

# **Gaseous and Particulate Emissions**

by


**Nisha Beebal  
BSc(Hons) UDW**

*Submitted in partial fulfilment of the academic requirements for the award of the degree of Masters of Science in Engineering, in the Department of Civil Engineering, University of Durban- Westville, February 2000.*

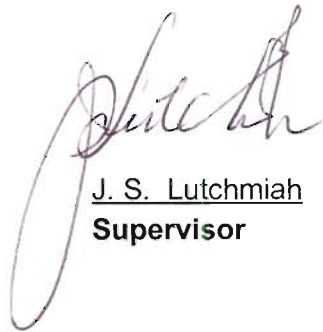
## DECLARATION

I, Nisha Beebal declare that unless otherwise indicated, this dissertation is my work and has not been submitted for degree purposes at any other University or Institution.

Signed

A handwritten signature in black ink, appearing to be 'Nisha Beebal', written over a horizontal line.

Nisha Beebal  
**Candidate**

A handwritten signature in black ink, appearing to be 'J. S. Lutchmiah', written over a horizontal line.

J. S. Lutchmiah  
**Supervisor**

## ABSTRACT

The need to limit sulphur dioxide, nitrous oxides and particulate matter emissions from fossil fuel combustion processes is now a major factor in the design, construction and operation of plants. Over the past 5 to 10 years, most developed countries have introduced legislations to control these emissions. This report was done to investigate emissions of sulphur dioxide ( $\text{SO}_2$ ), nitrogen oxides ( $\text{NO}_x$ ) and particulate matter from coal fired power plants.

$\text{SO}_2$  emissions contribute to an air pollution load that has been found to be linked to respiratory and heart problems. It increases the formation of acid precipitation, called acid rain.  $\text{NO}_x$  oxides to nitrogen dioxide ( $\text{NO}_2$ ).  $\text{NO}_2$  acts as an irritant to the respiratory tract, alveoli of the lungs and causes chest pains. Toxic elements concentrate on fine particulate matter because of their high surface-area to volume ratio. These particles penetrate and settle onto the lung wall, leading to respiratory disorders and lung diseases.

Due to the strict legislative requirements, flue gas detoxification processes were investigated. They include:

- Wet scrubbing
- Spray dry scrubbing
- Dual-alkali
- Wellman-Lord
- Activated carbon
- Magnesium oxides
- Selective catalytic recovery
- Circulating fluidised bed combustion
- Electromembrane desulphurisation

A study was done to show the possible cost implications associated with the wet scrubbing, spray dry scrubbing, selective catalytic recovery, circulating fluidised bed and electromembrane desulphurisation process. A variety of methods were used to calculate cost, due to costs being difficult to obtain. Most of the information was based on European, Asian and American installations. The objective of this economic analysis was to compare the different technologies. Because of the assumptions used, costs were always "order of magnitude" estimates. Capital and operating costs were calculated for these technologies, but they can only be used as an indication of costs and to allow for comparisons between the technologies. The costs were converted using the January 2000 exchange rates. Due to this, costs may be more inflated than what would normally be expected due to equipment being purchased earlier than January 2000. Of the processes investigated, the circulating fluidised bed process was found to be the most economical.

There is no inexpensive route for the removal of sulphur dioxide, nitrous oxides and particulate matter. The most suitable detoxification process will depend on the legislative requirements per country. South Africa is a signatory of the Montreal Protocol and the Framework for Climatic Change. Legislations for gaseous emissions were not considered a priority. Currently, Eskom's drive is to electrify townships rather than spend money on detoxification processes. However, future legislations may require more stringent control measures.

## ACKNOWLEDGEMENTS

I would like to extend my sincerest thanks to my sister Nita Beebal. Your advice, encouragement and help were invaluable to me.

I would like to thank Bruce Smith and Johan Liendenkamp, the environmental officers at Majuba and Tutuku Power Station, for affording me a glimpse of how a station is run.

Specials thanks to my dear friend Elton Singh who waited up for me, cooked me meals and provided a shoulder to lean on.

And last, but not least, I would like to express a debt of gratitude to my family and friends without whose support and understanding I could never have succeeded.

---

## TABLE OF CONTENTS

Declaration	ii
Abstract	iii
Acknowledgements	v
Table of Contents	vi
List of Figures	x
List of Tables	xi
List of Abbreviations	xii
Glossary	xiv

### CHAPTER 1: INTRODUCTION

1.1	Objectives	1-3
1.2	International Gas Emission Legislation	1-3
1.2.1	SO <sub>2</sub> Control Regulations	1-3
1.2.2	Nitrogen Oxides (NO <sub>x</sub> ) Legislation	1-4
1.2.3	Comparison of SO <sub>x</sub> and NO <sub>x</sub> emission limits between South Africa and World Energy Council (WEC) Members	1-4

### CHAPTER 2: SULPHUR DIOXIDE EMISSIONS

2.1	Effects on Human Health	2-1
2.2	Effects on Plants and Materials	2-1
2.3	Sulphur Dioxide (SO <sub>2</sub> ) control	2-2
2.4	Flue Gas Desulphurisation	2-3
	Non-Regenerable Processes	2-4
2.4.1.	Wet Scrubbing Process	2-4
	Process Description	2-4
	Advantages	2-7
	Disadvantages	2-7
2.4.2	Spray Dry Scrubbing (SDS)	2-7
	Process Description	2-7
	Problems	2-10
	Waste Disposal	2-10
2.4.3	Dual-Alkali	2-10
	Process Description	2-10
	Advantages	2-12
	Disadvantages	2-12

Regenerable Processes	2-12
2.4.4. Wellman-Lord	2-12
2.4.5 Activated Carbon	2-14
Advantages	2-14
Disadvantages	2-16
2.4.6 Magnesium Oxides	2-16
Advantages	2-16
Disadvantages	2-16

### **CHAPTER 3: NITROGEN OXIDES**

3.1 Effects on Human Health	3-1
3.2 Nitrogen Oxides (NO <sub>x</sub> ) Control	3-2
3.2.1 NO <sub>x</sub> Control by Combustion Modification	3-2
3.2.2 Selective Catalytic Reduction	3-3
Tail End Installations	3-4
High Dust Installations	3-4
3.2.3 Lurgi Circulating Fluidized Bed Reactor Process	3-4
Process Description	3-4

### **CHAPTER 4: PARTICULATE (FLY ASH) EMISSIONS**

4.1 Uses of Fly Ash	4-2
4.2 Effects on Human Health	4-2
4.3 Effects on Plants and Animals	4-3
4.4 Removal of Particulate (Fly Ash) Emissions	4-3
4.4.1 Cyclones and Multicyclones	4-4
4.4.2 Baghouse Filters	4-4
4.4.3 Electrostatic Precipitators (ESP)	4-4
4.4.4 Conclusion	4-5
4.4.5 Flue Gas Conditioning by Sulphur Trioxide	4-6
Chemical Conditioning Mechanism	4-6
The Process	4-6

### **CHAPTER 5: NEW EMERGING TECHNOLOGIES**

5.1 SNOX Technology for SO <sub>2</sub> and NO <sub>x</sub> Removal	5-1
5.1.1 Process Description	5-1
5.1.2 Advantages	5-1
5.2 Electromembrane Desulphurisation	5-3
5.2.1 Process Description	5-3
5.2.2 Process Equations	5-5
5.2.3 Conclusions	5-6
5.3 IHI In-Line Type FGD	5-6
5.3.1 Process Description	5-6

5.3.2 Advantages	5-7
------------------	-----

## **CHAPTER 6: MAJUBA POWER STATION**

6.1 Pulse Jet Fabric Filter	6-1
6.1.1 Design and Specifications of Pulse Jet Fabric Filter (PJFF)	6-3
6.1.2 Plant Layout and Operation	6-3
6.1.3 Correlation of Stack Emission Monitors	6-5
6.1.4 Plant Operational Problems	6-6
6.1.5 Waste Disposal	6-6
6.1.6 Advantages	6-7
6.1.7 Future Enhancements	6-7
6.2 Eskom's Enhancement Program	6-8

## **CHAPTER 7: ECONOMIC ANALYSIS**

7.1 Cost Components	7-1
7.1.1 Capital Costs	7-1
7.1.2 Annualized Costs	7-1
7.1.2.1 Fixed Operating Costs	7-1
Labor	7-1
Maintenance and Repairs	7-2
7.1.2.2 Variable Operating Costs	7-2
Utilities	7-2
7.3 Costs Calculations	7-2
7.3.1 Conversion of Currency	7-2
7.4 Economic Evaluation for a Wet Scrubbing and Spray Dry Scrubbing Process	7-3
7.4.1 Investment Costs	7-3
7.4.2 Operating Costs	7-3
Fixed Operating Costs	7-3
Variable Operating Costs	7-3
Waste Cost	7-3
Fly Ash	7-4
Values of Operating Costs	7-4
7.4.3 Discussion of Costs for Wet Scrubbing Process	7-6
7.4.4 Discussion of Costs for Spray Dry Scrubbing Process	7-6
7.5 Selective Catalytic Reduction	7-7
7.6 Circulating Fluidised Bed (CFB)	7-8
7.7 Electromembrane Desulphurisation	7-8
7.8.1 Discussion of Costs for Electromembrane Desuplurisation Process	7-10

## **CHAPTER 8: CONCLUSIONS AND RECOMMENDATIONS**

8.1 Conclusions	8-1
8.2. Recommendations	8-4



**REFERENCES****R-1****APPENDICES**

A1	Legislative Requirements	A1-1
A2	Gaseous and Particulate Emissions	A2-1
A3	Air Quality Results for Majuba Power Station	A3-1
A4	Cost Analysis	A4-1

---

## LIST OF FIGURES

Figure Number	Title	Page Number
1.1	Comparison of WEC member countries 1990 gaseous emissions and 1990 electricity generated	1-5
2.1	Relationship between coal sulphur content and the removal efficiency required to meet different emission limits	2-2
2.2	KRC limestone absorber process	2-5
2.3	Simplified flow diagram of Mitsubishi Heavy Industry wet lime scrubber/forced oxidation	2-6
2.4	Flow diagram of lime spray dry FGD process	2-8
2.5	Flow diagram of Thyssen dual alkali process	2-11
2.6	Simplified flowsheet of the Wellman-Lord process	2-13
2.7	Bergbau-Forschung activated carbon absorber for dual SO <sub>2</sub> and NO <sub>x</sub> removal	2-14
2.8	Simplified flowsheet of the Bergbau-Forschung/Uhde dry activated carbon process	2-15
2.9	Simplified diagram of United Engineers and Constructors wet magnesium oxide scrubber	2-17
3.1	Simplified flow diagram of the Lurgi circulating fluidised bed reactor process	3-6
4.1	Schematic diagram of an Electrostatic precipitator	4-5
5.1	Flow Diagram of the SNOX process	5-2
5.2	Schematic diagram of electromembrane sulphur dioxide treatment	5-3
5.3	Schematic diagram of electrodialysis cell with bipolar membrane	5-4
5.4	Overview of the IHI plant	5-7
5.5	Close up view of the U-bend	5-7
6.1	Schematic diagram of normal shaken bag filter	6-1
6.2	Schematic diagram of a Pulse Jet Fabric Filter	6-2
6.3	Schematic layout of Isolatable cells	6-3
6.4	Air system temperature control	6-4
6.5	Plan of waste disposal site at Tutuku Power Station	6-6
6.6	Landfill liner of hazardous waste disposal system	6-7
8.1	Total costs divided into capital and operating costs	8-3
A1.1	Relationship between calorific value and sulphur contents of various South African coals	A1-4
A3.1	Dust emission limits for 1998 in comparison with 1999	A3-3
A3.2	Dust concentrations from Majuba for 1998 and 1999	A3-6
A3.3	Ambient sulphur dioxide concentrations	A3-8
A4.1	Mass flow diagram of lime sulphur dioxide removal method	A4-3
A4.2	Mass flow diagram for electromembrane method of SO <sub>2</sub> removal	A4-9

## LIST OF TABLES

Table Number	Title	Page Number
7.1	Exchange rate	7-2
7.2	Values of constants of investment cost function	7-3
7.3	Operating costs	7-5
7.3	Summary of costs for wet scrubbing process	7-5
7.4	Summary of costs for spray dry scrubbing process	7-6
7.6	Summary of costs for the Electromembrane Desulphurisation Process	7-8
8.1	Capital and operating cost comparison	8-2
A1.1	Signatories of the sulphur dioxide protocol to the UN Convention on long-range transboundary air pollution	A-1
A1.2	Large Combustion Plant Ordinance (LCPO) European Community limits for bituminous coals	A1-1
A1.3	Provisions of the European Communities Large Combustion Plant Ordinance (LCPO): emission targets for existing plants	
A1.4	Current First World limits for sulphur dioxide	A1-2
A1.5	Table of comparison between WEC members SO <sub>2</sub> emissions, NO <sub>x</sub> emissions and electricity generated from hard coal power plant	A1-3
A2.1	Some gaseous and particulate substances from coal combustion	A2-1
A2.2	Characteristics of particulate control technologies	A2-6
A3.1	Majuba power station statistics	A3-1
A3.2	Emissions from Majuba from January 1998 to December 1999	A3-2
A3.3	Comparison of dust concentrations from mission monitors (Sick® OMD 41 to 3 transmissometers) are correlated against the actual measurements	A3-4
A3.4	Ambient sulphur dioxide concentrations	A3-5
A3.5	Maximum allowable concentration for atmospheric sulphur dioxide (DEAT, 2000)	A3-6
A4.1	Basic values and assumptions	A4-1
A4.2	Itemised cost analysis for sulphur dioxide removal from flue gasses by Electromembrane Desulphurisation Process	A4-15

