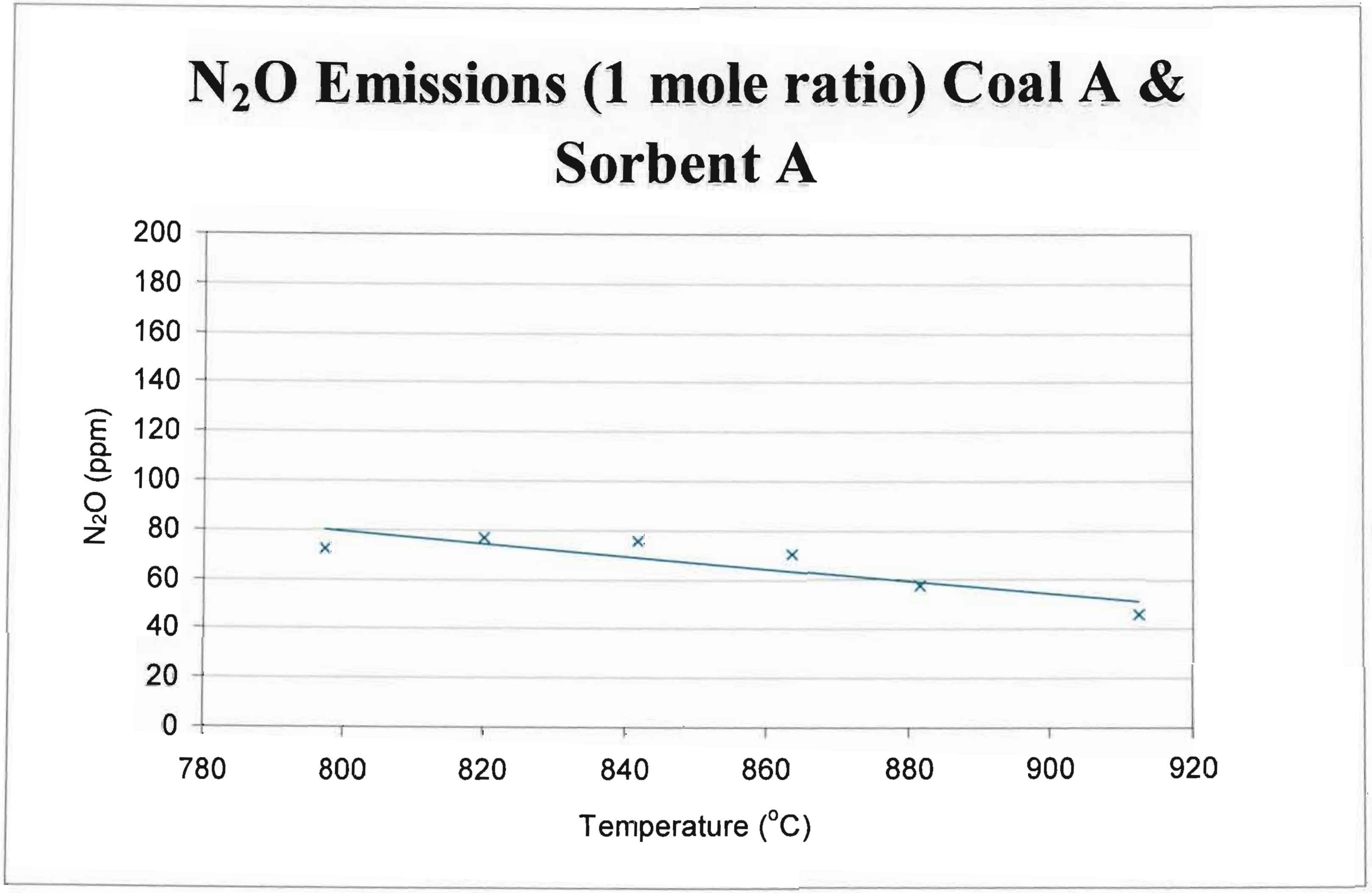


Figure A - 11 N_2O Emissions for Coal A Ca/S=1

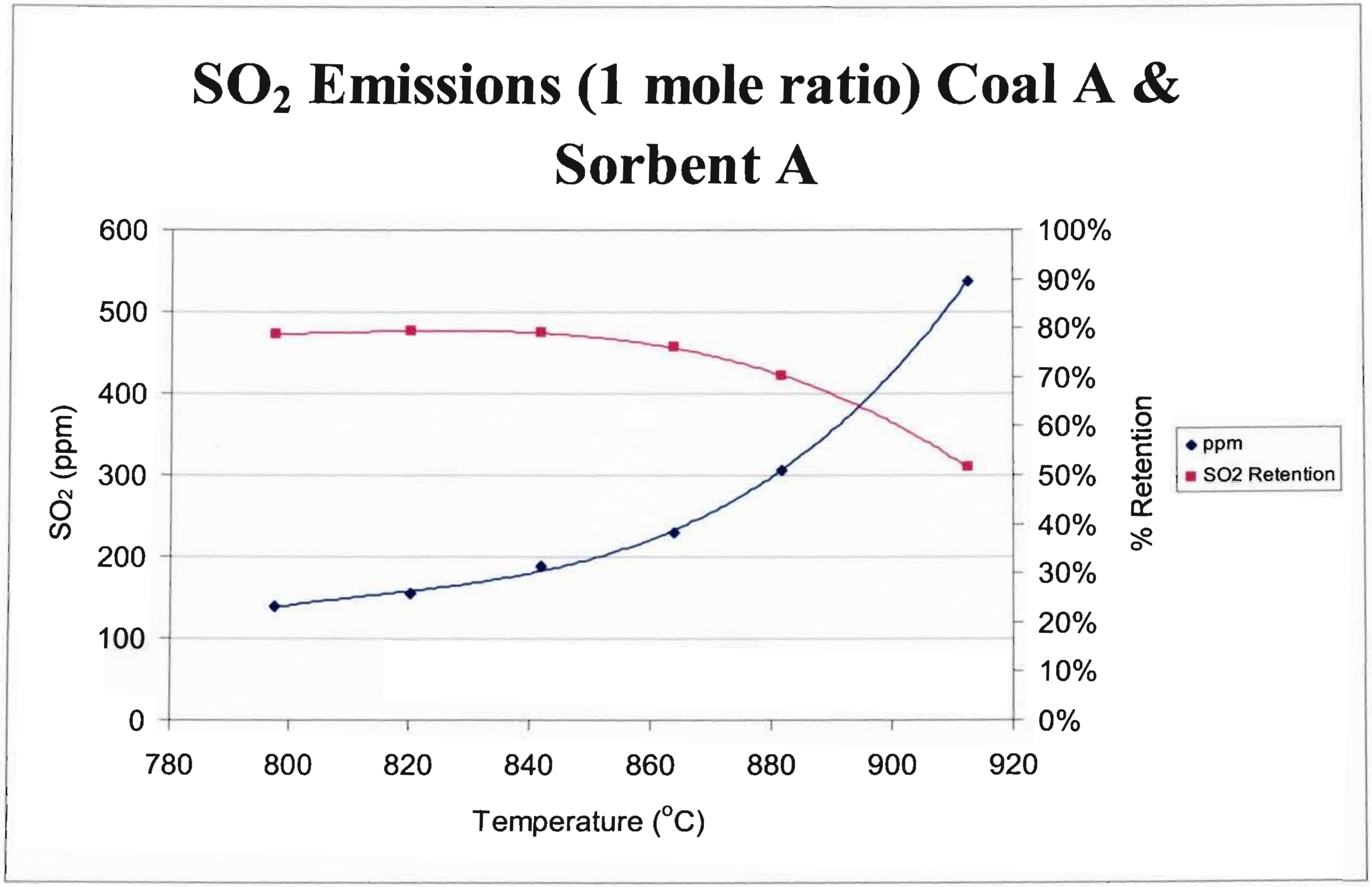


Figure A - 12 SO_2 Emissions for Coal A Ca/S=1

APPENDIX B

COAL B AND SORBENT A TESTS