## LIST OF ABBREVIATIONS

Abbreviations	Descriptions
BOD	Biochemical Oxygen Demand
CFB	Circulating Fluidised bed
CAAA	Clean Air Act Amendments
CAPCO	Chief Air Pollution Control Officer
CHE	Chemistry
DEAT	Department of Environmental Affairs and Tourism
DEM	Demark
DO	Dissolved Oxygen
EC	European Community
EIA	Environmental Impact Assessment
EPA	Environmental Protection Agency
ESP	Electrostatic Precipitator
ETH	Eastern Transvaal Highveld
FGC	Flue Gas Conditioning
FGD	Flue Gas Desulphurisation
FBA	Furnace Bottom Ash
Fin	Finland
FPM	Fine Particulate Matter
Fra	France
IHI	Ishikawajima-Harima Heavy Industries
Ire	Ireland
Isr	Israel
Ita	Italy
Kor	Korea
LCPO	Large Combustion Plant Ordinance
L/g	Liquid – gas ratio
LNB	Low NO <sub>x</sub> Burners
MCR	Maximum Continuous Rating
NAPAP	National Acid Precipitation Assessment Project
Neth	Netherlands
OHSP	Occupational Health and Safety Program Plan
PAN	Polyacrilonitrile
PFA	Pulverised Fuel Ash
PJFF	Pulse Jet Fabric Filter
PM	Particulate Matter
ppm	Parts per million
Pol	Poland
POM	Polycyclic Organic Matter
SA	South Africa
SCR	Selective Catalytic Recovery
SDS	Spray Dry Scrubbing
SNCR	Selective Non-Catalytic Recovery

SNOX	Sulphur Dioxide and Nitrogen Oxides Removal
SSI	Small Scale Industry
TDS	Total Dissolved Solids
TR	Transformer Rectifier
UCR	Unit Control Room
UK	United Kingdom
UNDP	United Nations Development Program
UNECE	United Nations Economic Commission for Europe
UNEP	United Nations Environmental Program
USA	United States of America
USEP	United States Environmental Protection Agency
UN	United Nations
WEC	World Energy Council
www	World Wide Web

## GLOSSARY

<b>Boiler slag</b>	Is a granular and angular material with the same upper and lower particle size limits as bottom ash. It has a uniform shiny-black colour, resembling crushed coal or black glass, but is very hard and abrasive.
<b>Bottom ash</b>	Bottom ash is a granular material with the same upper and lower particle size limits as concrete sand.
<b>Coal ash</b>	It is chemically similar to clay, essentially a calcined or fired clay which lends itself as a replacement for natural resources. There are generally three types of coal ash produced, fly ash, bottom ash and boiler slag.
<b>Epidemiology</b>	Study of the spread of diseases.
<b>Fly ash/PFA</b>	It is a powdery material made up of tiny, separate glass spheres. Chemically, fly ash consists of silicon, aluminium, iron and calcium oxides.
<b>Neutralisation</b>	An acid base reaction that produces neutral solution of a salt and water.
<b>Absorption</b>	Refers to a chemical reaction of sulphur dioxide with an absorbent.
<b>Adsorption</b>	Refers to sulphur dioxide and/or nitrogen oxides absorbing onto the surface of a solid, porous material. This material acts as a carrier of the chemicals undergoing a chemical reaction.
<b>Catalytic oxidation /reduction</b>	The catalyst facilitates the chemical reactions by participating in the intermediate reactions. However, in the final product, the catalyst has been restored.

## CHAPTER 1

### INTRODUCTION

Coal has been recognised for centuries as being a major source of energy and currently produces 40 % of the world's electricity (Tyson et al, 1988). With coal reserves currently forecast to last four times longer than oil and gas, coal is expected to be powering the world well into the 22nd century. World coal production is set to rise by 40% between 1993 and 2010 (Vernon, 1990). This is a reflection of the expected increase in demands worldwide, driven by the growing population and the desire for continuing improvement in living standards and industrialisation. In many developing countries such as South Africa, coal is realistically the only main source of fuel that can meet the escalating energy demand for the foreseeable future. Industries are now beginning to pay more attention to develop efficient and environmentally acceptable technology for coal extraction, preparation and use in combustion processes. Some of the methods currently employed include demand side management, clean coal technologies and energy efficiency.

Coal in its raw form contains many impurities, such as rocks that line coal seams, organic and inorganic sulphur, ash, trace metals, and other wastes (Hutton and Allen, 1982). When burned, the impurities are transformed into sulphur dioxide ( $\text{SO}_2$ ), nitrogen oxides ( $\text{NO}_x$ ), particulate and other wastes. The need to limit sulphur and nitrogen emissions from coal fired power stations is now a major factor in the design, construction and operation of a wide range of plants.

Controls over  $\text{SO}_2$  and  $\text{NO}_x$  emissions were initially prompted by high, localised concentrations of  $\text{SO}_2$  and  $\text{NO}_x$  in the air of towns and cities, arising from domestic and industrial coal burning. These high concentrations led to episodes such as the London smog of 1952 (Vernon, 1989), which resulted in over 40000 deaths. In response to this, the regulatory authorities introduced measures such as the burning of certain types of fuel, moving plants to less populated areas, usually outside cities and residential neighbourhoods, and requiring that all fossil fired plants be fitted with tall stacks. This method proved effective in reducing the levels of  $\text{SO}_2$  and  $\text{NO}_x$ . From 1970 onwards, research began to indicate that  $\text{SO}_2$  and  $\text{NO}_x$  emitted from tall stacks could be transported long distances through the atmosphere (Vernon, 1989). Pollutants released into the atmosphere undergo a series of chemical and physical reactions.  $\text{SO}_2$  present in the atmosphere reacts photochemically to produce sulphur trioxide, sulphuric acid droplets and salts of sulphuric acid. Nitrogen dioxide is converted in the atmosphere to nitrates and nitric acid (Wilson et al, 1980). Studies conducted have implicated  $\text{SO}_2$  and  $\text{NO}_x$  in damage to lakes, forest and rivers. This phenomenon has become known as acid rain.

Because acid rain is an international issue, more emphasis is being placed on the control and abatement of  $\text{SO}_2$  and  $\text{NO}_x$  emissions from coal fired power stations. In 1988, the Large Combustion Plant Ordinance (LCPO) was established, specifying the maximum  $\text{SO}_2$  emission limits for new combustion plants (Council of European Communities Directive, 1988). The limits for coal firing depend on the size of the plant and vary on a sliding scale from 2 000  $\text{mg}/\text{Nm}^3$  at 50 to 300 MW thermal input to 400 $\text{mg}/\text{m}^3$  for combustion plants greater than 300 MW plants (Council of European Communities Directive, 1988). In Germany,  $\text{SO}_x$  and  $\text{NO}_x$  emissions have been reduced by 90% and 75% respectively (World Energy Council Report, 1993). The costs however have been massive, with the total programme costing DEM 22bn in 1994, (approximately R 70.07bn, in 2000), with an annual operating cost of DEM 5bn in 1994 (approximately R15.93bn). All costs were converted using the current exchange rate of R6.05 per US\$ (Appendix 4, Table A4.1). Since the unification of Germany, this commitment has been extended to the formerly East German power stations (Financial Times, Germany). The measured  $\text{SO}_2$  emission concentrations, for Eskom power stations, ranged from 520 to 870 ppm, and the  $\text{NO}_x$  (measured as NO) emissions were 325 to 650 ppm (Tilly and Keir, 1995).

In South Africa, emission control has been primarily focused on particulate emission control, which have been reduced by 90.5% from 1982 to 1998, in terms of the mass emitted per unit of electricity generated (Smith, 1998). Majuba power station has been used to evaluate the mass of particles released into the atmosphere. Majuba is situated in the southeastern highveld of Mpumalanga Province at an altitude of 1 709 m above sea level. The power station consists 3 × 657 MW direct dry cooled, and 3 × 713 MW indirect wet cooled coal fired units totalling 4 110 MW generating capacity (Smith 1998). The primary control method used for the removal of boiler flue gas particulate is by means of the pulse jet fabric filtration (PJFF).

In order to meet the reduced  $\text{SO}_2$  and  $\text{NO}_x$  emission levels, desk top study of flue gas desulphurisation (FGD), flue gas denitrification and future emerging technologies was done. The following technologies were researched for  $\text{SO}_2$  dioxide removal only: wet scrubbing, spray dry scrubbing (SDS) and electrochemical membrane process. For  $\text{NO}_x$  removal only, the selective catalytic recovery (SCR) technique was investigated. Dual  $\text{NO}_x$  and  $\text{SO}_2$  removal techniques investigated were the circulating fluidised bed (CFB), SNOX and activated coke process.

The capital and operating costs were estimated for selected flue gas detoxification processes. The primary objective was to provide a cost comparison of the various technologies. Most cost estimates obtained were based on existing data from reference plants, assumptions made and by contacting suppliers for budget quotes e.g. LURGI South Africa.

---



## 1.1 OBJECTIVES

- To review international legislations of sulphur dioxide and nitrogen oxide and compare South African emission levels with the World Energy Council member countries.
- To review current and flue gas technologies to reduce sulphur dioxide, nitrogen oxides and particulate matter from coal fired power stations.
- To demonstrate particulate emission control using Majuba power station as a reference.
- To establish capital and operating costs for selected flue gas detoxification process. These include:
  - Wet scrubbing process
  - Spray dry scrubbing process
  - Circulating fluidised bed process
  - Selective catalytic recovery
- To do a cost comparison of the different flue gas detoxification processes.

## 1.2 INTERNATIONAL GAS EMISSION LEGISLATION

Because acid rain is associated with relatively low concentrations of sulphur dioxide ( $\text{SO}_2$ ) and nitrogen oxides ( $\text{NO}_x$ ) in the atmosphere, international regulations to control acid rain have focused on reducing total  $\text{SO}_2$  and  $\text{NO}_x$ . Stringent control are in place for large plants requiring removal of 80 to 90 % of  $\text{SO}_2$  and 70 to 80 %  $\text{NO}_x$  (Vernon, 1990). This primarily applies for already existing boilers. Depending on the legislative requirements per country, FGD processes will be used to control gaseous emissions.

### 1.2.1 $\text{SO}_2$ Control Regulations

The UN Convention on Long-Range Transboundary Air Pollution established the responsibility of governments to reduce cross-boundary pollution. This convention came into force in 1983, and in 1985 21 countries signed a protocol to this convention, pledging a reduction in national  $\text{SO}_2$  emissions by 30 %. The signatory countries are listed in Table A1.1 (Appendix 1), (Vernon, 1990). Missing from the list are UK and USA because they did not sign the protocol. The United Nations (UN) following the 1992 Stockholm Conference on the human environment took the first international action on  $\text{SO}_2$  control.

In 1988, a more detailed and enforceable regulation on  $\text{SO}_2$  emissions was adopted by the European community known as the Large Combustion Plant Ordinance (LCPO). The emission limits set out by the LCPO are given in Table A1.2 (Appendix 1). Table A1.2 shows that the limits for coal firing depend on the size of the power plant and vary on a sliding scale from 2 000  $\text{mg/Nm}^3$  at 50 to 300 MW thermal input to 400  $\text{mg/Nm}^3$  for

combustion plants greater than 300 MW (Council of European Communities Directive, 1988). The eventual aim of the LCPO is to gradually reduce SO<sub>2</sub> emission levels by 70 % of the 1980 levels, by 2003 (Table A1.3, Appendix 1).

### 1.2.2 Nitrogen Oxide (NO<sub>x</sub>) Legislation

Two international agreements are in existence for nitrogen oxides (NO<sub>x</sub>).

#### NO<sub>x</sub> Protocol

This is applicable to the European Community (EC) and was signed in 1988. It specifies that signatories are committed to reduce their emissions by 30 % between 1995 to 1998 in order to achieve a 10 % reduction in the total EC emissions

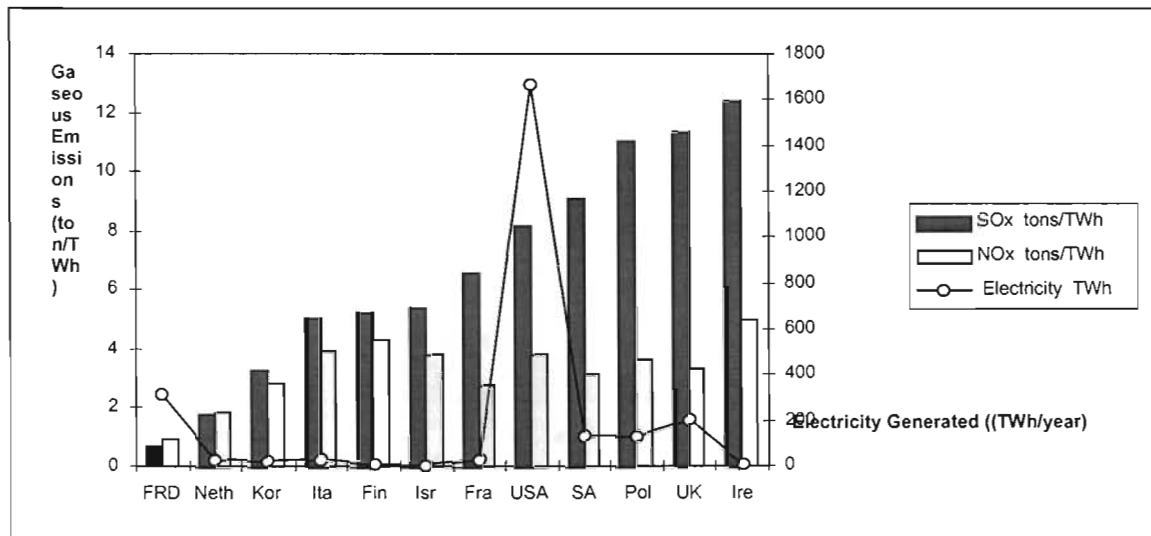
#### United Nations Economic Commission for Europe (UNECE) Convention of Long-Range Transboundary Air Pollution

The UNECE Convention of Long-Range Transboundary Air Pollution covers all emissions from stationary and mobile sources. On signing the agreement, NO<sub>x</sub> emissions were frozen at the 1987 level until 1994. This protocol was signed in 1988.

### 1.2.3 Comparison of SO<sub>x</sub> AND NO<sub>x</sub> emission limits between South Africa and World Energy Council (WEC) Members

South Africa is not a signatory to either the LCPO or the UNECE, primarily because these international agreements were designed for the European Community. South Africa is a signatory to the Montreal Protocol and the Framework for Climatic Change (World Energy Council Report, 1993).

South Africa generally burns coals that have low sulphur contents. Figure A1.1 (Appendix 1) shows the relationship between the calorific value and the sulphur content for a range of low sulphur coals. Although South Africa burns lower sulphur coal than other WEC member countries, the emissions per terawatt per hour are relatively high compared to the WEC member countries (Figure 1.1). South African coals have a relatively low calorific value, therefore large quantities of coal are burnt to compensate for this (Naude, 1991). Cumulatively this has a negative impact overall, due to more gaseous and particulate emissions being released.



**Figure 1.1:** Comparison of WEC Member countries 1990 gaseous emissions and 1990 electricity generated (Naude, 1991).

## CHAPTER 2

### SULPHUR DIOXIDE EMISSIONS

Eskom Power Stations collectively burn about 500 million tons of coal on the Eastern Transvaal Highveld, and produce about 1 million tons of sulphur dioxide gas (Tyson et al, 1988). The tall stacks fitted to Eskom power stations are very effective in limiting ground level concentrations of sulphur dioxide ( $\text{SO}_2$ ) and nitrogen oxides ( $\text{NO}_x$ ). Stack heights of 220 metre are optimal for the Eastern Transvaal Highveld (Smith 1998).

The atmosphere at an altitude of 1000m contains  $30 \times 10^6 \text{ m}^3$  of air. If all the sulphur dioxide produced by the power stations were confined over a full year in this volume of air, the resulting concentration (after one year) of sulphur dioxide would be 17 parts for every one billion parts of air. At this concentration it would take a person living on the Eastern Transvaal Highveld 20 years to circulate just one litre of sulphur dioxide through their lungs. Compared to the 16 000L of naturally occurring hydrogen, methane, nitrogen oxide and carbon monoxide that they would also breathe in over the same period, the concentration of sulphur dioxide is thus low and well within the limits considered to be harmless. The EPA standards considers harmless levels of  $\text{SO}_2$  to be 100 parts per billion averaged over any 24 hour period, and 30 parts per billion averaged over the whole year (Power-Gen, 1998). The average  $\text{SO}_2$  concentrations on the ETH over a whole year as measured by all monitoring stations is about 7 parts per billion (Cock and Koch, 1991).

#### 2.1 EFFECTS ON HUMAN HEALTH

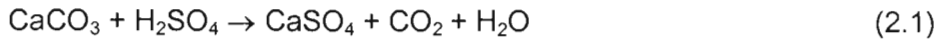
Both sulphur dioxide and sulphuric acid molecules paralyse the hair-like cilia which line the respiratory tract. Without the regular sweeping action of the cilia, particles are able to penetrate the lungs and settle there. These particles usually carry with them concentrated amounts of sulphur dioxide, thus bringing the irritant into direct contact with the delicate lung tissue. This often fosters the development of chronic respiratory diseases, particularly bronchitis and pulmonary emphysema. Appendix 2, Table A2.1 shows the effect of  $\text{SO}_2$  on human health.

#### 2.2 EFFECTS ON PLANTS AND MATERIALS

Exposure at low concentrations of sulphur dioxide for long periods results in brownish red or bleached white areas on the blades of the leaves. Plants are sensitive during the growing season especially during periods of intense light, high relative humidity, adequate moisture, and moderate temperature.

The sulphur dioxide released during the burning of coal reacts photo-chemically to produce sulphuric acid aerosols. Sulphuric acid aerosols formed, readily attacks

building materials, especially those containing carbonates such as marble, limestone and roofing slate. The carbonates are replaced by the sulphates that are water soluble as indicated by the following equation:



The calcium sulphate ( $\text{CaSO}_4$ ) formed in this process is washed away, leaving a pitted discoloured surface. Sulphur dioxide can remain airborne for 2 to 4 days and can be transported as far as 1000 km. Many of our cathedrals, public buildings and sculptures have deteriorated due to the exposure of the by-products of sulphur oxides (Fuggle and Rabie, 1983).

Sulphuric acid mists damage cotton, linen, rayon, nylon and is also found to weaken and disintegrate leather. Excess exposure to sulphur dioxide accelerates corrosion rates for metals such as iron, zinc, copper, and nickel, especially at relative humidities over 70 % (Ballinger Energy Series, 1979).

### 2.3 SULPHUR DIOXIDE ( $\text{SO}_2$ ) CONTROL

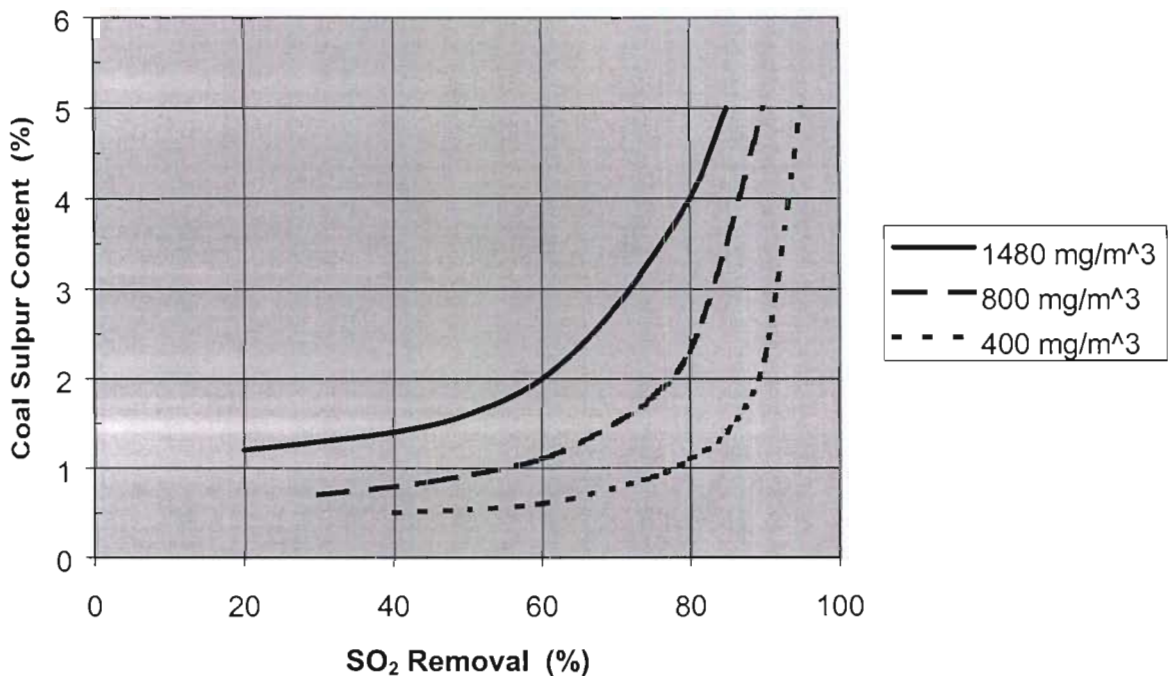


Figure 2.1: Relationship between Coal Sulphur Content and the Removal Efficiency Required to meet Different Emission Limits (Takeshita, 1993).

The flue gas desulphurisation (FGD) technology selected by an industry is dependent on the applicable legislative requirements.

Germany has strict legislative requirements of a maximum of 400 mg/m<sup>3</sup> of sulphur dioxide emitted into the atmosphere. Germany burns coals with low sulphur content (1% to 2%). A 1.5% sulphur coal requires top of the range FGD installations that has a greater than 85% removal efficiency (Figure 2.1). Due to this, German utilities have embarked on a DEM22 billion program to install FGD on all power stations, a commitment that has been extended to East Germany (Takeshita, 1993)

In the USA, the Clean Air Act Amendments (CAAA) phase two limit is set at 1480 mg/m<sup>3</sup> (Appendix 1, Table A1.4). It is therefore probable that sorbent injection can be used because for the same type of coal (S = 1.5%) it has a removal efficiency of 50 to 70% (Figure 2.1).

## 2.4 FLUE GAS DESULPHURISATION

Three options are available to reduce the amount of sulphur in coal emitted as sulphur dioxide during combustion:

- The coal is treated to reduce the sulphur content.
- The combustion process may be such that low levels of sulphur dioxide are released.
- Treat the combustion gases prior to their release to the atmosphere. This process is known as flue gas desulphurisation (FGD).

In the United Kingdom, there are well over 100 FGD processes available at various stages of development, from laboratory/pilot scale through to full-scale commercial operation (Klingspor and Cope, 1987). Generally, 90% SO<sub>2</sub> removal efficiencies are achieved with most FGD processes. The majority of FGD systems use an alkaline sorbent to absorb sulphur dioxide. FGD systems can be grouped into two categories, regenerable and non-regenerable systems. In non-regenerable systems, the sulphur dioxide is permanently bound by the alkaline sorbent and must be disposed of as waste or sold as a by-product, i.e. gypsum. In regenerable systems, the sulphur dioxide is released in a regeneration process following the absorption or adsorption process. The regenerated absorbent/adsorbent is recycled to the sulphur dioxide scrubbing step. The recovered sulphur dioxide from the regeneration process may be further processed to obtain sulphuric acid, elemental sulphur or liquid sulphur dioxide.



## NON-REGENERABLE PROCESSES

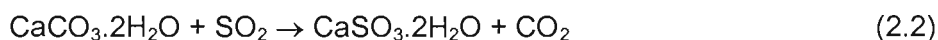
### 2.4.1. Wet Scrubbing Process

The wet scrubbing process is the most widely used FGD process. Of these scrubbers, 85% used calcium based sorbents, 10% use sodium based sorbents and the remaining 5% use other reagents (Klingspor and Cope, 1987). The main advantage of wet scrubbers is the high removal efficiencies achieved and the saleable by-products produced from the process. South African local markets can only absorb a limited amount of gypsum and the remaining by-products would have to be disposed of in landfills. The waste from the tower consists of a solution containing calcium sulphite. As such, the wastes need to be stored in a clay-lined dumping area. This increases disposal costs. South Africa has a limited and distant supply of limestone and the alternative of dolomite would require increase tonnage, leading to increased transport, operating and waste disposal (Weiler, 1992).

#### Process Description

(Refer to Figure 2.2)

The sulphur dioxide reacts with the sorbent in the spray tower to form a calcium compound. If limestone ( $\text{CaCO}_3$ ) is used as the sorbent, the primary chemical reaction is:

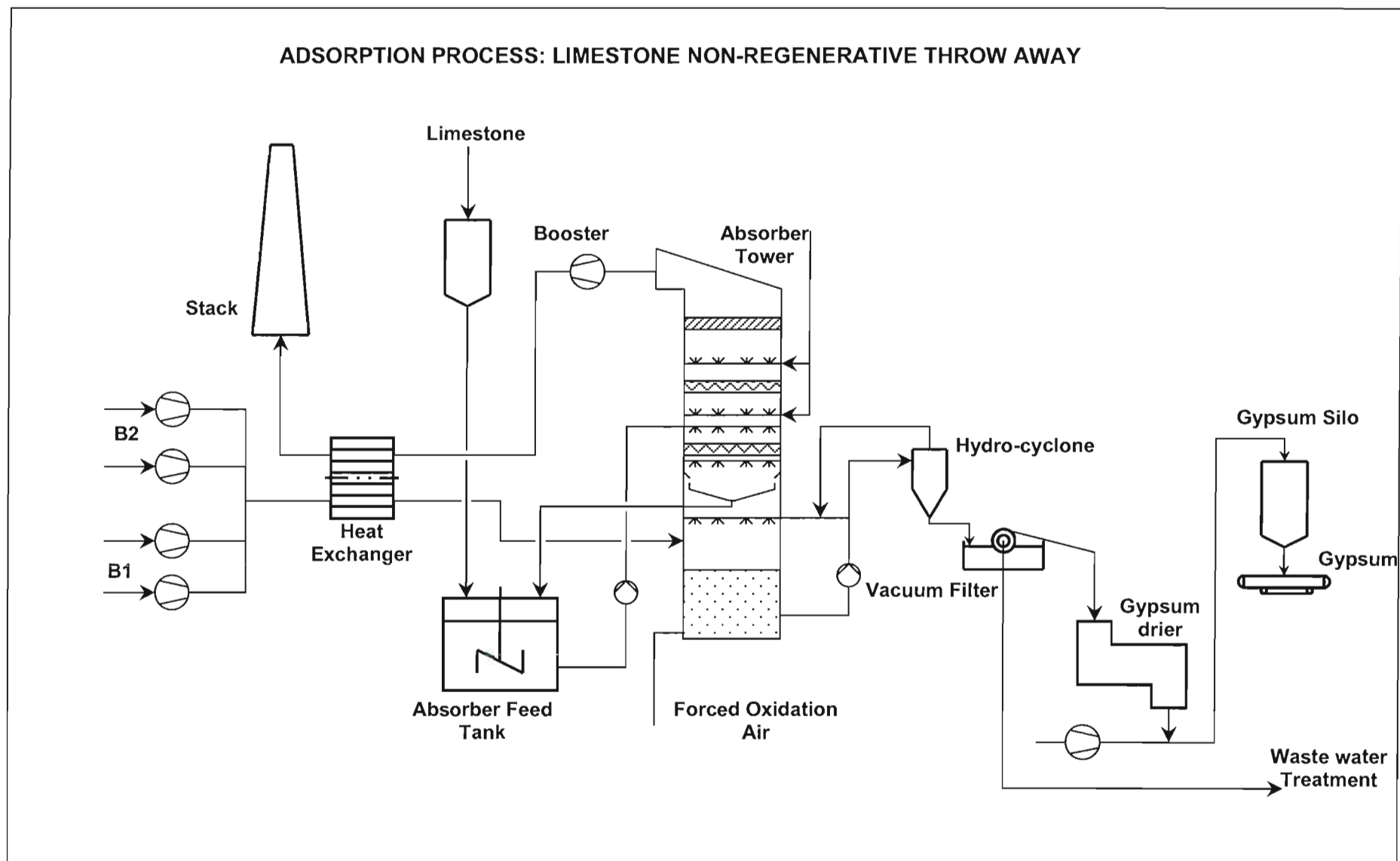


If lime ( $\text{CaO}$ ) is used as the reagent, it is first processed by mixing it with water to form slaked lime, also known as calcium hydroxide ( $\text{Ca(OH)}_2$ ). The basic chemical reaction is:



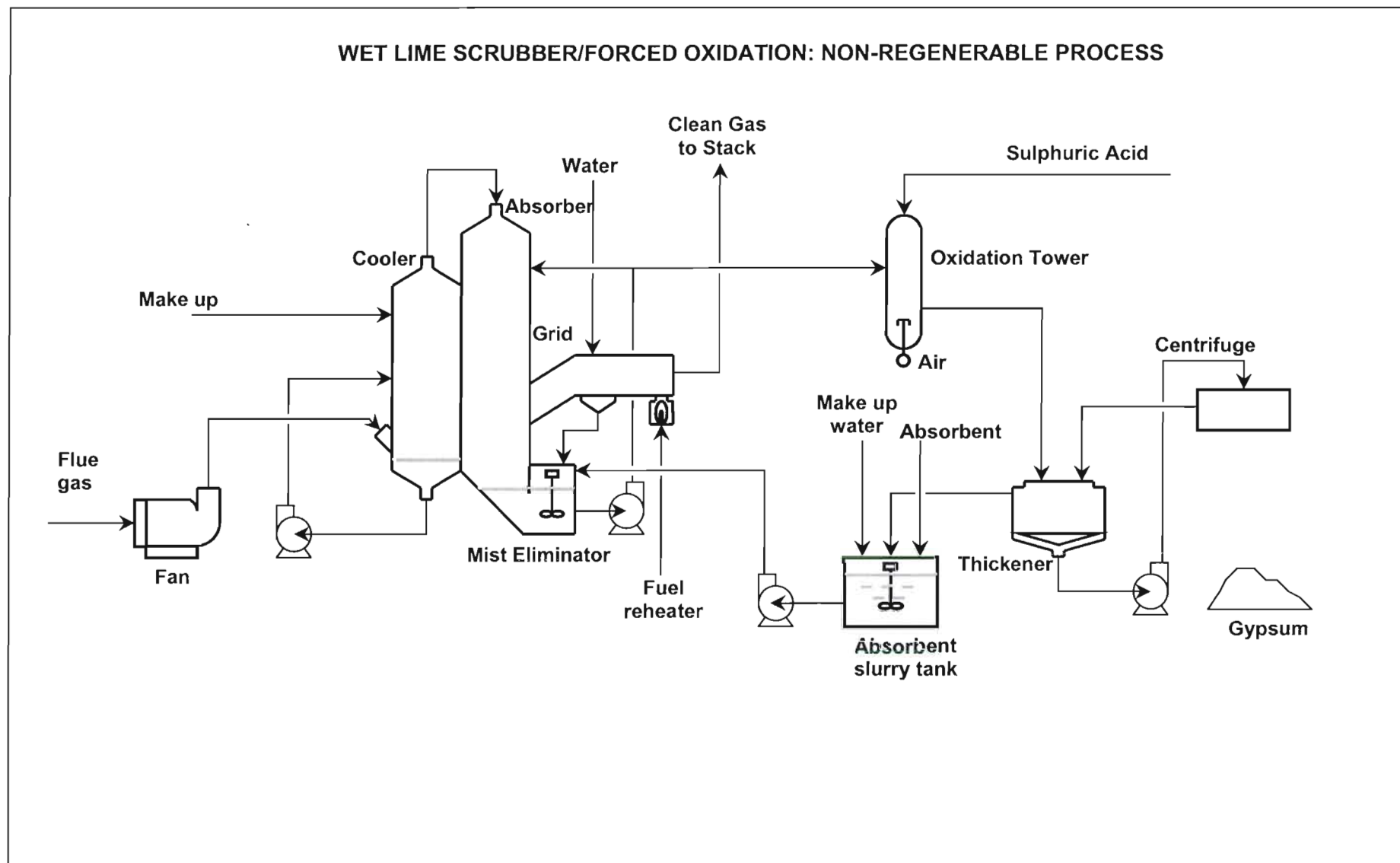
Through the use of an oxidation step, either as a part of the scrubbing process, or in a separate vessel, gypsum may be produced. Wet scrubbers can be divided into two oxidation processes which tend to influence the quality and type of waste product produced. Natural oxidation results in a mixture of calcium sulphite hemihydrate and calcium sulphate bihydrate. This mixture is difficult to dewater. A thickener is thus required and a secondary dewatering process utilizing centrifuges or filters is used. The waste products contains 40 to 50% water, which is mixed with fly ash or lime before disposing in a landfill (Rinaldi, 1995).

Forced oxidation produces a better quality gypsum and is therefore the process most widely used (Refer to Figure 2.3). Although two dewatering processes are still required,



**Figure 2.2:** KRC Limestone absorber process (Taken from Noell – KRC marketing brochure)





**Figure 2.3:** Simplified Flow Diagram of Mitsubishi Heavy Industries Wet Lime Scrubber/Forced Oxidation (taken from Klingspor and Cope, 1987)

dewatering is easier because the crystals are larger and therefore no thickener is necessary. The final product is 90% solid and can be sold for use as plaster, cement and wall board gypsum, thus reducing the overall costs (Rinaldi, 1995).

The absorber is equipped with a prescrubber, partly to provide additional particulate removal but mainly to cool and humidify the flue gas before it enters the absorber section. The prescrubber also removes significant portions of the halides (chloride and fluoride). The absorber removes sulphur dioxide with the limestone slurry.

### **Advantages**

- High removal efficiency of  $\text{SO}_2$  (80 to 90%), (Klingspor and Cope, 1987)
- Production of a saleable by-product.
- Decrease in capital cost.
- Increase in efficiency and reliability (95 to 98%).
- The gypsum produced is of a high quality.

### **Disadvantages**

- Increase in disposal cost of gypsum.
- Corrosion of material of construction due to the presence of chlorides in the absorber. Eskom flue gas has a chloride content of 3ppm (Smith, 1998).

### **2.4.2 Spray dry scrubbing (SDS)**

Spray dry scrubbers are generally found on medium capacity boilers (approximately 300MW) burning low to medium sulphur coal (<1.5%) (Bogucki and Pragnell, 1989). SDS has the ability to remove 99% of  $\text{SO}_2$  from the flue gas although they are generally designed for a 70 to 98% removal efficiency. The design efficiency is dependent on legislative requirements of the country and on the sulphur content of the coal. The greater the removal efficiency required, the more lime sorbent is required. See Figure 2.4 for a flow diagram of a spray dry FGD process.

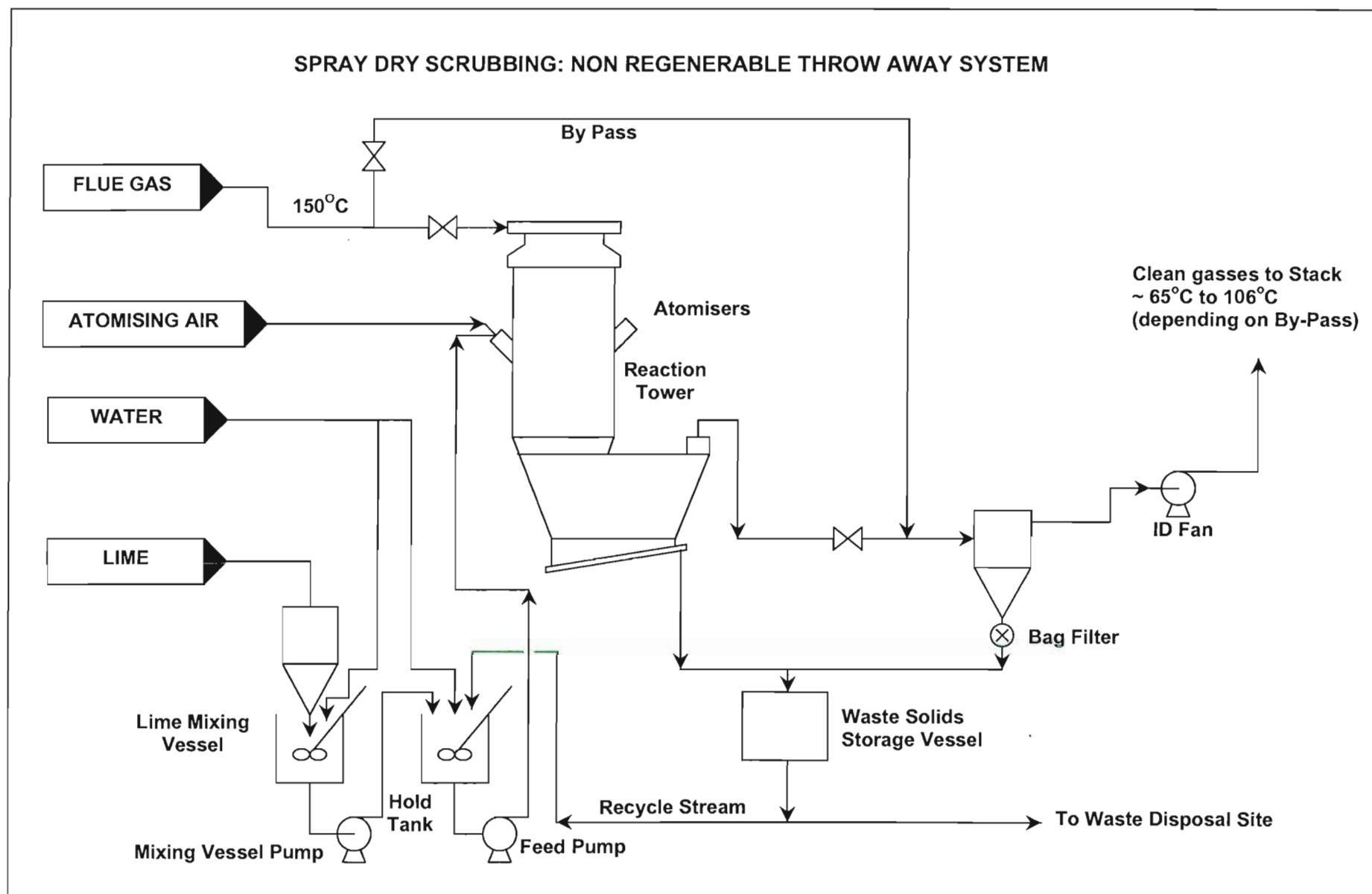
### **Process description**

The major plant components are:

- Spray dry absorber
- Fly ash removal system (fabric filter or ESP)
- Recycling/disposal device for the reaction product

The absorbents used in the SDS are mostly slaked lime and calcium oxides. The lime is mixed with water to form a concentrated slurry. In the spray dry absorber, the slurry

---



**Figure 2.4:** Flow Diagram of Lime Spray Dry FGD Process

is atomised by a rotary atomiser to mix intimately with the flue gases. Heat from the process evaporates the remainder of the moisture before the hydrated lime reacts with the sulphur dioxide ( $\text{SO}_2$ ), sulphur trioxide ( $\text{SO}_3$ ) and hydrochloric acid ( $\text{HCl}$ ). A residence time of approximately 10 seconds is required to allow the calcium sulphate/sulphite ( $\text{CaSO}_3/\text{CaSO}_4$ ) and calcium chloride ( $\text{CaCl}_2$ ) to form. There is no wastewater treatment required because the water is completely evaporated in the absorber. The process is dependent on the temperature of the flue gas, gas humidity,  $\text{SO}_2$  concentration and atomised slurry droplet size.

The removal efficiency of  $\text{HCl}$  and  $\text{SO}_3$  is much higher in SDS than in wet scrubbing, which may prove to be disadvantageous on Eskom plant because of the high resistivity ash. During the combustion process, sulphur in coal converts to  $\text{SO}_2$  and a small proportion converts to  $\text{SO}_3$  through the catalytic effect of bare metals. If sufficient  $\text{SO}_3$  is formed, it combines with moisture in the flue gas to form  $\text{H}_2\text{SO}_4$  which **condenses** onto the fly ash. This reduces the resistivity of the fly ash to an acceptable range for effective precipitation of the fly ash. Since South African coals are generally low in sulphur (<1.5%),  $\text{SO}_3$  production is minimal. To compensate for the low natural  $\text{SO}_3$  concentration, flue gas conditioning (FGC) with  $\text{SO}_3$  has been installed at Hendrina, Matla and Kriel Power Stations (Smith, 1998). Small quantities of  $\text{SO}_3$  (approximately 30 to 40ppm) are injected into the flue gas stream after the air heater to supplement the natural formation of  $\text{SO}_3$ . With the aid of FGC, efficiencies of the ESP has improved quite substantially (Bosch and Hansen, 1997).

The by-product is completely dry and contains unreacted lime. Part of the by-product is recycled and mixed with a fresh lime slurry to enhance lime utilisation. Some commercial plants have a prescrubber situated between the air heater and the absorber. This reduces the volume of waste product and often results in a decrease in erosion in the absorber.

Studies carried out at the Västerå Plant in Sweden found that the absorber nozzles had a life span of only 300 hours due to abrasion of the fly ash. This however was overcome by recycling a full complement of fly ash particle sizes rather than only the coarse particles. The nozzles are now lasting up to three years (Bogucki and Pragnell, 1989).

The SDS plant has a power consumption in the order of 0,5 and 1,0% of the electricity production. This is almost a third of the power consumption required by wet scrubbing. If Kendal Power Station is used as an example, auxiliary power consumption would be increased by 11,12GWh (at 1%) per annum (Technical Information, 1994). At a cost of 0,18c/kWh this equates to R2 million per annum as opposed to wet scrubbing costs of R5,37 million per annum.

## Problems

Although the plant has a reliability of 90%, many problems are present. Tank and pipe plugging causes the slurry rates to be slow. This can lead to the settling of lime in the pipework. Lime is hygroscopic and therefore absorbs moisture readily, resulting in hardened deposits. Corrosion of ESP and fabric filters is primarily due to operation within the dew point range. The rotary atomiser found in the Niro system requires stricter maintenance due to atomiser erosion and cracking as a result of the thermal shock of the cold slurry introduced into the heated flue gas.

## Waste disposal

The most common means of disposal is in landfills, old strip mines or gravel pits. The wastes must be conditioned with water and fly ash before disposal, because of the possibility of leaching of any hazardous components. Various investigations have been done to use the by-product in construction materials like cement and sand lime bricks.

### 2.4.3 Dual-alkali

Unlike the wet scrubbing process, the absorption of sulphur dioxide and production of waste are separated, with the addition of limestone/lime occurring outside the scrubber loop. This process utilises soda ash ( $\text{Na}_2\text{CO}_3$ ) for sulphur dioxide removal. The soda ash is regenerated for recycle by the reaction with slaked lime, and the gypsum produced is disposed of (Shannon, 1982). Due to the disposal of gypsum, this process is called a throw away system. Dual-alkali FGD system typically has  $\text{SO}_2$  removal efficiencies of up to 90%, with  $\text{NO}_x$  removal efficiencies expected to be very low (Shannon, 1982). Refer to Figure 2.5.