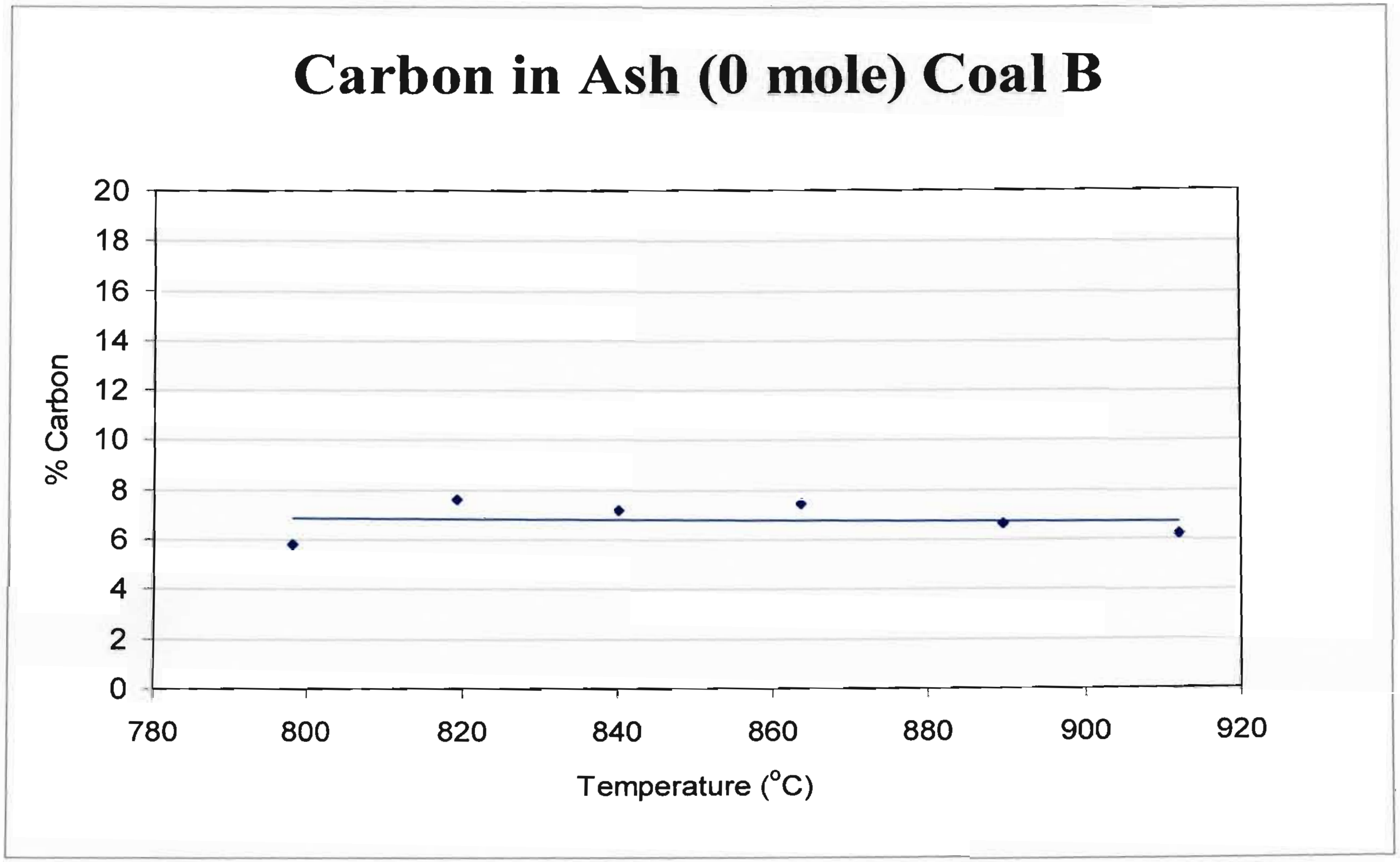


Figure B - 1 Carbon in Ash for Coal B Ca/S=0

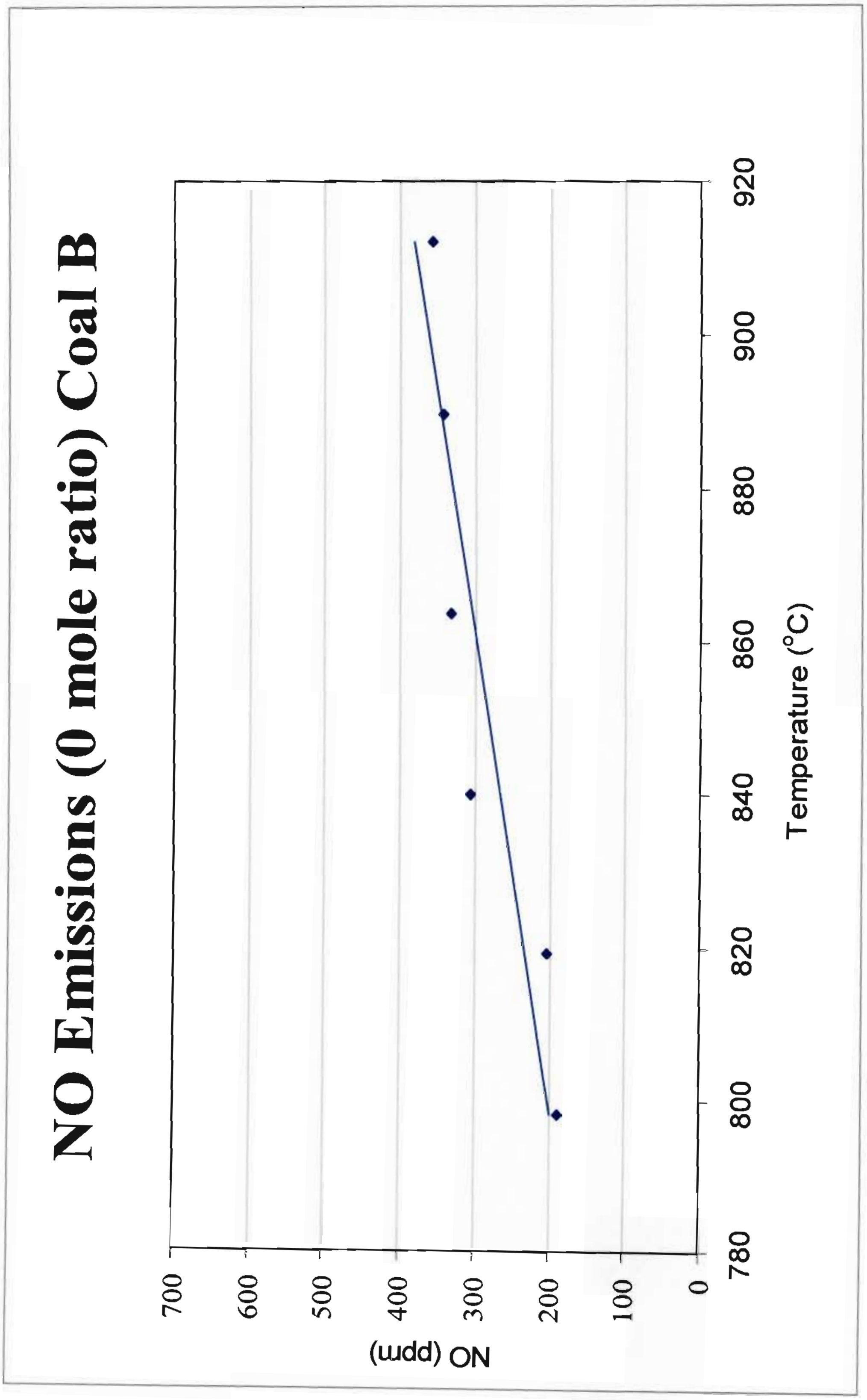


Figure B - 2 NO Emissions for Coal B Ca/S=0

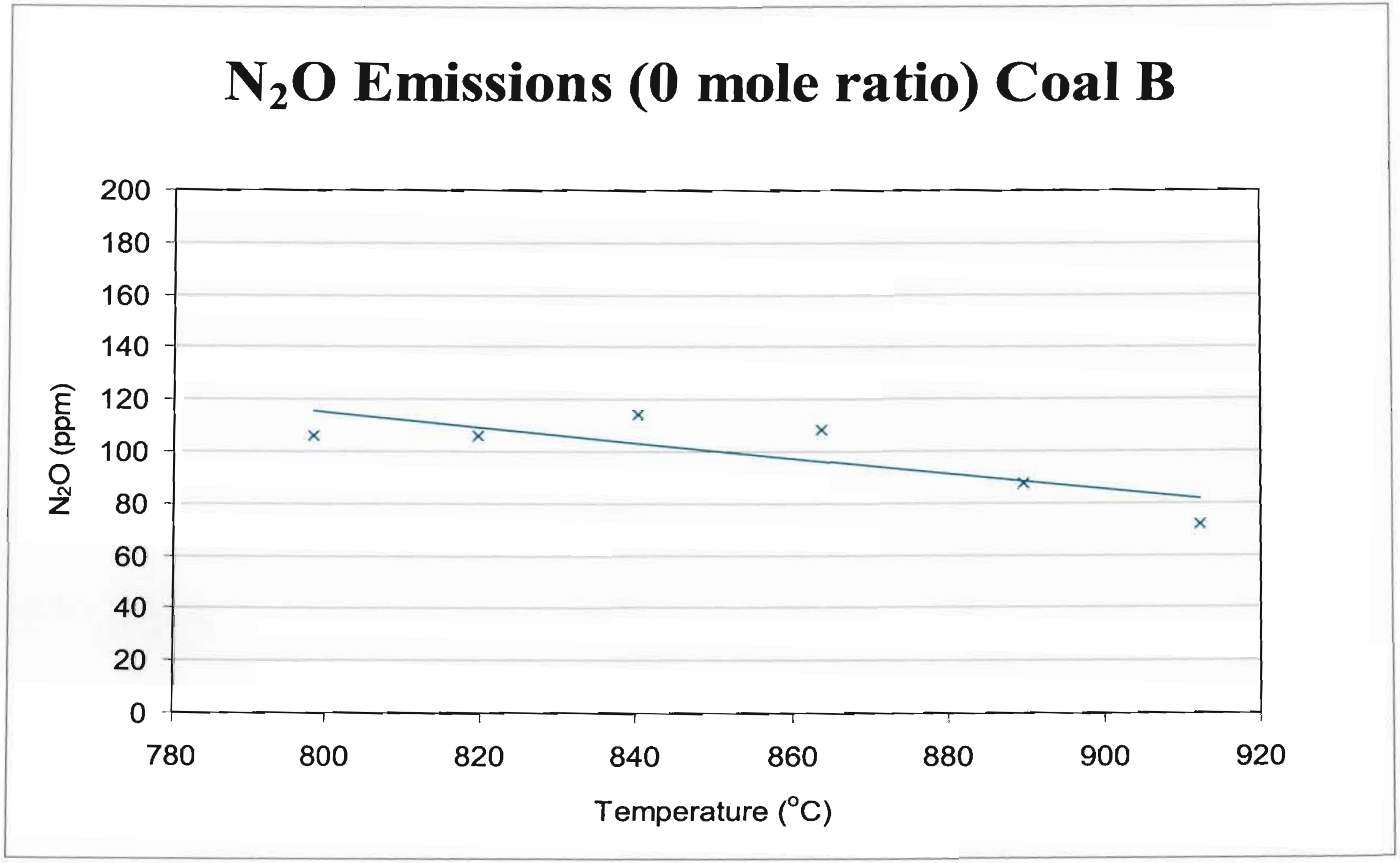
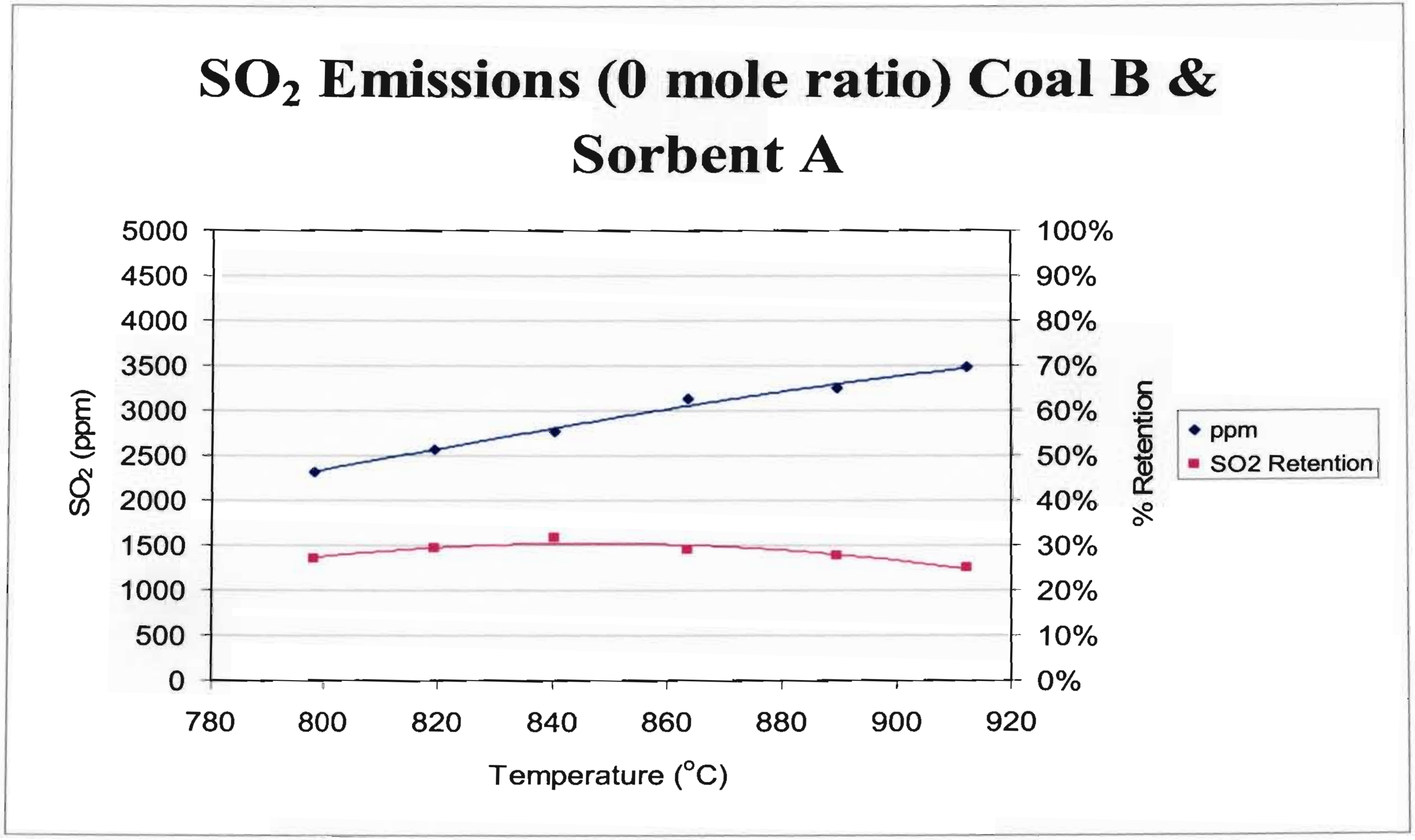