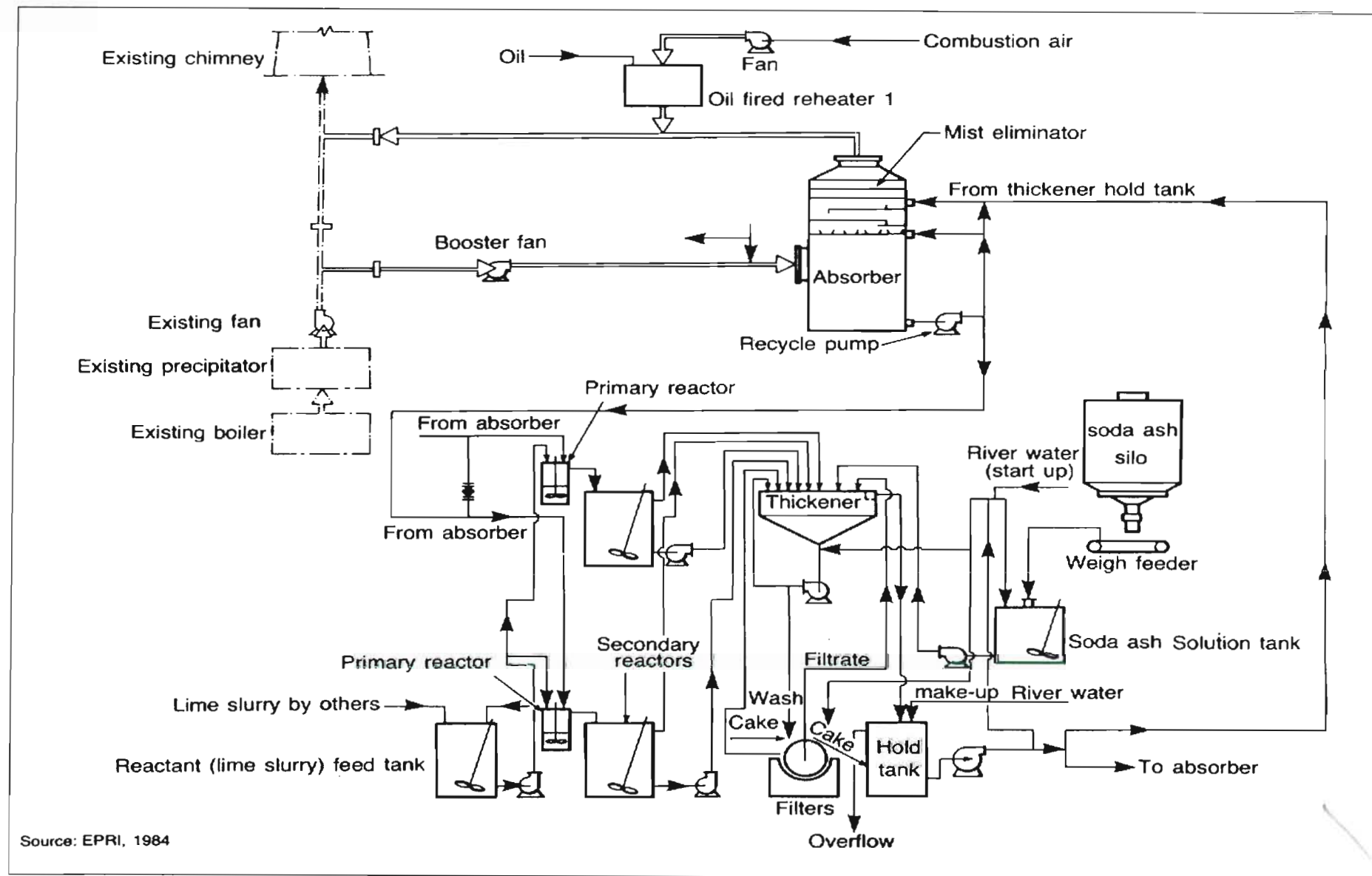
## Process Description

The soda ash ( $\text{Na}_2\text{CO}_3$ ) is slaked to give caustic soda ( $\text{NaOH}$ ). The caustic soda is used to absorb  $\text{SO}_2$  according to the following reaction:



Lime is slaked to give calcium hydroxide ( $\text{Ca}(\text{OH})_2$ ). The spent sorbent ( $\text{Na}_2\text{SO}_3$ ) is then reacted with the slaked lime to regenerate the sodium hydroxide. The regeneration reaction is:





Source: EPRI, 1984

Figure 2.5: Flow Diagram of Thyssen Dual Alkali Process (taken from Klingspor and Cope, 1987)

### Advantages

- Sulphur dioxide scrubber operation is less sensitive to pH control.
- Scrubber scaling and plugging problems are reduced.

### Disadvantages

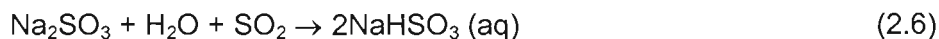
- More complex system, therefore increase in capital cost of construction.
- Potential increase in operating and maintenance costs.
- Non marketable wastes.

## REGENERABLE PROCESSES

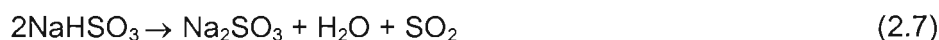
### 2.4.4. Wellman-Lord

The Wellman-Lord is the most common regenerable FGD system in operation today. It involves the absorption of sulphur dioxide in a sodium sulphite solution ( $\text{Na}_2\text{SO}_3$ ). The sulphite is converted to bisulphite during absorption and the solution is regenerated by thermal decomposition in an evaporator/crystalliser unit. Refer to Figure 2.6.

The primary absorption reaction is:



The regeneration reaction is:



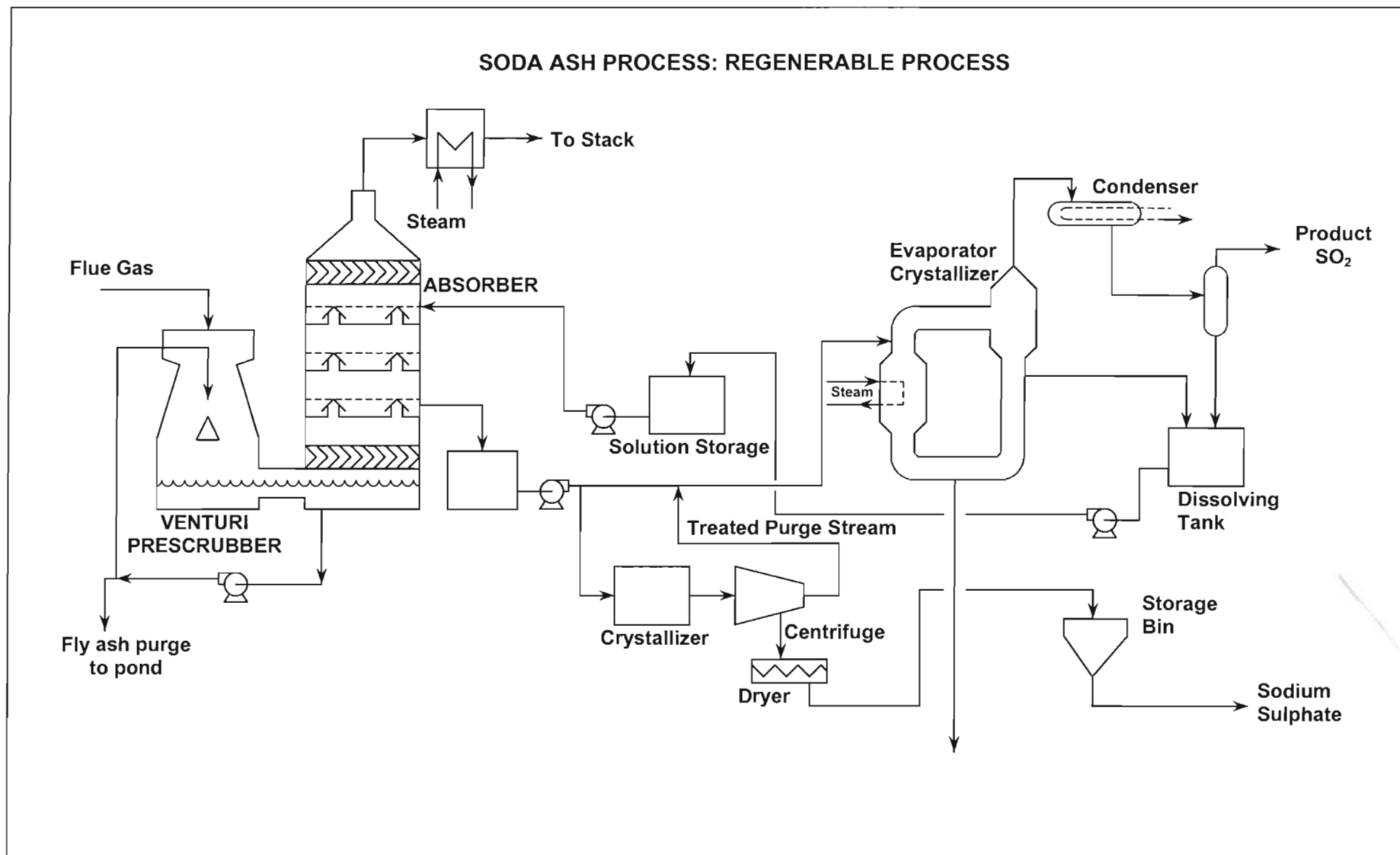
In the regeneration section the sulphur dioxide is recovered in a gas rich stream and may be used for the production of sulphuric acid, elemental sulphur, or liquid sulphur dioxide. A small portion of the spent absorption solution is selectively crystallised in the crystallisation section to remove any sodium sulphate that is formed during the absorption process

The following waste/effluents are produced from the Wellman-Lord process:

- Liquid effluent containing fly ash, chlorides and sulphuric acid from the prescrubber.
- Liquid effluent from the evaporator-crystalliser purge.

Disadvantages of the Wellman-Lord process:

- High energy demand in the regeneration section.
  - High loss in power plant efficiency due to the evaporation unit.
-



**Figure 2.6:** Simplified Flowsheet of the Wellman Lord Process (Adapted from Kohl and Reisenfeld, 1985)

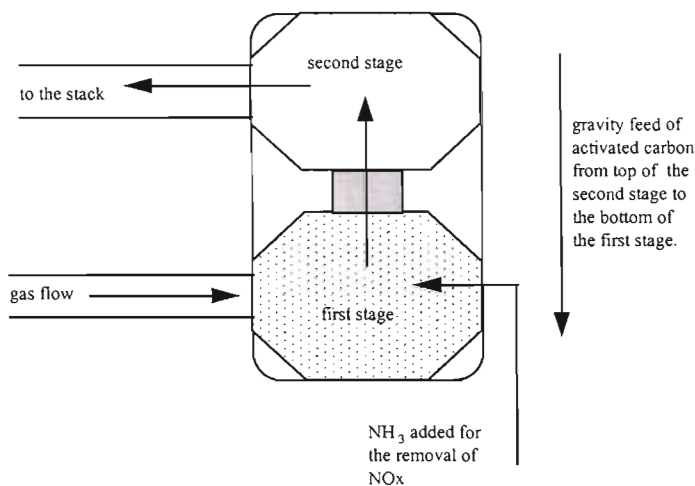


- High labour demand.
- High maintenance due to numerous control loops required.
- Complex plant thus giving rise to high construction costs.
- Process does not accommodate nitrogen oxide removal

### 2.4.5 Activated carbon

This process is based on the adsorption of sulphur dioxide on the surface of porous charcoal. The adsorbed sulphur dioxide reacts with oxygen and water vapour to form sulphuric acid. The activated charcoal is thermally regenerated and recycled to the adsorption reactor. Nitrogen oxide removal also becomes possible if ammonia is added to the process. See Figure 2.7.

Activated carbon FGD has 98% and 80% removal efficiency of  $\text{SO}_2$  and  $\text{NO}_x$  respectively (Klingspor and Cope, 1987). The reactor consists of two moving beds of activated carbon. The first bed removes the  $\text{SO}_2$  and the second bed removes  $\text{NO}_x$  if ammonia ( $\text{NH}_3$ ) has been injected to the process.

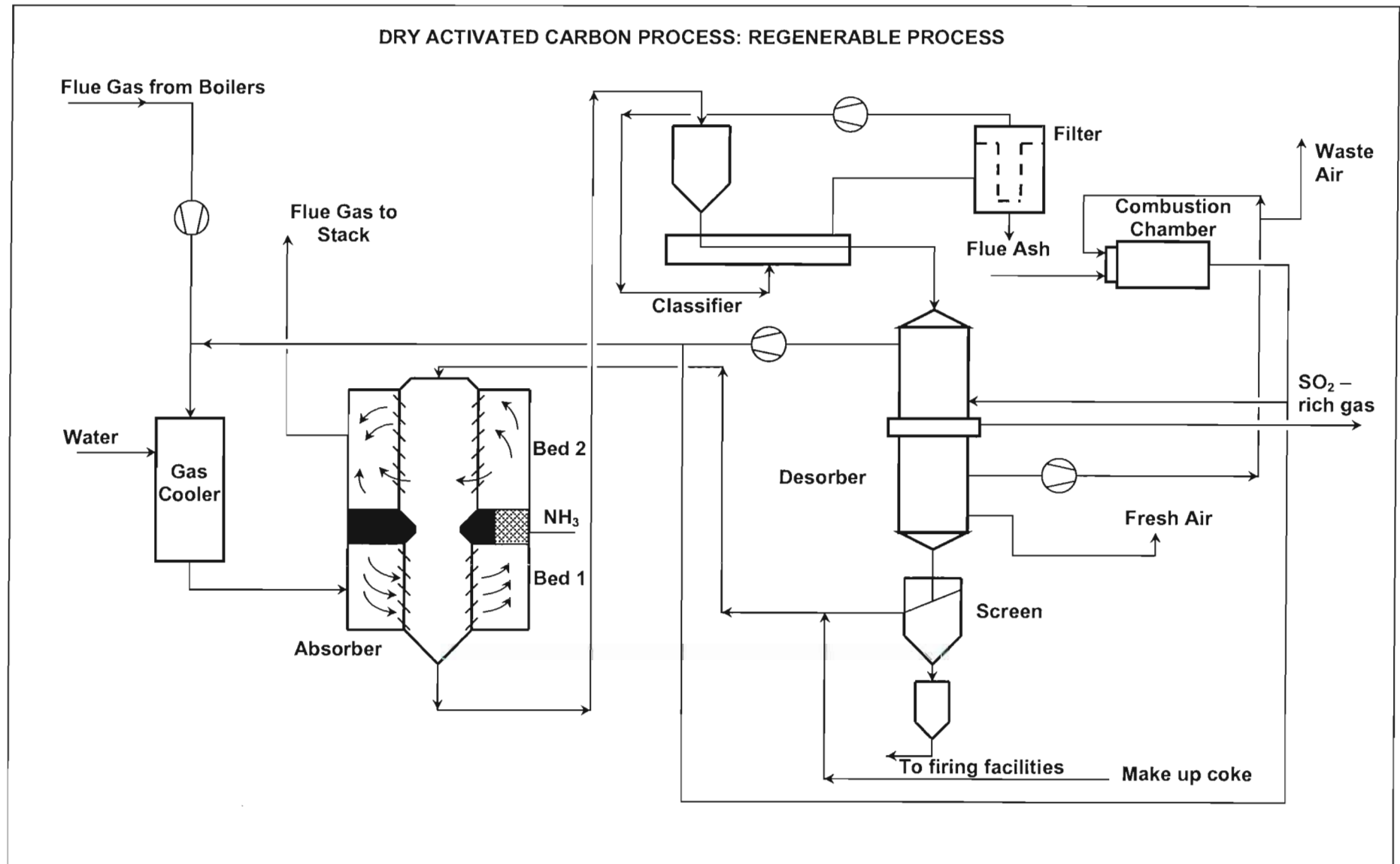


**FIGURE 2.7:** Bergbau-Forschung activated carbon adsorber for dual  $\text{SO}_x$  and  $\text{NO}_x$  removal

The adsorption process occurs at a temperature of  $120^\circ\text{C}$ . The regeneration process is performed at a temperature of  $400^\circ\text{C}$ . The regeneration step produces a sulphur rich gas stream which can be used to form sulphuric acid, elemental sulphur or liquid sulphur dioxide.

#### Advantages

- High removal efficiencies of  $\text{SO}_2$  and  $\text{NO}_x$ .
- No process water and waste water treatment required.



**Figure 2.8:** Simplified Flowsheet of the BERGBAU-Forschung/Uhde Dry Activated Carbon Process (taken from Klingspor and Cope,1987)

- Produces sulphuric acid which can be sold to offset costs.
- Also removes heavy metals and hydrocarbons.

### **Disadvantages**

- South Africa has very limited amounts of activated carbon. This means that the adsorbent will have to be imported at cost (Tilly, 1996).

### **2.4.6 Magnesium oxides**

The magnesium oxide (MgO) process is a wet system using hydrated magnesium oxide to remove flue gas sulphur dioxide in a spray tower. Refer to Figure 2.8. The sulphur dioxide reacts with the magnesium to form magnesium sulphate and small quantities of magnesium sulphite. The reaction products are removed from the scrubbing solution, dried and transferred to a regeneration facility. The regenerated magnesium oxide is recycled to the FGD system for reuse in the scrubbing process.

The dried magnesium sulphite and magnesium sulphate are regenerated in an oil-fired calciner in the presence of coke to enhance the reduction of magnesium sulphate. A sulphur dioxide rich gas stream is produced which may be further processed to produce elemental sulphur, sulphuric acid or liquid sulphur dioxide (Klingspor and Cope, 1987).

### **Advantages**

- Regeneration of the absorbent.
- MgO can be obtained from dolomite which South Africa has an abundant supply of. However, the recovery of MgO may prove to be costly.

### **Disadvantages**

- High construction costs due to the many unit operations.
  - High power consumption and running cost.
  - High labour demand.
  - Increased load on cooling towers (if absorber temperature control is to be obtained).
  - Expensive construction materials.
  - Liquid effluents containing fly ash, chlorides and sulphuric acid.
-

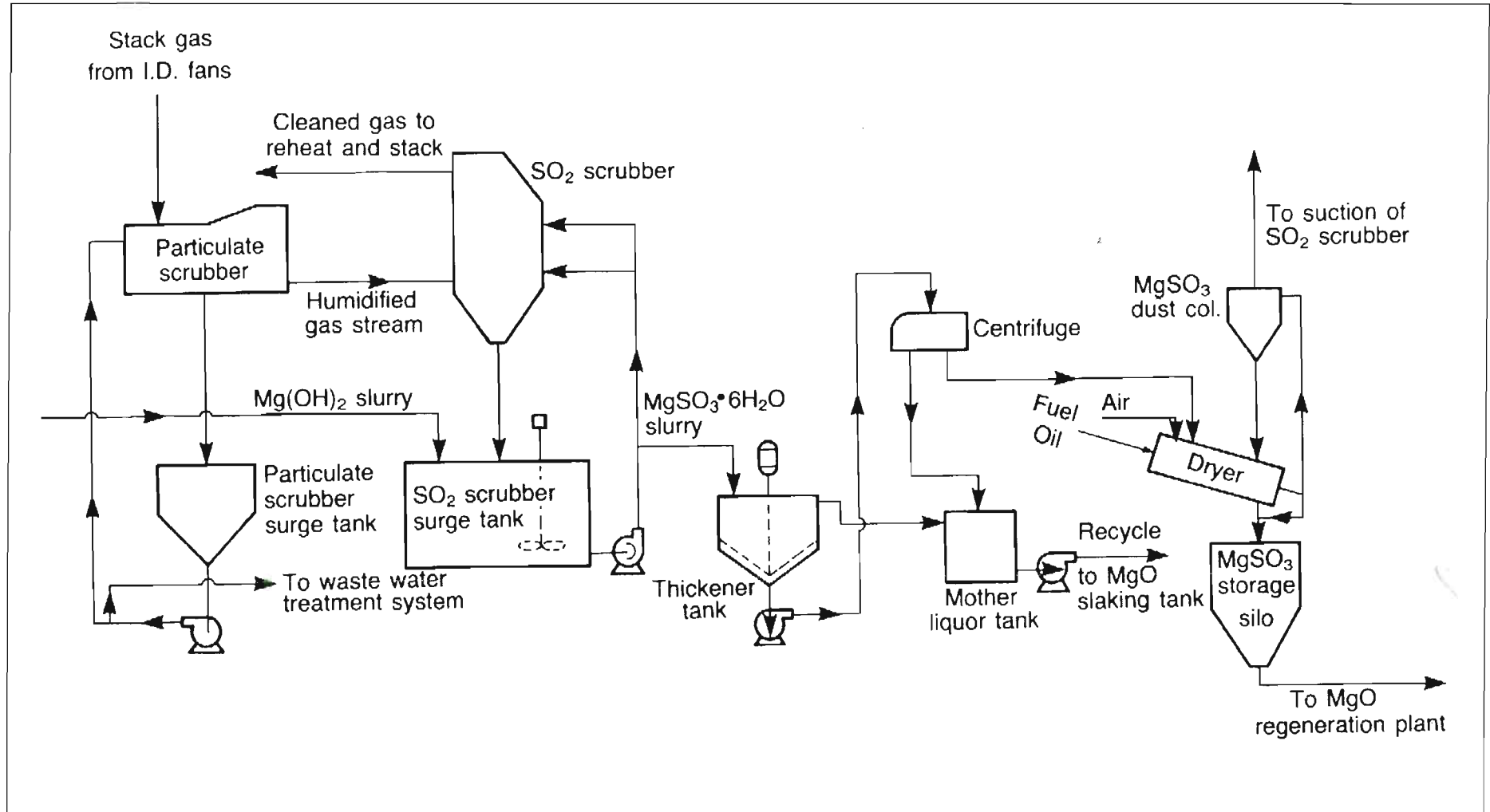


Figure 2.9: Simplified flow diagram of United Engineers and Constructors Wet Magnesium Oxide Scrubber (Klingspor and Cope, 1987)

## CHAPTER 3

### NITROGEN OXIDES

Nitrogen oxides from coal combustion are associated with plant damage, possible health effects, acid rain, and production of deleterious secondary pollutants. Nitrogen oxides directly affect health and are also transformed in combination with other air pollutants into harmful secondary pollutants, in particular, photochemical oxides. Some of the gaseous compounds include nitric oxide (NO), nitrogen dioxide (NO<sub>2</sub>), nitrogen oxide (N<sub>2</sub>O), nitrogen sesquioxide (N<sub>2</sub>O<sub>3</sub>), nitrogen tetroxide (N<sub>2</sub>O<sub>4</sub>), and nitrogen pentoxide (N<sub>2</sub>O<sub>5</sub>). The two oxides of primary concern in air pollution are NO and NO<sub>2</sub>. NO<sub>2</sub> is harmful in high enough concentration as well as being the primary constituent of secondary pollutants. It is heavier than air and is readily soluble in water, forming nitric acid and either nitrogen acid or nitric oxide, as indicated by the following equations:



NO<sub>2</sub> is a good absorber of energy in the ultraviolet range, therefore plays a major role the production of secondary air contaminants such as ozone (O<sub>3</sub>). Nitric oxide is emitted to the atmosphere in much larger quantities than NO<sub>2</sub>. It is formed in high temperature combustion processes when atmospheric oxygen and nitrogen combine. NO<sub>x</sub> under certain conditions produces photochemical ozone (O<sub>3</sub>), the brownish haze found in the polluted atmosphere, commonly referred to as smog. During photochemistry, NO<sub>2</sub> breaks down to produce atomic oxygen which reacts with oxygen molecules to form ozone.

#### 3.1 EFFECTS ON HUMAN HEALTH

Nitric oxide (NO) is a relatively inert gas and only moderately toxic. It leads to oxygen deprivation, because it has a stronger affinity for haemoglobin than oxygen does. NO is readily oxidised to NO<sub>2</sub>, which has biological significance.



NO<sub>2</sub> is an irritant to the respiratory tissues, the alveoli of the lungs, and causes chest pains, pulmonary edema and death. It also appears to reduce the body's resistance to infection.

Experiments were conducted on a group of volunteers by exposing them to 9.4 mg/Sm<sup>3</sup> (5 ppm) of nitrogen dioxide, considerably above the 7.0 mg/Sm<sup>3</sup> (3.75 ppm) peak recorded in Los Angeles, for 10 minutes produced a substantial but transient increase in the resistance of the lungs airways to air movement. Concentrations

from 47 to 141 mg/m<sup>3</sup> (25 to 75 ppm) cause reversible pneumonia. Nitrogen dioxide at high level exposures of 285 mg/Sm<sup>3</sup> (150 ppm) and above are fatal to humans (Stenby, 1990). A study in Chattanooga found a correlation between NO<sub>2</sub> exposure and reduced respiratory function in children, and a greater likelihood of acute respiratory illness (Wilson et al, 1980).

The main cause of concern is the contribution of nitrogen oxides to form more dangerous chemicals in the air. Nitrogen dioxide is converted in the atmosphere to nitrates and nitric acid. NO<sub>x</sub>, in the presence of sunlight and sulphur oxides, is a key contributor to photochemical production of sulphates in the air, which depletes the stratospheric ozone layer that shields the earth from the harmful ultraviolet radiation, thereby increasing the incidence of skin cancer.

### 3.2 NITROGEN OXIDES (NO<sub>x</sub>) CONTROL

The most popular choice of nitrogen oxide control is the installation of low NO<sub>x</sub> burners on large power generating units. The choice of NO<sub>x</sub> removal is generally dependent on the stringency of the legislation requirements. Selective catalytic reduction (SCR) and selective non-catalytic reduction (SNCR) methods can also be employed to reduce nitrogen oxide emissions.

The reassessment of furnace designs, upstream preparation of coal and modified boiler operation is found to further reduce NO<sub>x</sub> emissions. Detmarovice Power Station in Czechoslovakia achieved a 60% reduction in NO<sub>x</sub> emissions with the installation of low NO<sub>x</sub> burners, in conjunction with overfire-air, staging systems and flame monitoring. This process was modeled by IVO International (Ltd), Finland (Makansi, 1995).

#### 3.2.1 NO<sub>x</sub> Control by Combustion Modification

Foster Wheeler's low NO<sub>x</sub> burner (LNB) with advance overfire air is one of the promising new technologies being demonstrated in the Department of Energy's Clean Coal Technology Program for reducing nitrogen oxides (NO<sub>x</sub>) pollutants from coal fired utility boilers. NO<sub>x</sub> can contribute to smog and acid rain especially in the urban areas.

Advanced overfire air involves:

- Improving the mixing of over-fire air with the furnace gases to achieve complete combustion.
- Depleting the air from the burner zone to minimise NO<sub>x</sub> formation.
- Supplying air over furnace wall tube surfaces to prevent slagging and furnace corrosion.

The advanced over-fire air technique is expected to reduce NO<sub>x</sub> emissions by about 35 %. In a LNB, fuel and air mixing is controlled to prevent the formation of NO<sub>x</sub>.

This is accomplished by regulating the initial fuel-air mixture, velocities, and turbulence to create a fuel-rich flame core and by controlling the rate at which additional air is required to complete combustion. Tests conducted indicate that LNB technology is capable of reducing NO<sub>x</sub> emissions by about 45 % (Power-Gen. International, 1998).

Based on earlier experience, the use of advanced over-fire air in conjunction with LNB can reduce NO<sub>x</sub> emissions by as much as 65 % as compared with conventional burners.

Commercialisation of this technology will be aided by the following characteristics:

- Reduced NO<sub>x</sub> emissions by as much as 65 %.
- Competitive capital and operating costs.
- Relatively easy retrofit.
- Little or no derating of the boiler.
- Use of commercially available components.
- Automatic control of boiler efficiency.
- Maximum pollution abatement through the use of artificial intelligence technology.

This technology is currently being employed mainly in the United States, but given the many advantages of this technique, it will not be too long before South Africa follows suit.

### **3.2.2 Selective Catalytic Reduction**

Selective catalytic reduction (SCR) technology is the more common, but also the most expensive of all the NO<sub>x</sub> reducing systems. The SCR process uses ammonia (NH<sub>3</sub>) to reduce NO<sub>x</sub> to nitrogen and water vapor. The ammonia is vaporized and injected into the flue gas upstream from a catalyst. The purpose of the catalyst is to improve the efficiency for the reduction of NO<sub>x</sub>. Commonly used catalyst for this process is vanadium and titanium compounds. The catalysts are deposited on metallic or ceramic parallel-plate substrates. Alternatively, they are impregnated on a honeycomb substrate (Khan et al, 1990).