Figure B - 3 N$_2$O Emissions for Coal B Ca/S=0

Figure B - 4 SO_2 Emissions for Coal B Ca/S=0



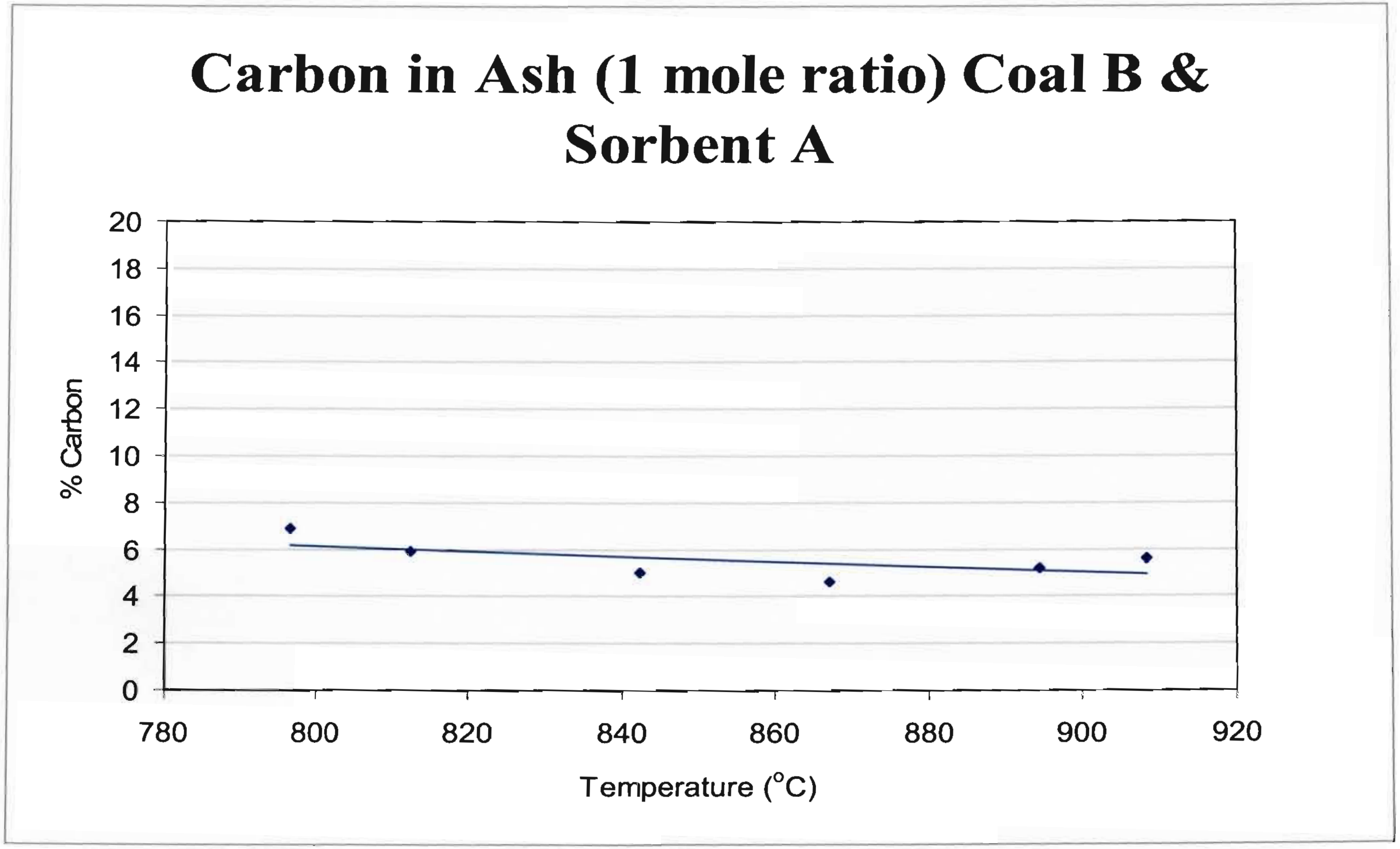


Figure B - 5 Carbon in Ash for Coal B Ca/S=1

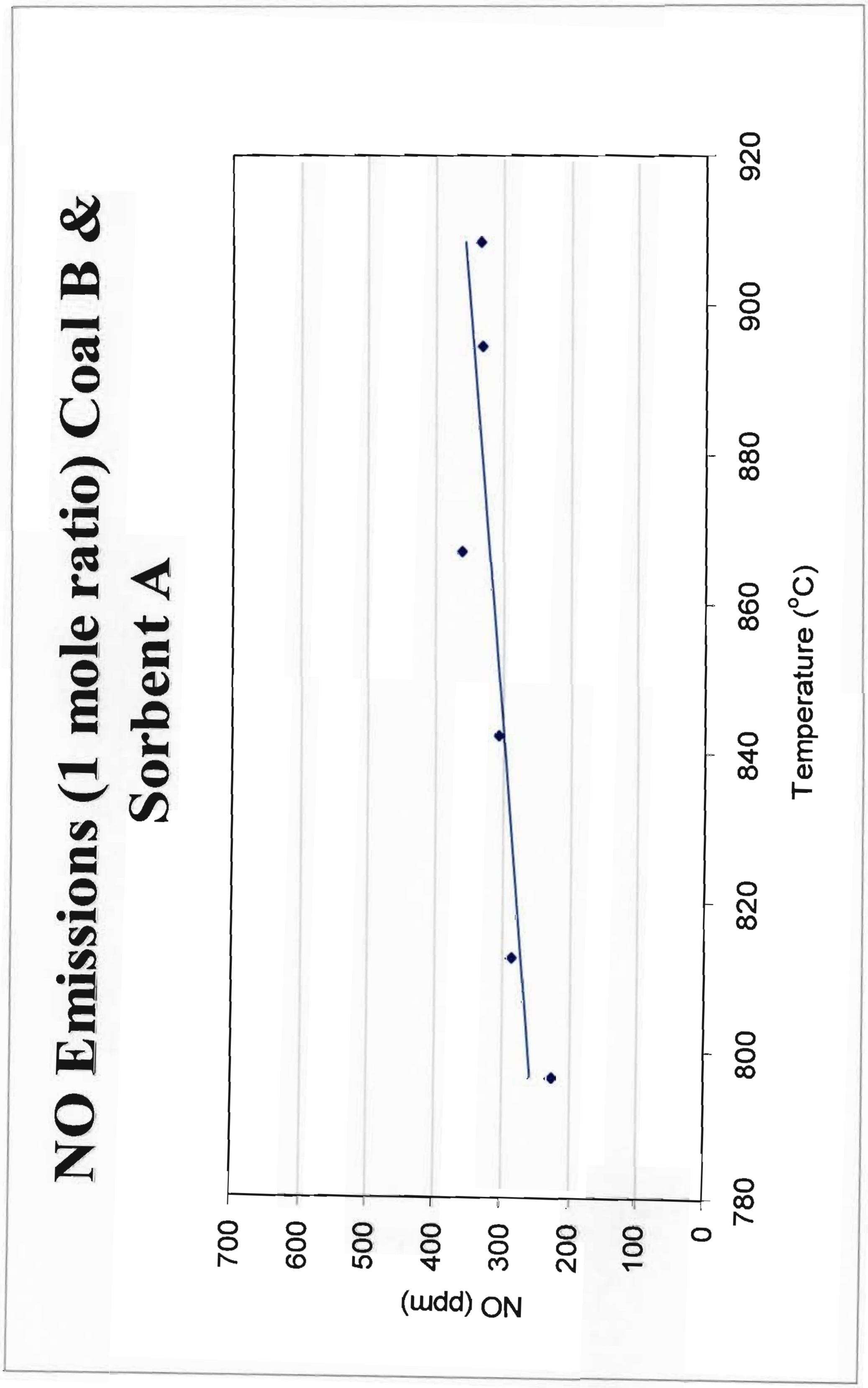


Figure B - 6 NO Emissions for Coal B Ca/S=1

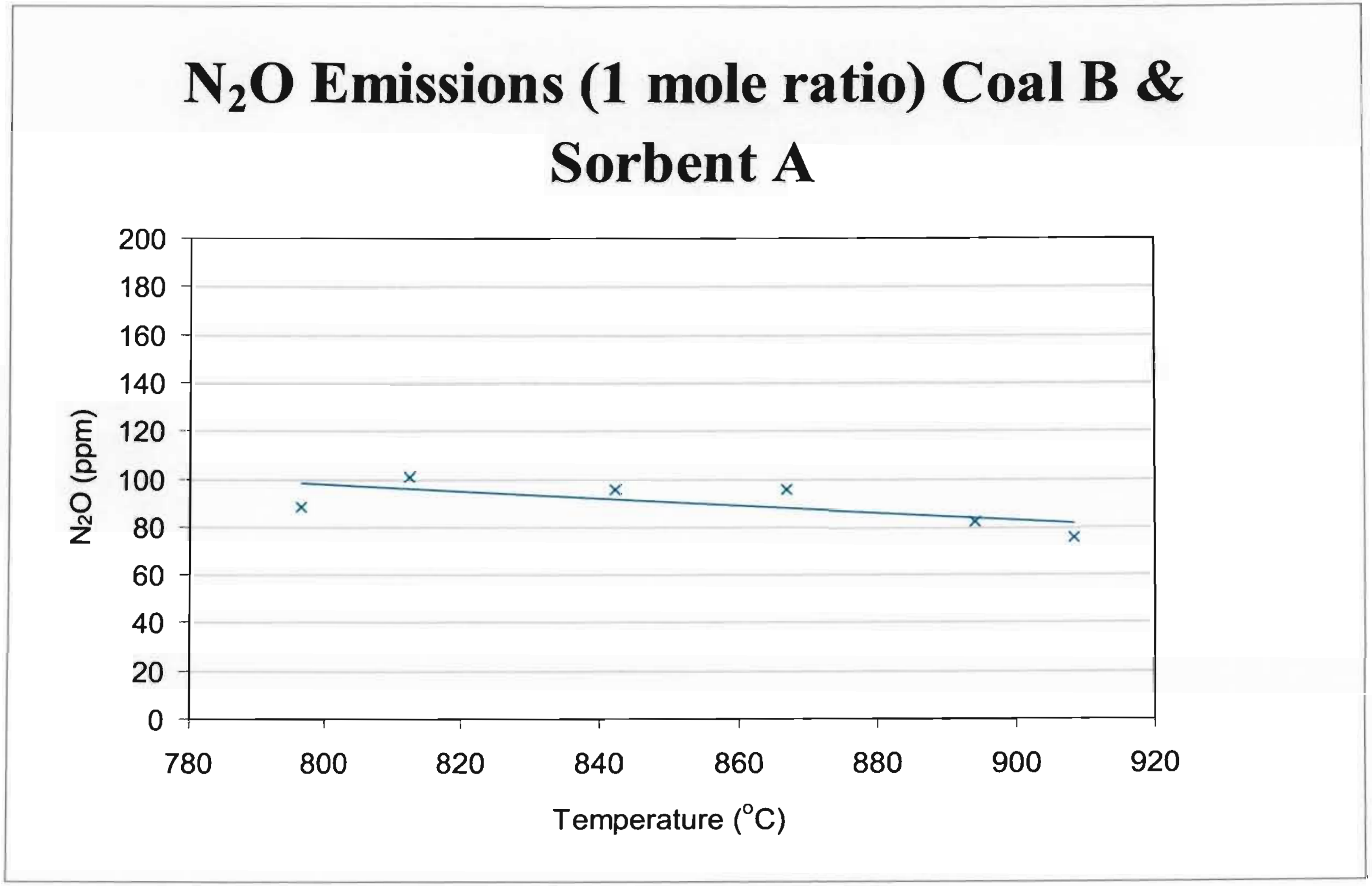


Figure B - 7 N_2O Emissions for Coal B Ca/S=1

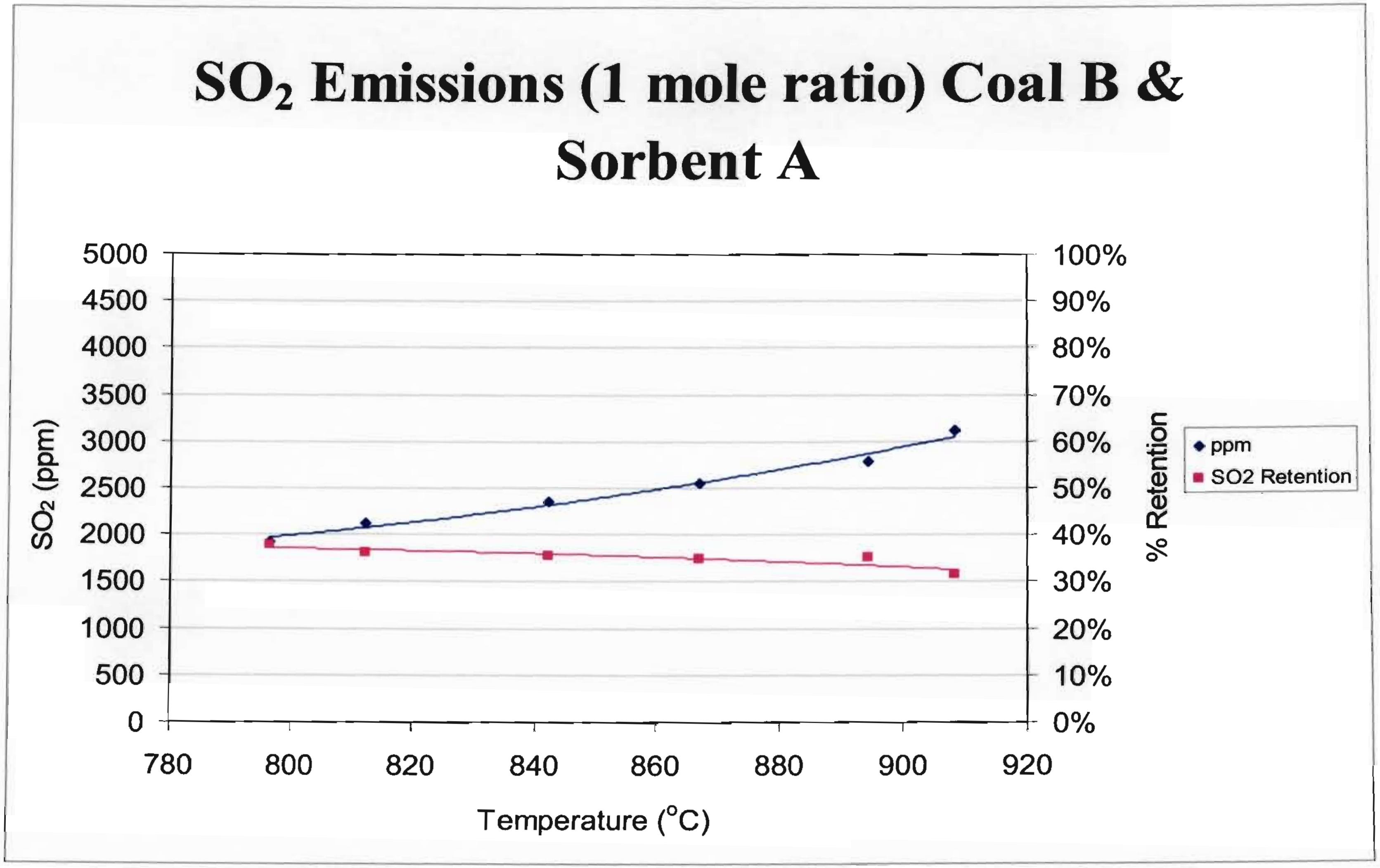


Figure B - 8 SO$_2$ Emissions for Coal B Ca/S=1

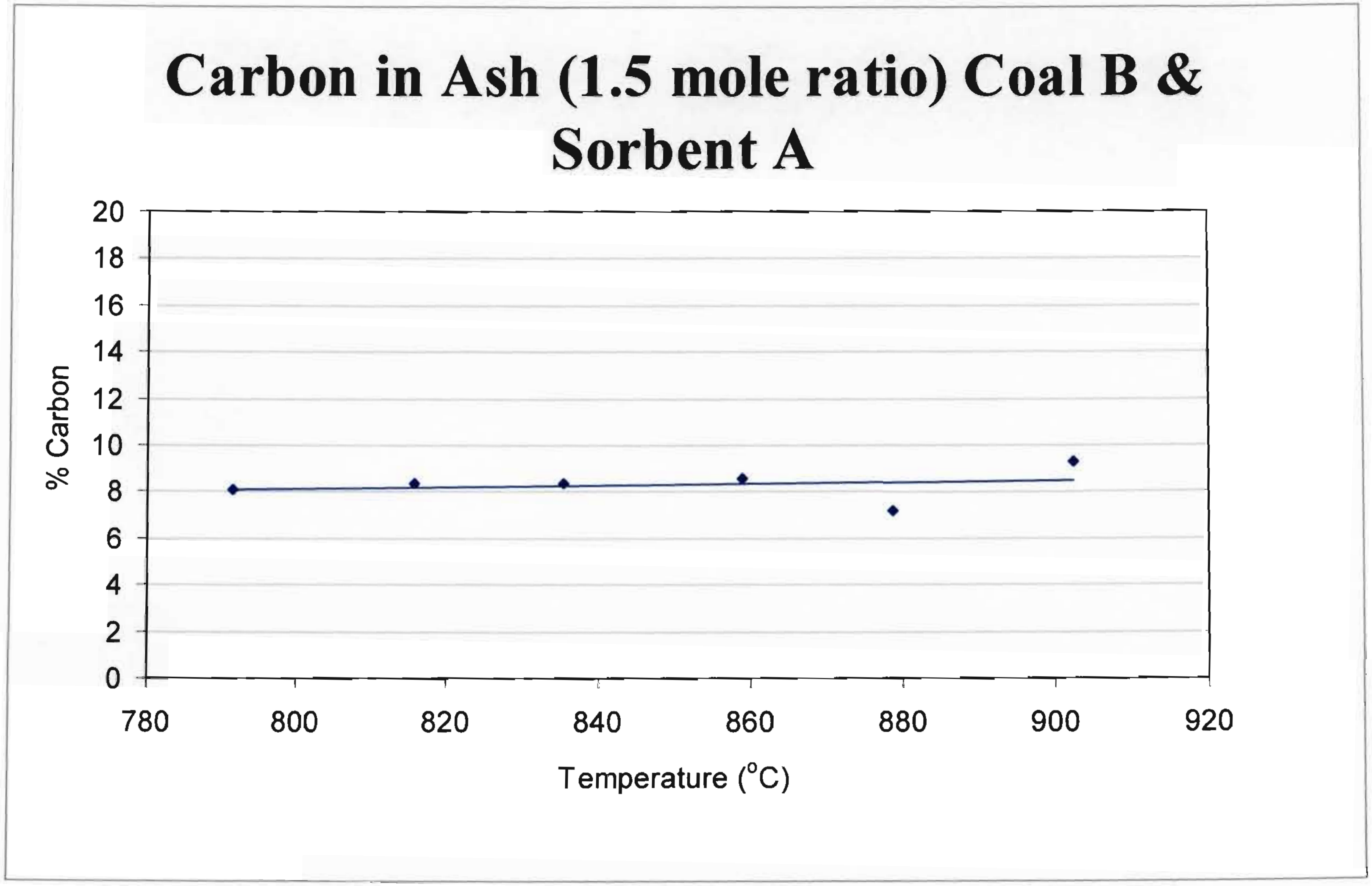


Figure B - 9 Carbon in Ash for Coal B Ca/S=1.5

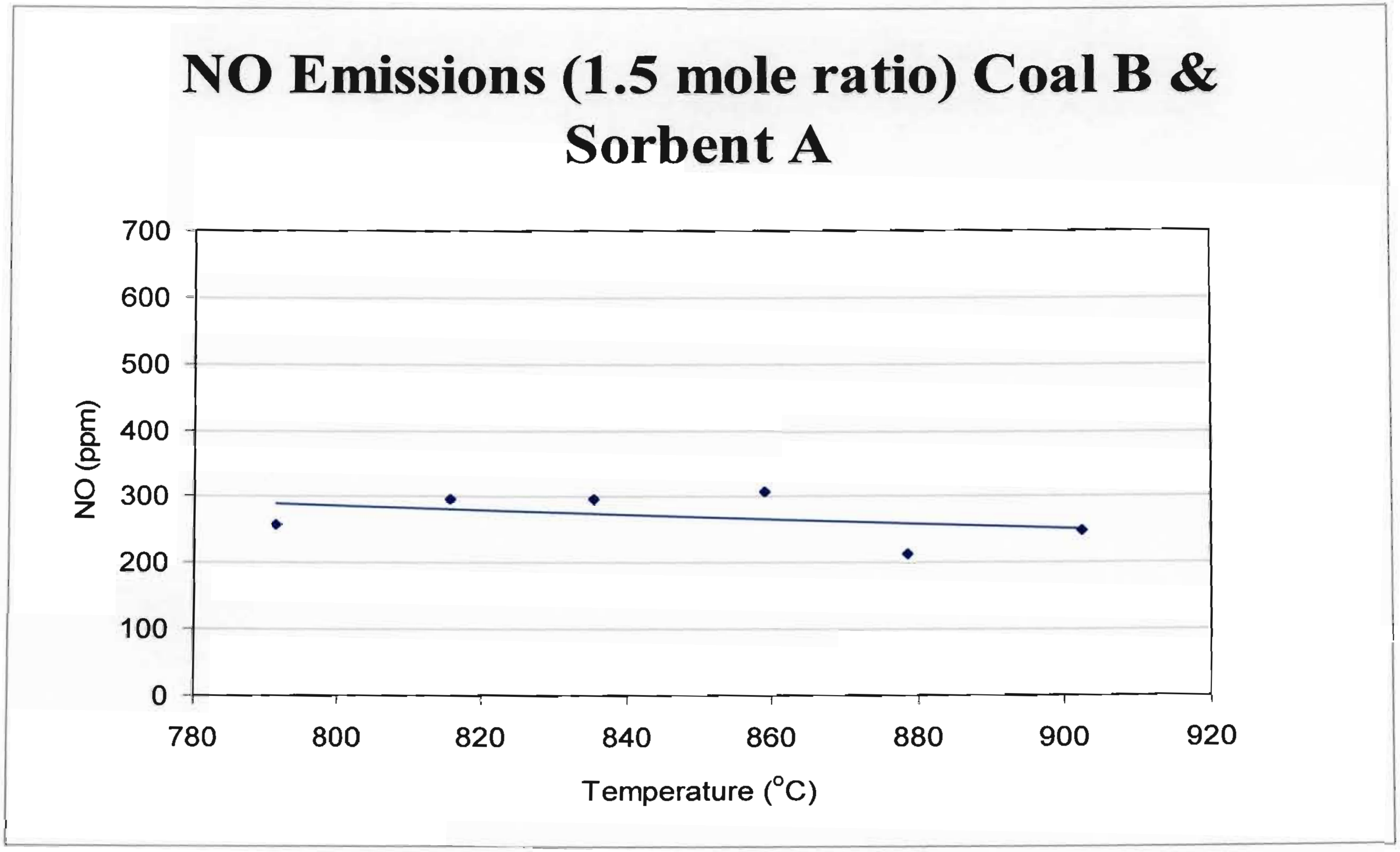


Figure B - 10 NO Emissions for Coal B Ca/S=1.5