There are three types of SCR plants in operation. These include:

- High dust installations, situated between the boiler and the ESP.
- Low dust installations, situated downstream from the ESP.
- Tail end installations, situated downstream from the FGD plant.

Most of the information obtained for SCR processes is based on existing plants in Japan, that typically burn coal that has a sulfur content of less than 1% (Khan et al, 1990). Eskom power station burn coal with a sulphur content of 0.5% to 1.5% (average 1%) (Tilly, 1996). This is very similar to the Japanese.



Major differences exist between high dust installations and tail end installations because of the structural and operational differences.

### **Tail end installations**

- Requires heat exchange equipment to reheat the flue gas to the SCR operating temperature, approximately 400°C. This results in an increase in capital costs.
- These plants require one third less catalyst than the high dust installations, since the catalytic plant is situated in a “clean” flue gas environment.
- The advantage of this system is that the catalyst plant can be located such that it minimizes construction time.
- A cheaper catalyst can be used because of the cleaner location.
- Long catalyst life of some 10 years.

### **High dust installations**

- Does not require heat exchange equipment. The SCR operates at low loads and lower flue gas temperatures therefore requires an economizer by-pass system to be installed. This increases the initial capital outlay.
- The location of the plant may cause problems in retrofit situations because of the limited space between the boiler and the ESP. This will result in higher construction times and cost.
- Short catalyst life of 3 to 5 years before it has to be replaced.
- Additional layers are installed to increase the life span of the catalyst. As the catalytic action deteriorates, the additional layers are used, until the full life span of the catalyst is reached.
- Installation of corrosion resistant material.

### **3.2.3 Lurgi Circulating Fluidized Bed Reactor Process**

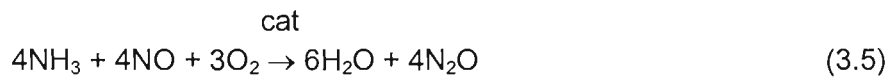
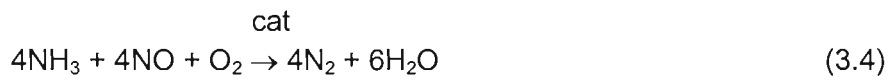
This process involves the simultaneous removal of sulphur dioxide and nitrogen oxides by a circulating fluidized bed reactor (Sauer et al., 1989), (Sauer and Anders, 1989). Refer to Figure 3.1.

#### **Process Description**

The flue gas is heated to 400°C prior to the entrance of the fluidized bed. The flue gas fluidizes catalyst particles of size 500  $\mu\text{m}$  and the hydrated lime ( $\text{Ca}(\text{OH})_2$ ) particles. The catalyst particles consist of  $\text{Al}_2\text{O}_3$  with active compounds  $\text{FeO}/\text{FeSO}_4$  and  $\text{MnO}/\text{MnSO}_4$ .

The alumina based particles catalyses the  $\text{deNO}_x$  reaction by the addition of  $\text{NH}_3$  to reduce  $\text{NO}_x$  to  $\text{N}_2$ . The primary reactions are:

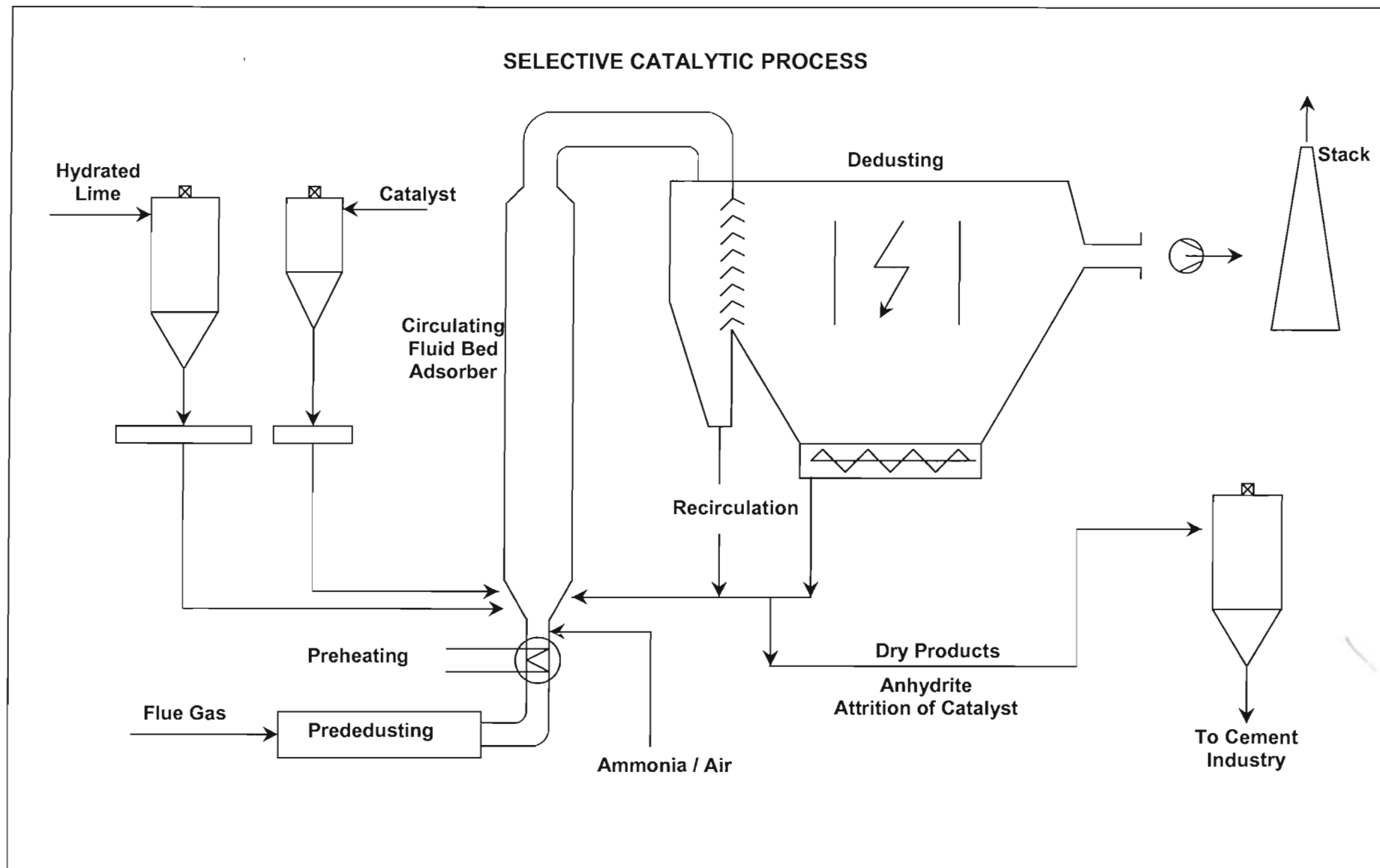




MnO/FeO catalyst has a dual oxidizing reducing effect. As the  $\text{NO}_x$  is reduced, the  $\text{SO}_2$  is oxidized by the reaction:



The lime that is fed to the reactor reacts with  $\text{SO}_3$  to form  $\text{CaSO}_3$  and  $\text{CaSO}_4$ . The catalyst and lime is elutriated from the reactor and separated in the ESP unit. The catalyst is recycled to the fluidized bed reactor. The fines which are  $\text{CaSO}_3$ ,  $\text{CaSO}_4$ , unreacted sorbent, ash and attrition of catalyst, are discarded.



**Figure 3.1:** Simplified flow diagram of the Lurgi Circulating Fluidised Bed Reactor Process (taken from Sauer et al., 1989)

## CHAPTER 4

### PARTICULATE (FLY ASH) EMISSIONS

The prevailing national policy is to wash and export the higher quality coals, yielding typically 7 to 8% ash, while the poorer quality coal is burnt in Eskom power stations. This ash content varies between 15 to 50% ash, with an average of about 25% (Tyson et al, 1988).

Ash is emitted from the pulverised fuel fired boilers in the form of fly ash or pulverised fuel ash (PFA), a fine incombustible with particles ranging from sub to 50  $\mu\text{m}$  in size, the majority falling in the 10 to 20  $\mu\text{m}$  range. Some 15 Mton of PFA leave Eskom power station boilers each year, and the considerable impact this would have on our environment is attenuated by the installation of Electrostatic Precipitators (ESP) (Smith, 1998).

Precipitators are not 100% efficient, a quantity of dust will not be collected, but will pass through the ESP, and be discharged via the chimney to the atmosphere. In Eskom's older stations the ESP had relatively low efficiencies. Today, however, considerably higher efficiencies are being specified. At Arnot power station the electrostatic precipitators have been designed such that units 1 to 3 have 3 field ESP's with an efficiency of 98% and units 4 to 6 have 2 field ESP's with an efficiency of 96%. Thus, 2% to 4% of the fly ash escapes to the atmosphere (Smith, 1998).

The sulphur content of the coal has a direct and marked effect on the electrical resistivity of the fly ash produced. Coal with a low sulphur content will yield a high-resistivity ash, which builds up on the collector plate as an insulating barrier layer. The electrical charge on the particles does not leak off, and the electrical field between discharge and collector is reduced, with accompanying loss of efficiency. This problem exists at many of Eskom's stations, with the result that even the relatively low specified efficiencies could not be attained in practice.

Studies were carried out to determine the most effective technique to improve performance of existing precipitators. These included the electrical modification of the waveforms of the DC voltage applied to the discharge electrode and injection of chemical conditioners (ammonia, tri-ethylamine and sulphur trioxide) into the flue gas to modify the characteristics of the PFA particles (Smith 1998). Sulphur trioxide proved to be cost effective, and most effective.

## 4.1 USES OF FLY ASH

Fly ash is used as a partial replacement for cement in concrete, concrete products and grouting applications. By replacing a portion of the cement with fly ash (usually 15% to 30%), raw material costs are reduced, while the ultimate strength and quality of the concrete are improved (Dacey and Cope, 1986).

The advantages of fly ash in concrete are (Jacobs, 1986):

- Lower cost
- Improved workability
- Greater ultimate strength
- Lower permeability
- Less thermal cracking
- Increased sulphate resistance
- Greater pumpability
- Improved finishing characteristics

Fly ash is also used as a fill material and in soil reclamation. By using fly ash as a fill material, poor load bearing soils and uneven terrain can be converted into useable land. In the soil reclamation process, the fly ash is mixed with very acidic soils. This technique adjusts the pH of the soil and allows vegetation to grow, and is found to be most useful in reclaiming surface-mined areas. Further, because fly ash is a by-product from coal combustion and does not have to be specifically manufactured, soil improvements can be made at a much lower cost than with most other alternatives.

Additional fly ash applications include (Ballinger Energy Series, 1979):

- Lightweight aggregate
- Paints and coatings
- Soil conditioner in agriculture
- Plastics
- Filter aid in the vacuum filtration of industrial sludges
- Sludge fixation and stabilization
- Coagulants in sewage treatment
- Asphalt roofing and sliding materials

## 4.2 EFFECTS ON HUMAN HEALTH

Particulates within the size range of 1 to 2  $\mu\text{m}$  in size are retained in the bronchioles and alveoli, which are sites of gas exchange in the lung. These smaller particles settle and diffuse onto the lung wall. Many toxic elements, including heavy metals, have been found to concentrate on these smaller particles because of their high surface-area-to-

---

volume ratio. Therefore, the small particles that penetrate deep within the lungs will also concentrate toxic elements.

Wilkelstein (1996) in a study found a strong positive association between suspended particulate levels and gastric and prostatic cancer and respiratory diseases. He found a weaker association with cardiovascular diseases, and a strong suspicion of association between pollution and cirrhosis of the liver. Lave and Seskin, 1995, found a significant correlation between particulate concentration and total mortality, and found particulates to be the pollutant most closely associated with the death of infants under one year of age.

Coal dust is considered a hazard because it can cause black lung disease, medically known as Coal Workers Pneumoconiosis. This disease is primarily predominant in mine workers and their families. Some of the signs of black lung disease are difficulty in breathing, focal emphysema, progressive scar tissue formation (fibrosis), heart failure and, increased risk of rheumatoid arthritis with some form of black lung disease i.e. Caplan's Syndrome (Stenby et al, 1990). Table A2.1 shows a summary of acute and chronic effects of particulates.

#### **4.3 EFFECTS ON PLANTS AND ANIMALS**

Unfortunately, very little information is available on the detrimental effect of particulates on vegetation. Coal dust in the presence of moisture inhibits growth of plant tissue. Dust coating on leaves reduces photosynthesis, since no sunlight is able to penetrate this layer. The increased plugging of the stomata reduces plant growth. Animals that eat plants covered with particulate may suffer some ill effects.

Particulate matter can damage materials by soiling clothing and textiles, corroding metals in the presence of moisture, eroding building surfaces, and discolouring and destroying painted surfaces. For example, at particulate matter concentration ranging from 130 to 180  $\mu\text{g}/\text{m}^3$  and in the presence of  $\text{SO}_2$  and moisture, corrosion of steel and zinc is 3 to 4 times greater than in an area with particulates of 60  $\mu\text{g}/\text{m}^3$  (Hutton and Allen, 1982).

#### **4.4 REMOVAL OF PARTICULATE (FLY ASH) EMISSIONS**

During the combustion of coal, large quantities of particulates are produced. Therefore, it is very important to have a reliable, low cost removal of particulates from the flue gas. There are three main types of equipment available for reliable, low cost removal of particulates for the flue gas. These are:

- Cyclones and multicyclones
  - Bag filters
-

- **Electrostatic precipitators (ESP).**

A summary was compiled of the characteristic for the particulate removal equipment. This can be found in Table A2.2 (Appendix 2). This was done to provide a comparison of the different technologies, and to determine the most reliable and economic option available for particulate removal. This is examined below.

#### **4.4.1 Cyclones and Multicyclones**

From the above table it can be concluded that where possible cyclones and multicyclones should be used as they provide a simple, low cost means of removing particles. However, under characteristic, effect of coal type, it can be seen that cyclones/multicyclones are relatively ineffective for collecting small particles (less than 5 to 10  $\mu\text{m}$ ) and with the more stringent laws and regulations being introduced today, cyclones/multicyclones may not be the most effective means of particulate removal. Also, the presence of moisture in the ash causes blocking, thereby reducing the efficiency of the equipment.