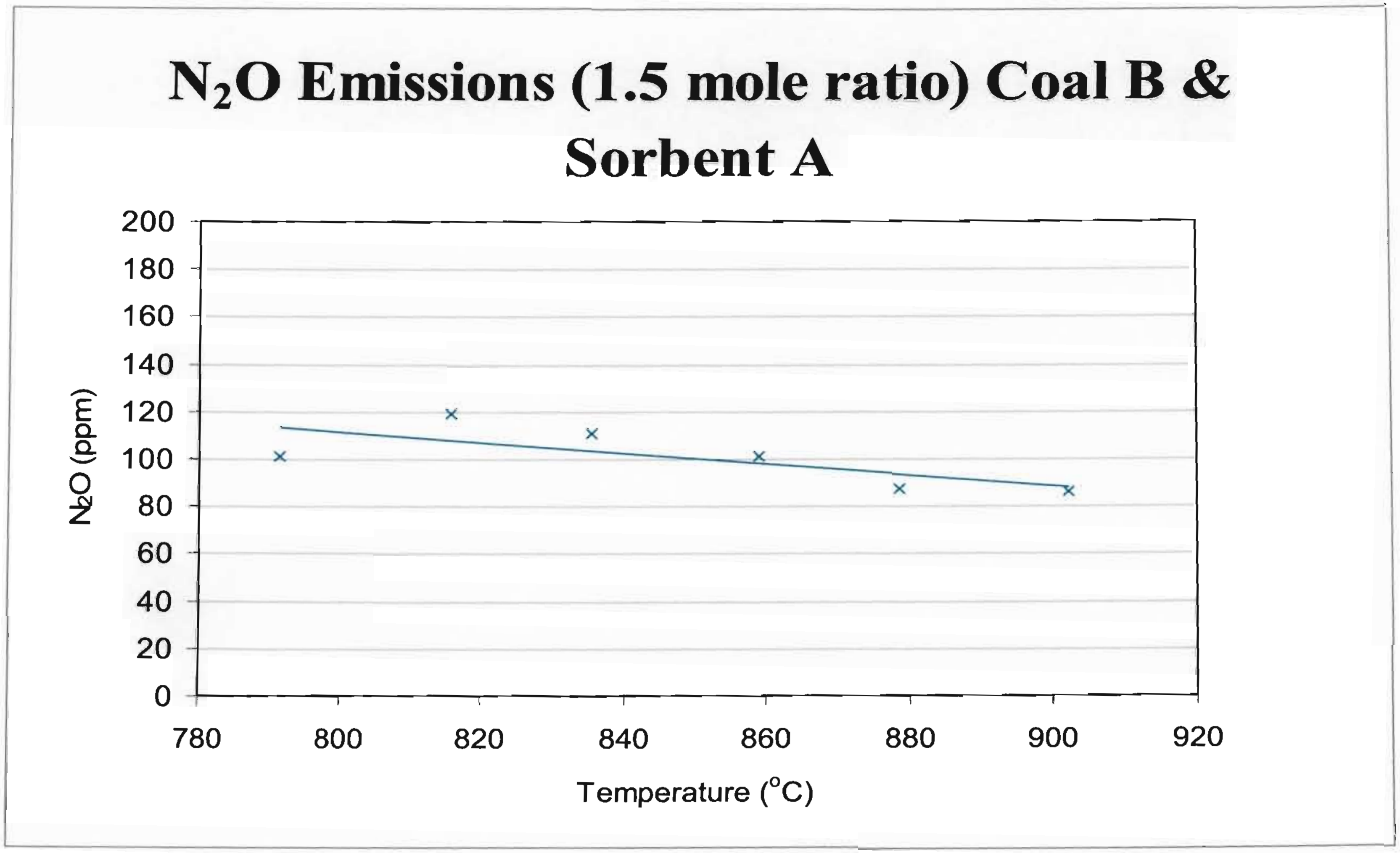


Figure B - 11 N_2O Emissions for Coal B Ca/S=1.5

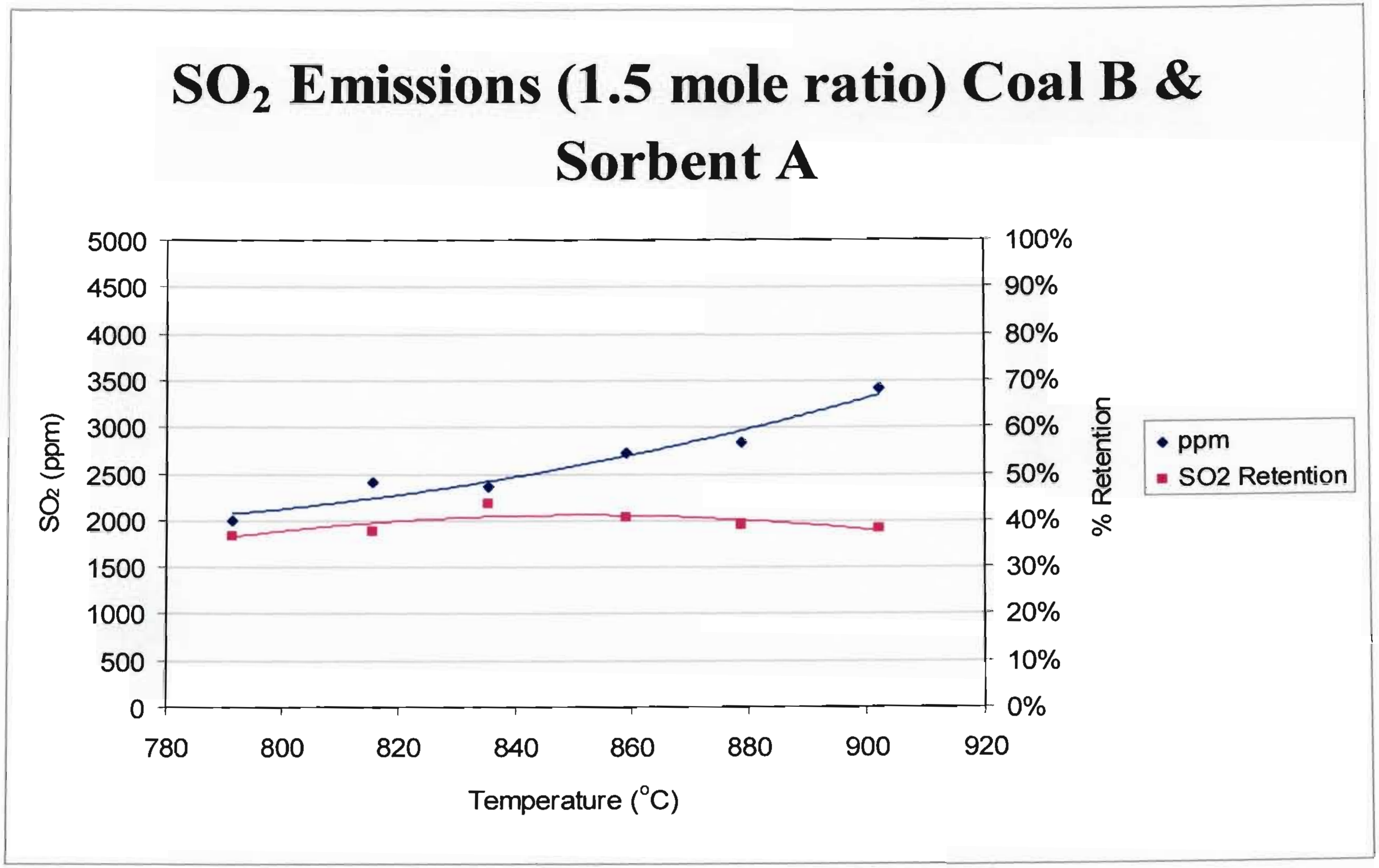


Figure B - 12 SO$_2$ Emissions for Coal B Ca/S=1.5

APPENDIX C

COAL C AND SORBENT A TESTS

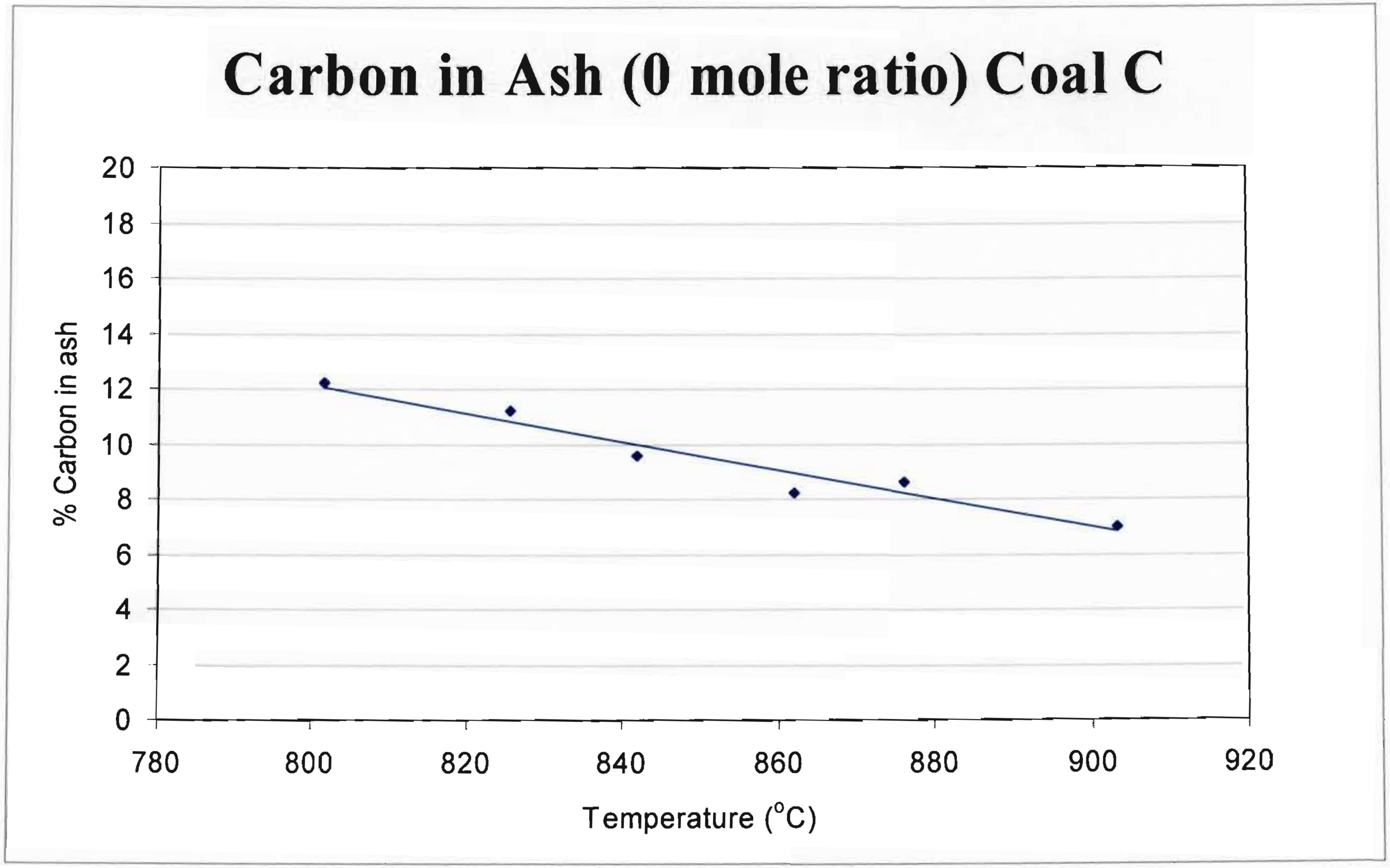


Figure C - 1 Carbon in Ash for Coal C Ca/S=0

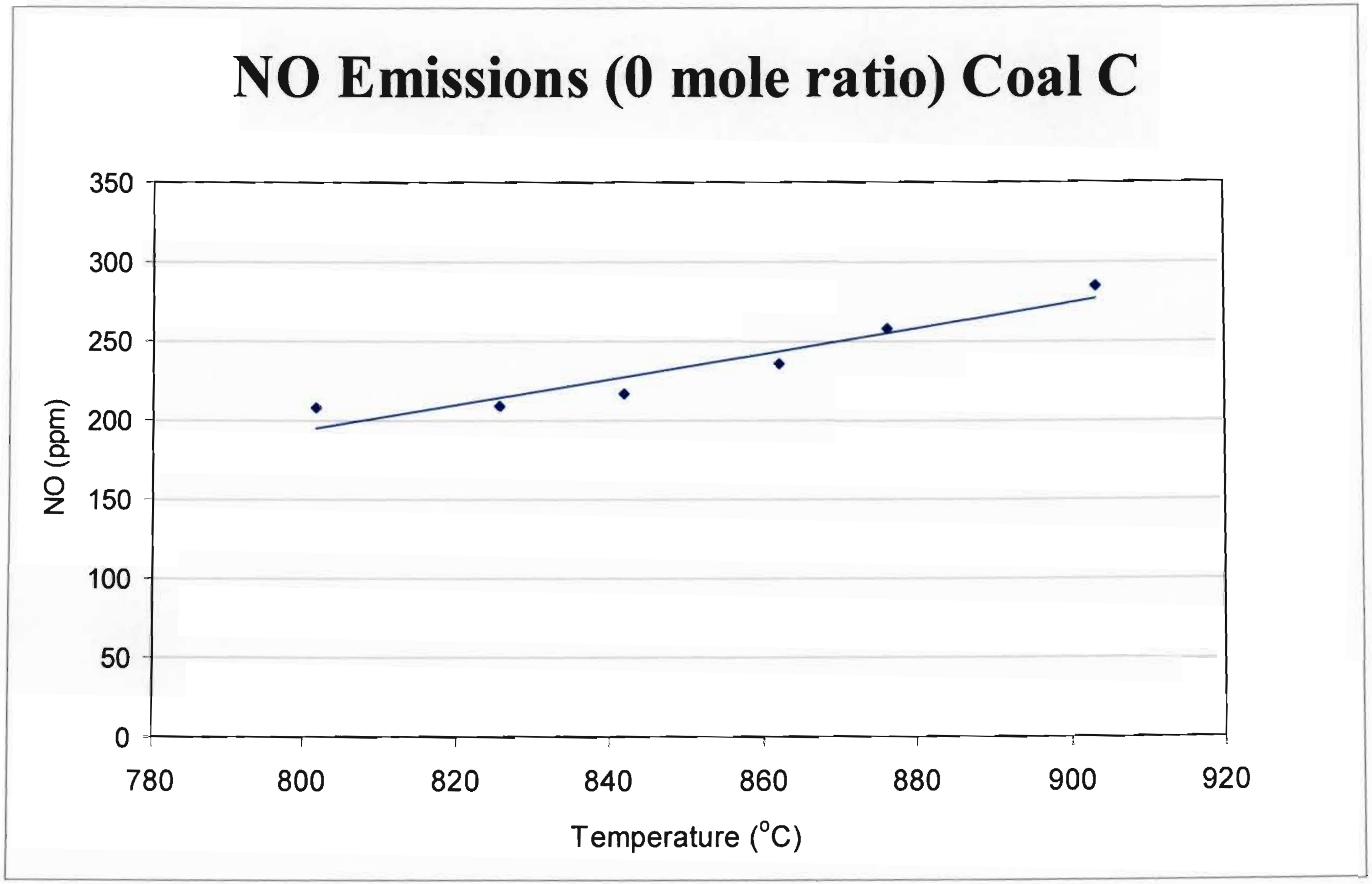


Figure C - 2 NO Emissions for Coal C Ca/S=0

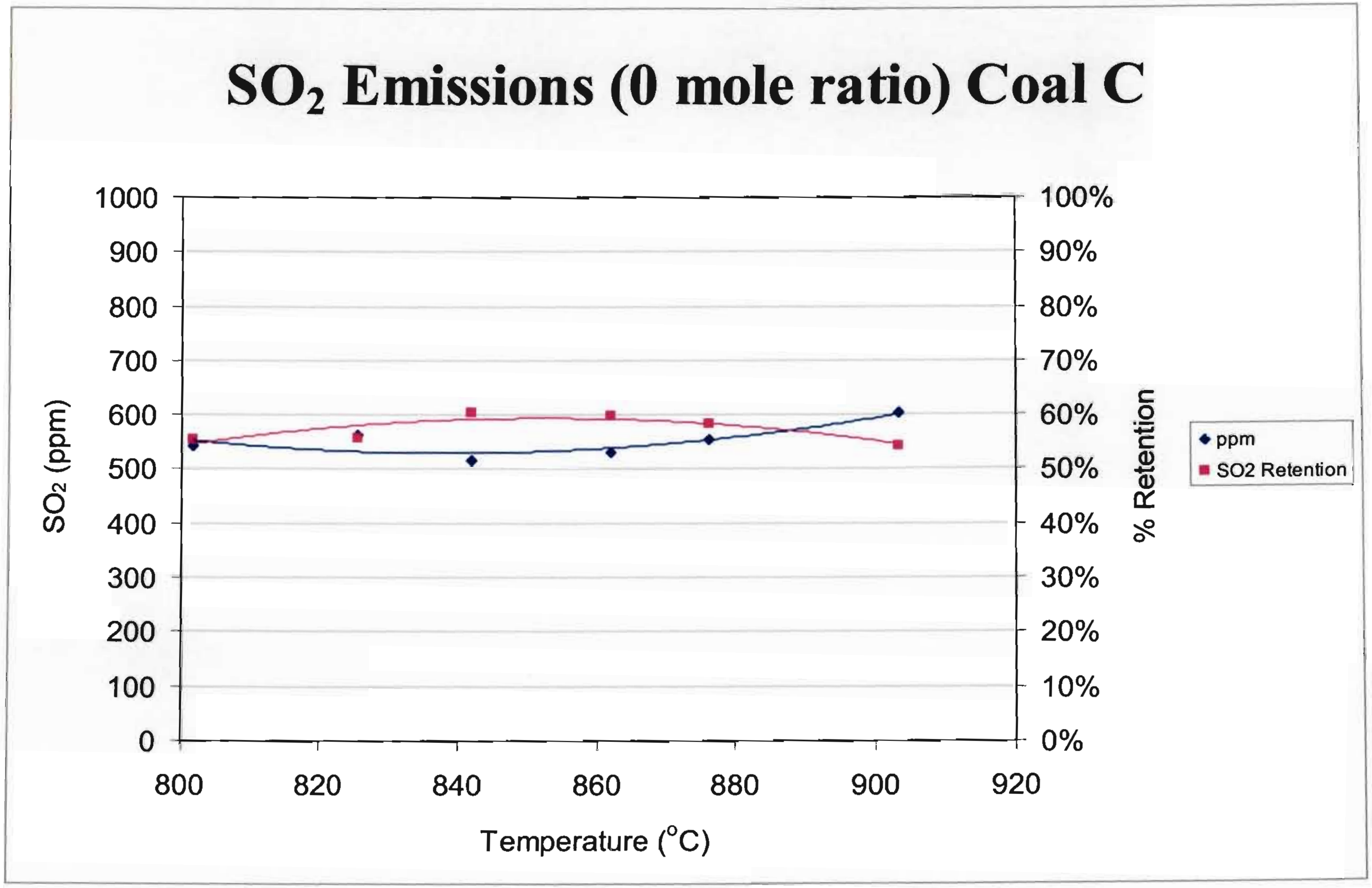


Figure C - 3 SO_2 Emissions for Coal C Ca/S=0

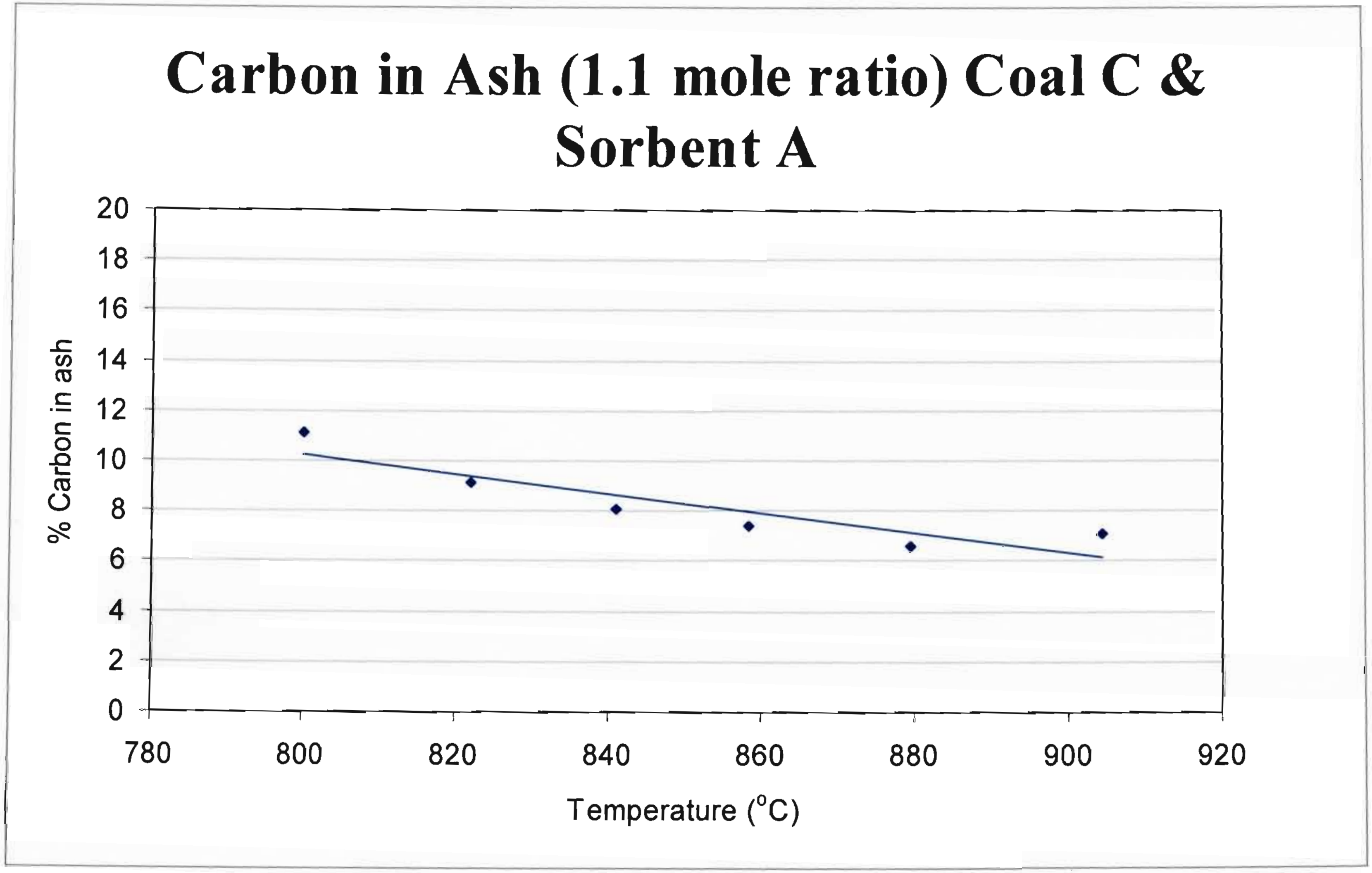


Figure C - 4 Carbon in Ash Coal C Ca/S=1.1

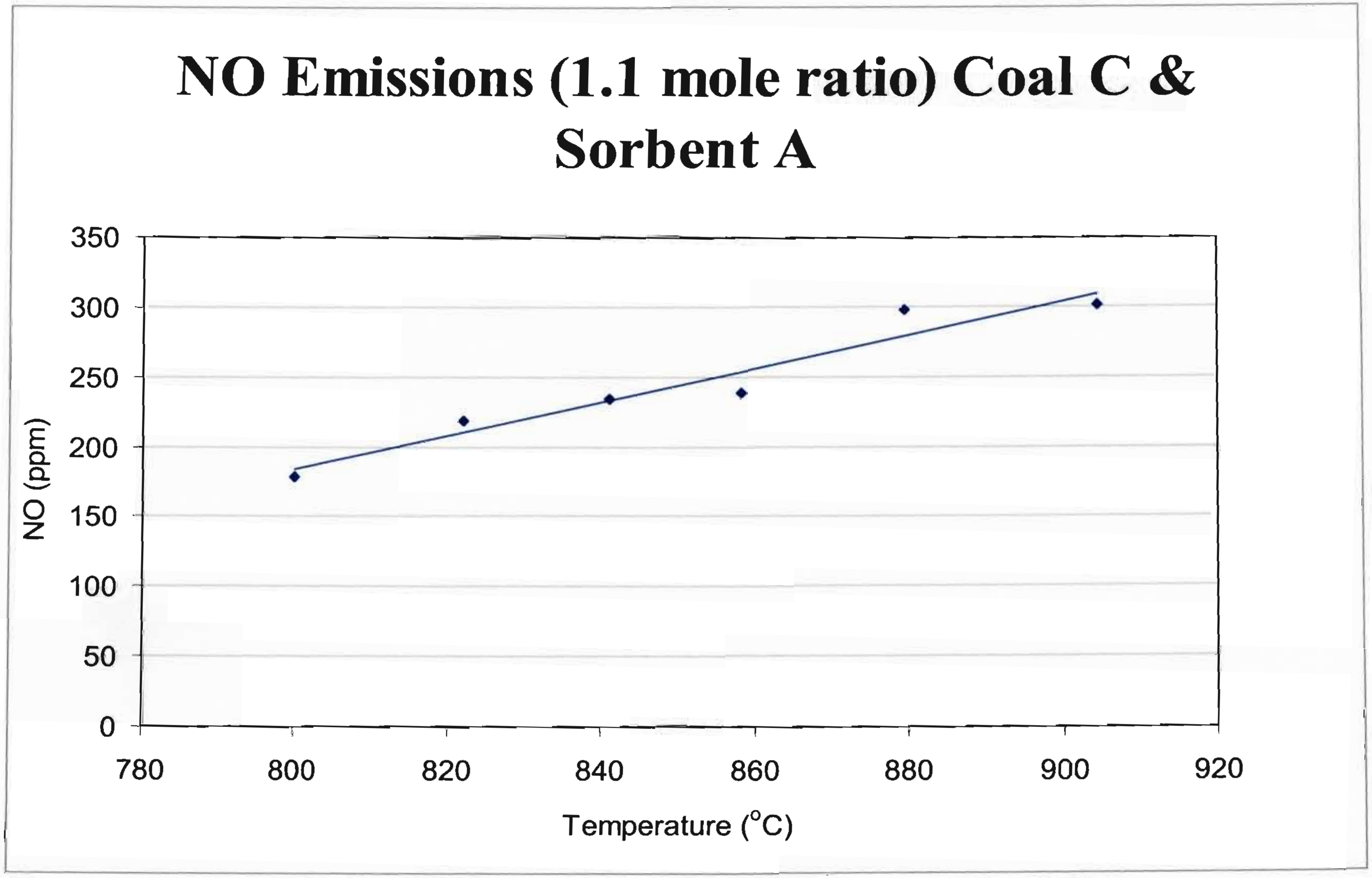
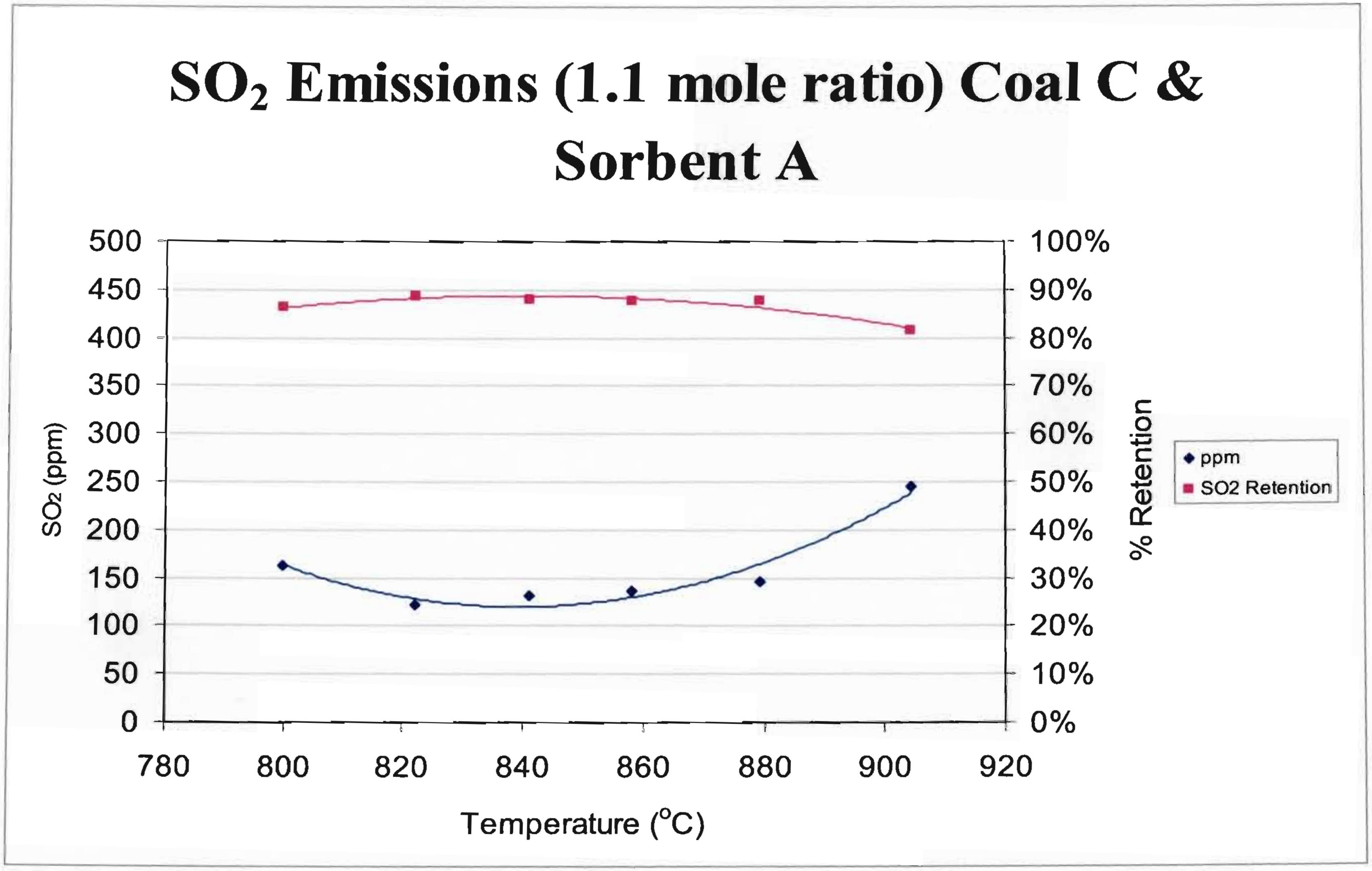


Figure C - 5 NO Emissions Coal C Ca/S=1.1



SO$_2$ Emissions (1.1 mole ratio) Coal C & Sorbent A

Figure C - 6 SO$_2$ Emissions Coal C Ca/S=1.1

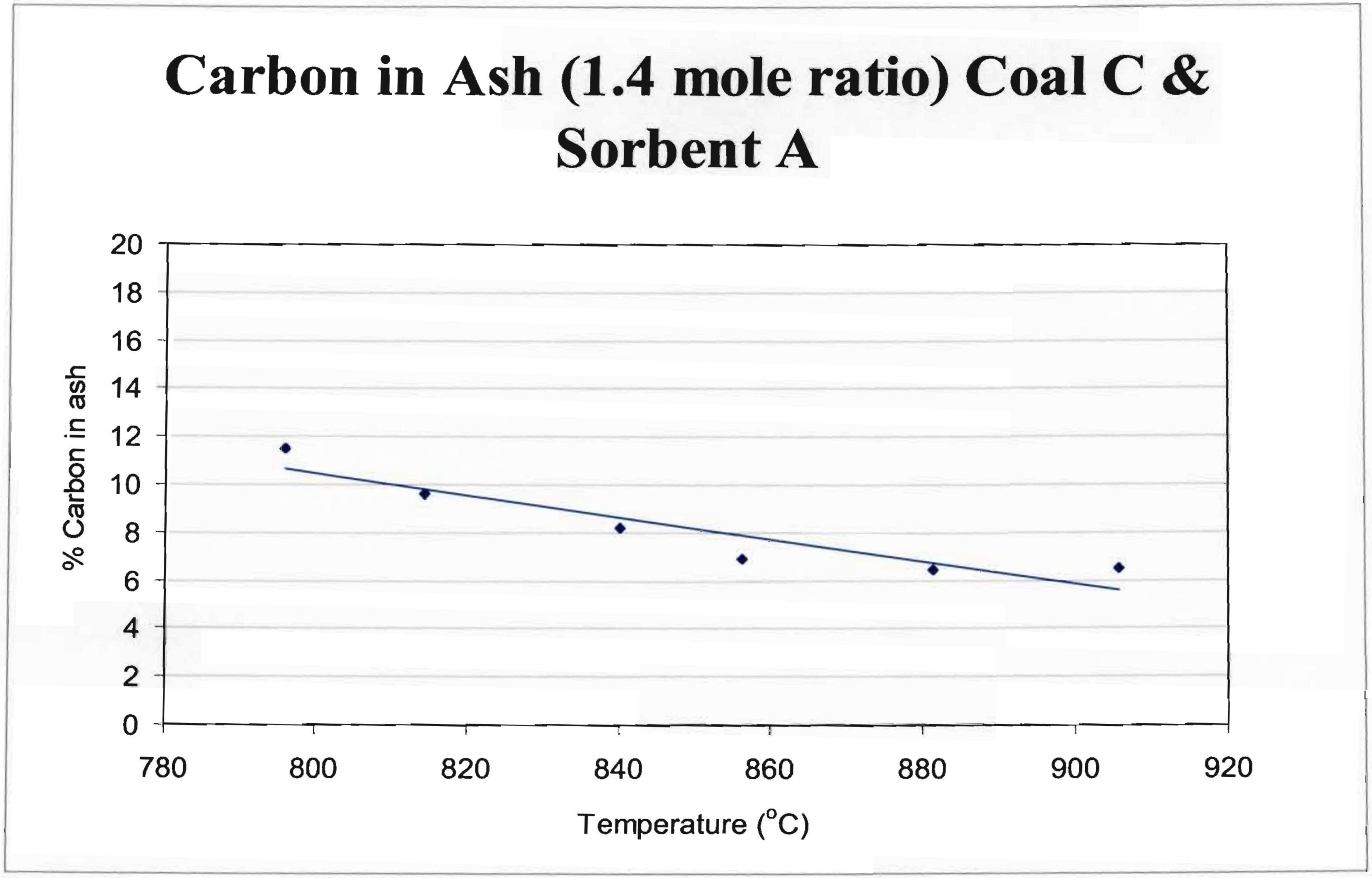


Figure C - 7 Carbon in Ash for Coal C Ca/S=1.4

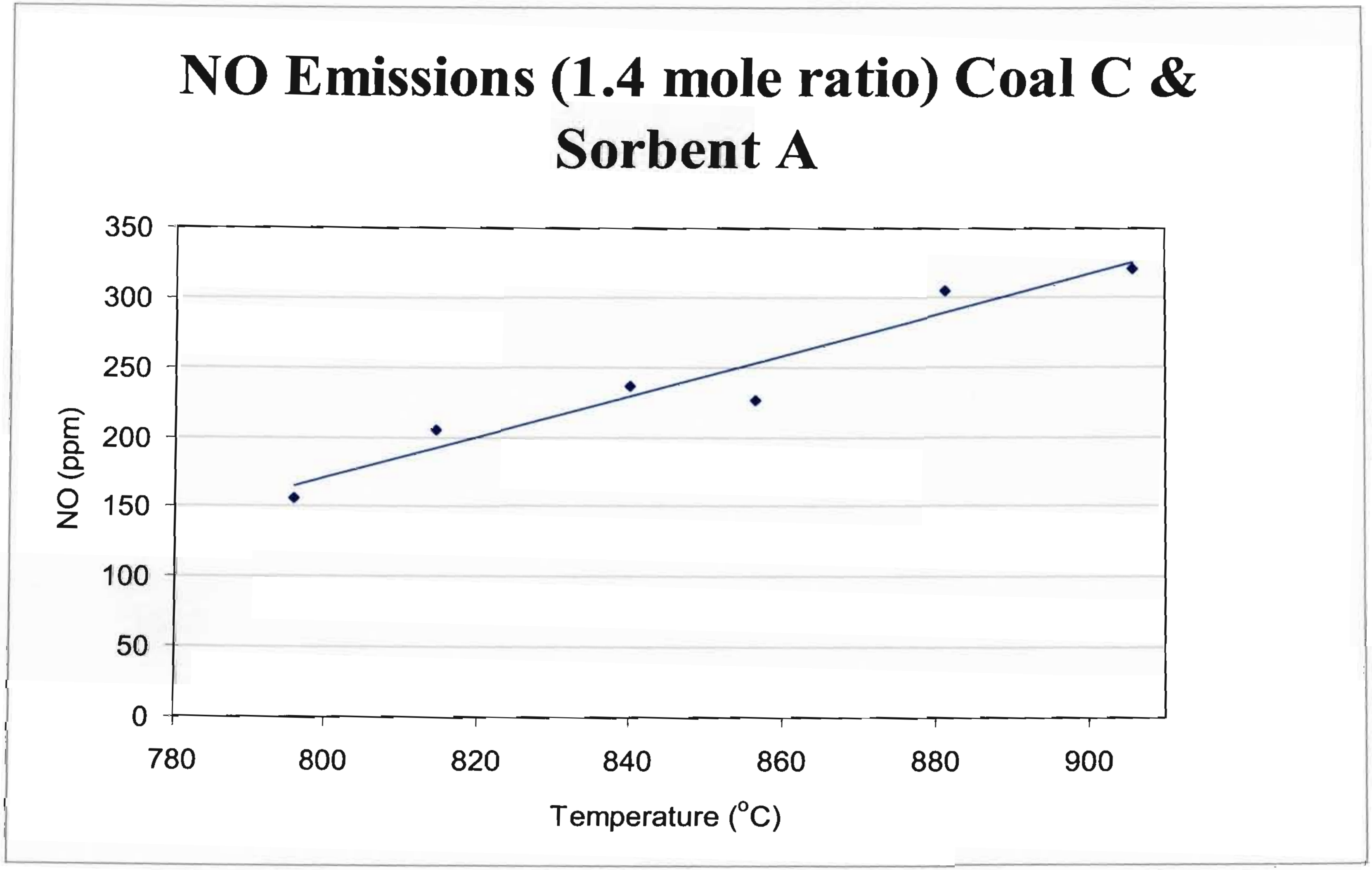


Figure C - 8 NO Emissions for Coal C Ca/S=1.4

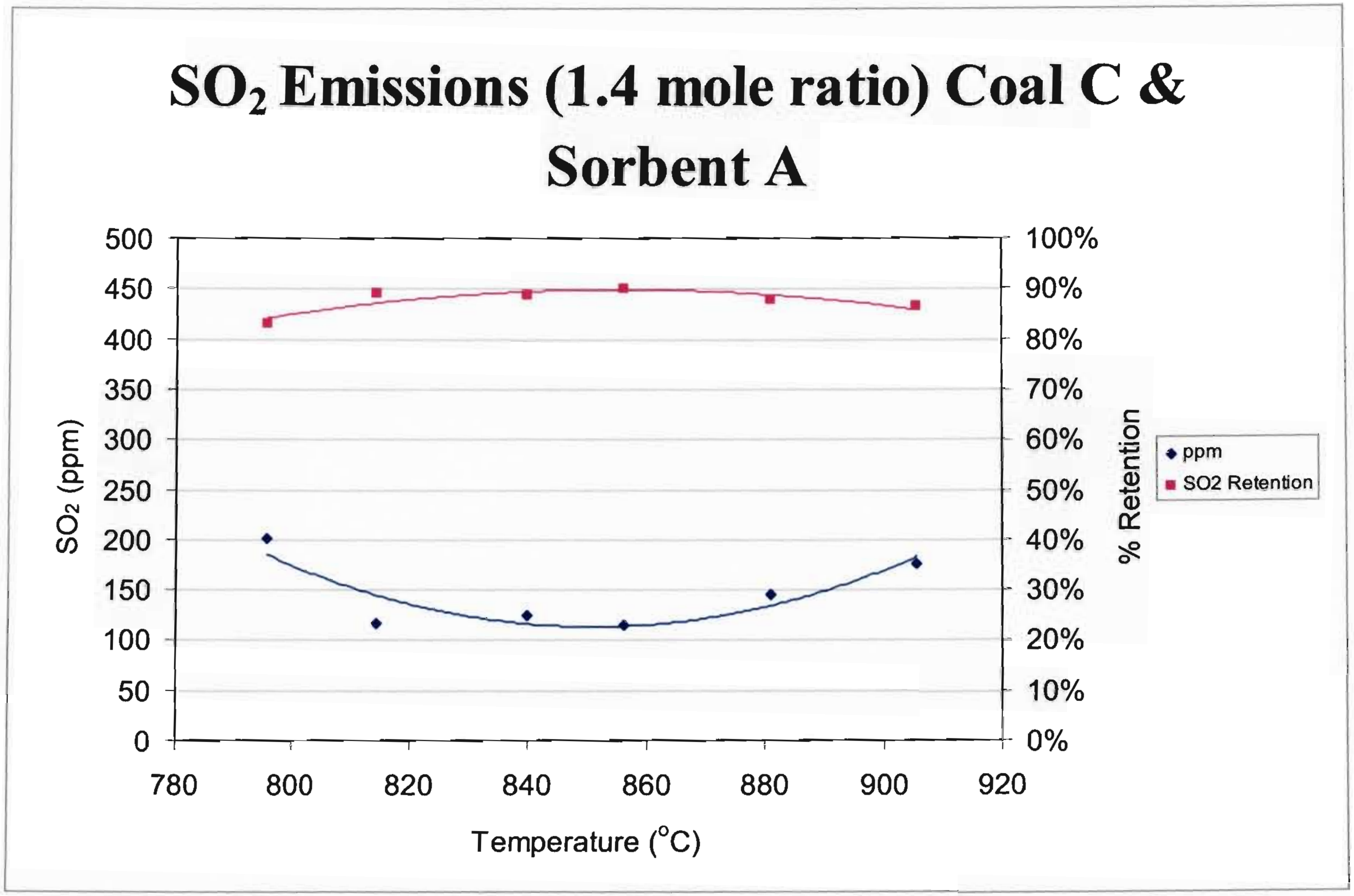


Figure C - 9 SO$_2$ Emissions for Coal C Ca/S=1.4

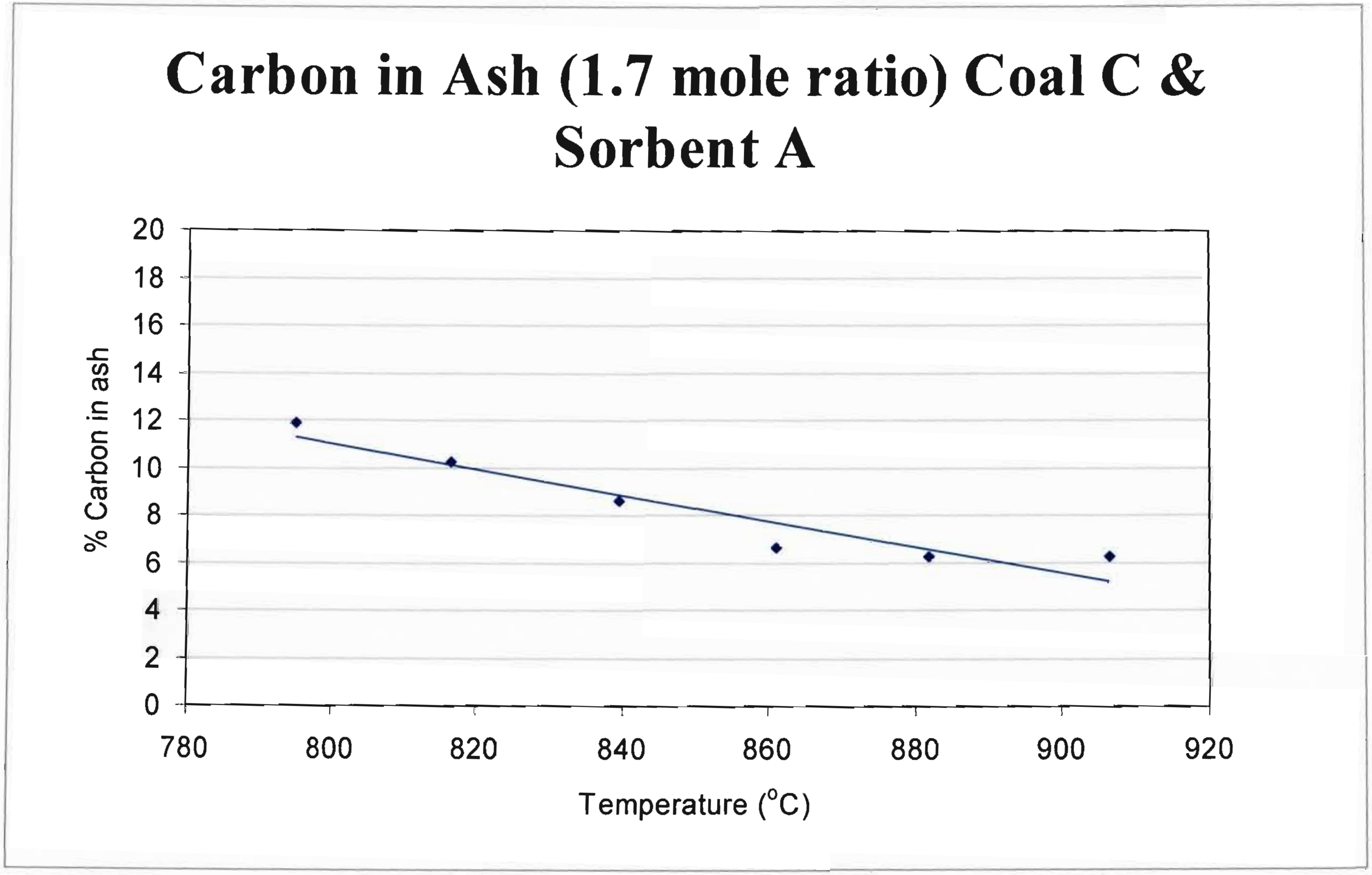


Figure C - 10 Carbon in Ash for Coal C Ca/S=1.7

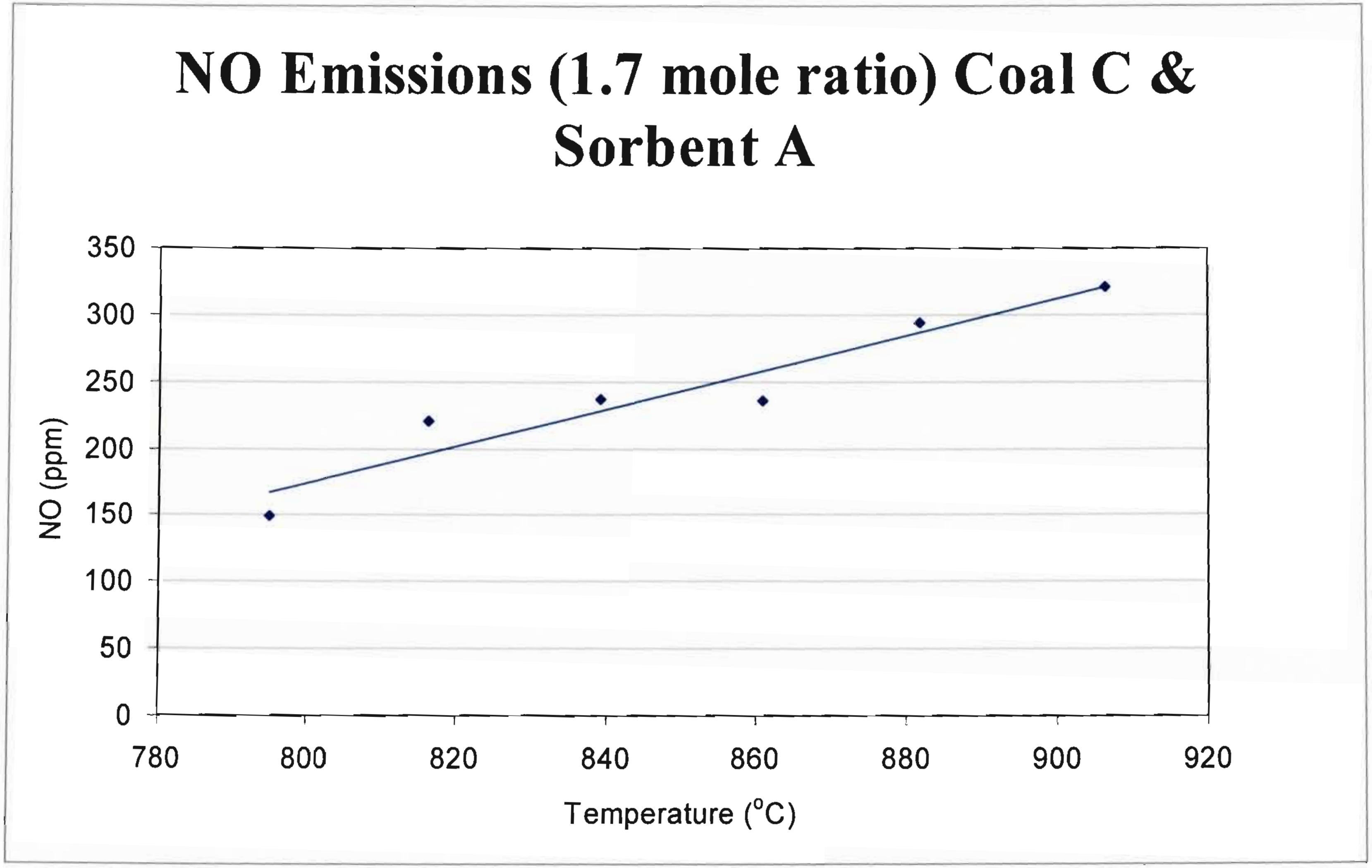


Figure C - 11 NO Emissions for Coal C Ca/S=1.7

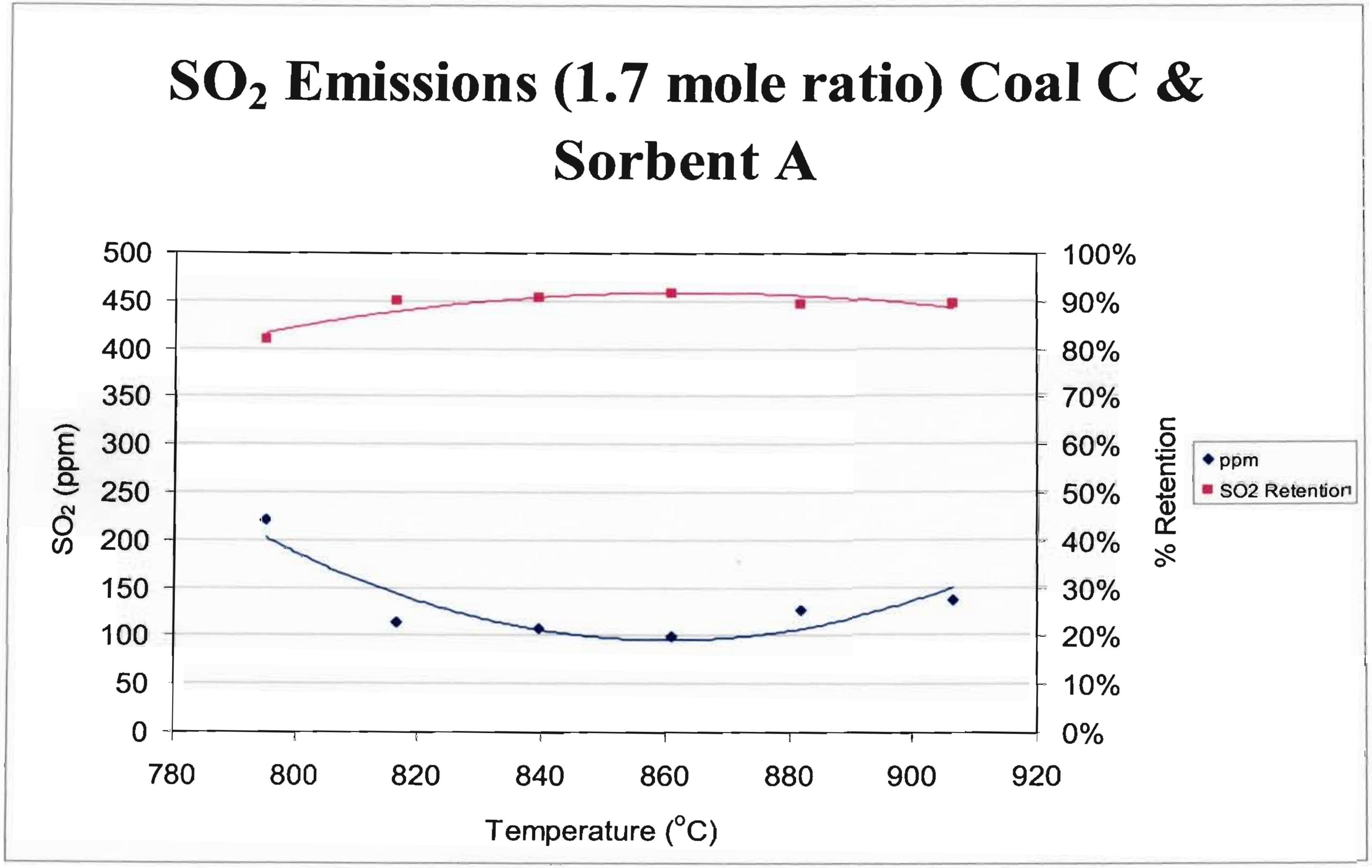


Figure C - 12 SO_2 Emissions for Coal C Ca/S=1.7