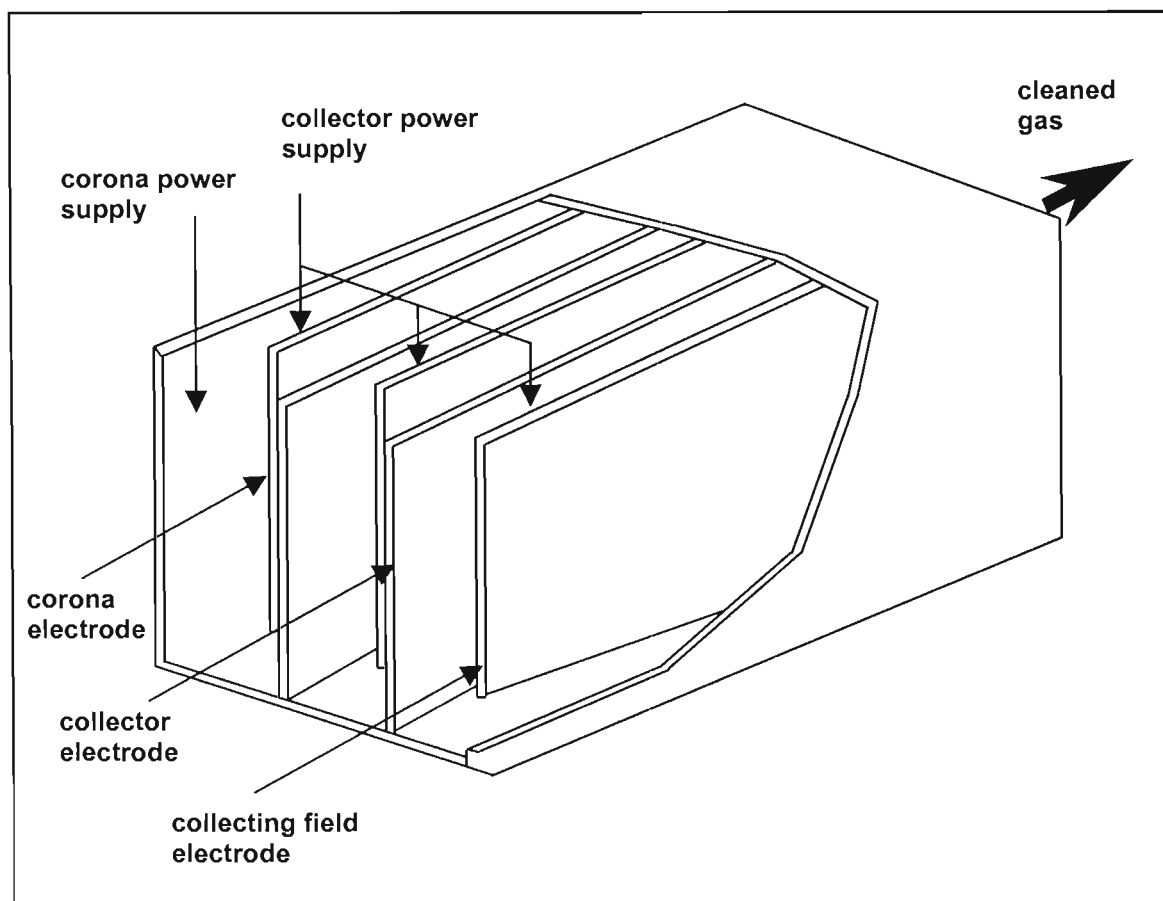
#### **4.4.2 Baghouse Filters**

Baghouse filters are widely used throughout the world owing to their high collection efficiency (approximately 99.5 %) and their ability to capture very small particles (less than 1  $\mu\text{m}$ ) (Dunderdale, 1990). The selection of the fabric for the bag filter is very important. The fabric must be resistant to the high operating temperatures and acidity of the ash, otherwise there could be the potential of a fire hazard. The presence of moisture in ash causes caking and this will tend to affect the performance of the filter. Release of the collected dust cake can be achieved by reversing the flow or applying pulsed jets of compressed air at regular intervals. Due to the performance being affected by the coal sulphur content, it is recommended that coals with low sulphur content be used. In the event of this being unavoidable, sorbents can be used to capture the sulphur. Despite the high capital and moderate operating cost, it may prove to be more beneficial in the long term to use baghouse filters capable of removing very small particles, than using cyclones, and later on having to install more control equipment to meet present emission standards.

#### **4.4.3 Electrostatic Precipitators (ESP)**

Most power stations use ESP. This is because it is the most efficient and effective method to collect small particles. The performance of the ESP is best when high sulphur coal is used. This gives power plants the advantage of using coals that may or may not require desulphurisation. The only drawback of the EPSs, is that their performance depends critically on the ash resistivity. Coal with a low sulphur content will yield a high-resistivity ash. This problem can however be overcome by conditioning the flue gas with sulphur trioxide.

---



**Figure 4.1: Schematic diagram of an Electrostatic Precipitator**

#### **4.4.4 Conclusion**

Fabric filters have high removal efficiencies, are capable of removing fine particulate matter and most importantly, provide a long-term solution to maintaining particulate emission levels below the legislative limits. In South Africa, this limit has been set to  $50\text{mg}/\text{Sm}^3$  by the Chief Air Pollution Control Officer (CAPCO) in liaison with the Department of Environmental Affairs and Tourism (DEAT) (Presidents Council, 1990). It is therefore the preferred option, however, most fossil fired power stations already in operation use ESP for particulate removal. The cost of retrofitting existing stations with fabric filters would prove too expensive, although fabric filters are more beneficial with regards to long term removal of particulates. All new power stations should install fabric filters for particulate removal. Existing power stations should improve ESP performance by flue gas conditioning using sulphur trioxide as the chemical conditioning mechanism.

#### 4.4.5 Flue Gas Conditioning by Sulphur Trioxide

The progressive improvement of particulate emission standards has resulted in the occurrence of occasional emissions in excess of the limits set, mainly since many electric utilities have chosen to switch to low sulphur coals to reduce sulphur dioxide emissions. It has therefore become necessary to increase the collection efficiency of the power station flue gas precipitators in order to meet more stringent emission levels.

Flue gas conditioning (FGC) is used to improve electrostatic precipitator (ESP) performance by reducing the fly ash resistivity and increasing the cohesivity of the fly ash. The collection efficiency of an ESP depends on the chemical and physical properties of the fly ash, sulphur trioxide generated from the coal borne sulphur, excess air and the flue gas temperature. One of the major factors affecting the performance of an ESP is the electrical resistivity of the fly ash. The electrical resistivity of fly ash typically increases as the ratio of sulphur to ash content decreases, assuming other variables such as temperature, pressure remain constant. If the resistivity is too high, which often occurs when low sulphur coal is burned, the collection efficiency is poor because the amount of useful power that can be supplied to the ESP is limited, mainly by the layer on the collecting plate. Corona current from the discharge electrodes pass through the collected dust layer on the plates to reach the grounded collecting plate surface.

##### Chemical conditioning mechanism

Sulphur trioxide is the most widely used form of FGC. The sulphur trioxide concentration is determined by the kinetics of sulphur oxidation and is typically limited to a small fraction (1 to 2%) of the total sulphur in the coal.

##### The process

The  $\text{SO}_3$  injected into the raw gas ducts is completely adsorbed by the fly ash dust particles, thereby reducing the dust resistivity to within a readily collectable resistivity range.

The amount injected is from 10 to 30 ppm vol (typically 20 ppm vol) into the gas-stream just downstream of the air heater (which already contains about 1 000 ppm vol due to the combustion of the natural sulphur in coal). The  $\text{SO}_3$  is hydrolysed to  $\text{H}_2\text{SO}_4$ , and the sulphate ions adhere to the dust particles, and hence are collected on the precipitator plates with no net increase in sulphate emitted. The overall sulphate emission is usually reduced as more of the dust (carrying some of the inherent sulphate) is retained within the precipitator.

---



The resultant combination of flue gas precipitator and gas conditioning plant offers an economical solution without prejudicing the acknowledged precipitator advantages of low maintenance, low energy consumption, long service life, and maximum plant availability. The collection performance of existing precipitators can thus be improved so that operation becomes largely independent of coal quality.

## CHAPTER 5 FUTURE TECHNOLOGIES

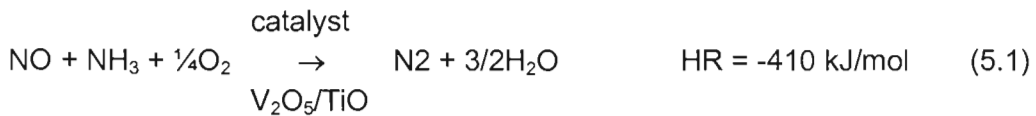
### 5.1 TECHNOLOGY FOR SO<sub>2</sub> AND NO<sub>x</sub> REMOVAL

This process is referred to as the SNOX (SO<sub>2</sub>, NO<sub>x</sub>) removal process. The SNOX process uses two catalytic reactors situated sequentially for the removal of sulphur dioxide (SO<sub>2</sub>) and nitrogen oxides (NO<sub>x</sub>). Ammonia is used to reduce NO<sub>x</sub> catalytically to nitrogen (N<sub>2</sub>) and water. SO<sub>2</sub> is oxidised to SO<sub>3</sub> in the SO<sub>2</sub> converter after which it condenses as 95% pure sulphuric acid (H<sub>2</sub>SO<sub>4</sub>) (Livengood and Marussen, 1992). The H<sub>2</sub>SO<sub>4</sub> produced is a saleable by-product, therefore there are no waste disposal costs. The removal efficiencies for SO<sub>2</sub> and NO<sub>x</sub> are 97% and 95% respectively.

#### 5.1.1 Process Description

The flue gas is passed through the baghouse filter which is used to remove fly ash. The flue gas blowers pass the flue gas to the heat exchangers where it is heated to about 400°C with the gases from the sulphur dioxide converter. Refer to Figure 5.1.

After an air rich ammonia mixture is injected into the stream, the flue gas is passed through a selective catalytic reduction reactor where the following reaction occurs:



The gas is heated further by another 20°C before entering the SO<sub>2</sub> converter where SO<sub>3</sub> is formed. The reaction is



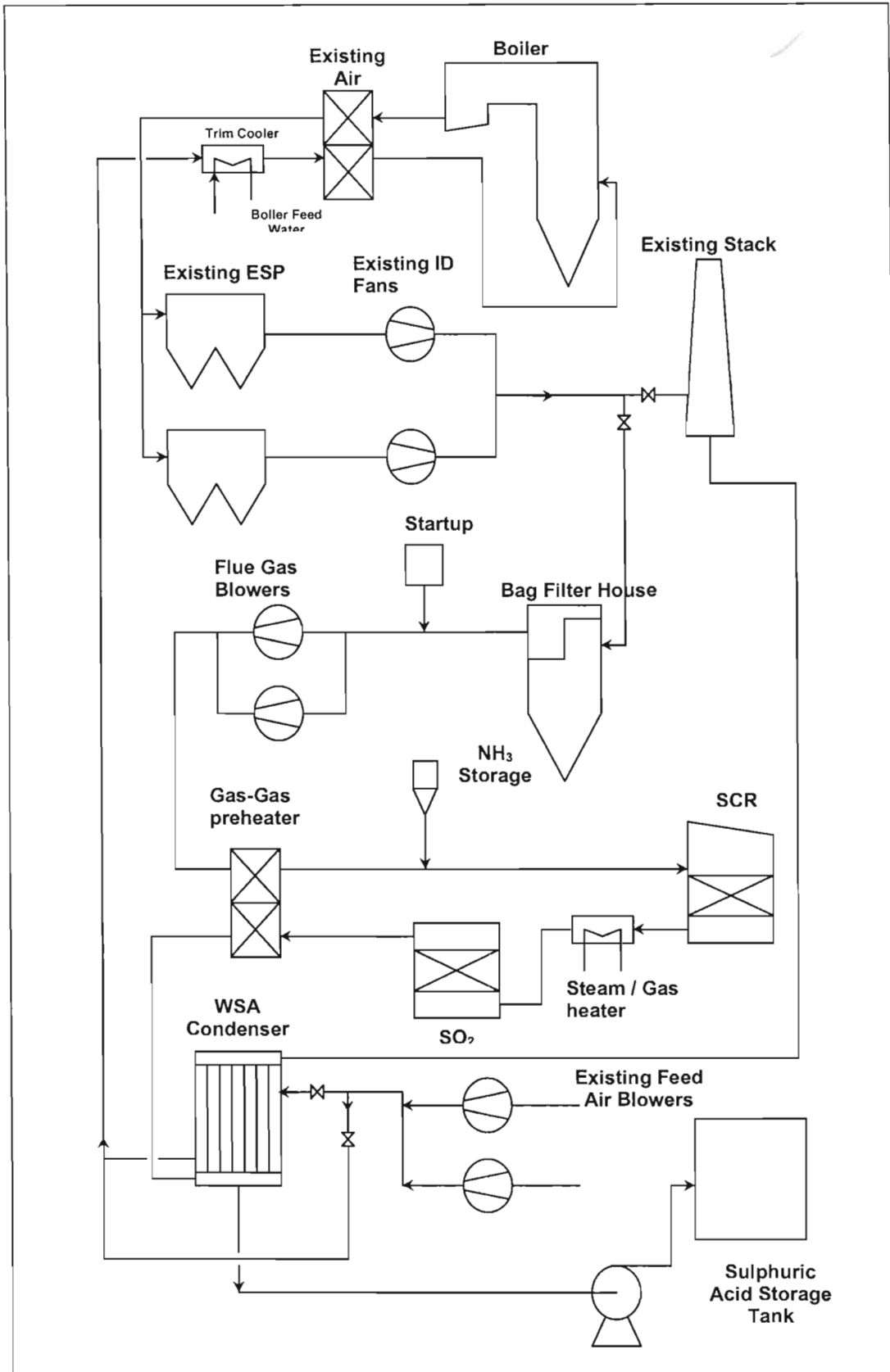
The hot gases are cooled by heat transfer and during this cooling process SO<sub>3</sub> reacts with water as follows:



The sulphuric acid is condensed out as 95% sulphuric acid.

#### 5.1.2 Advantages

- The NO<sub>x</sub> removal efficiency is better than a conventional SCR plant situated behind a conventional FGD scrubber. The main reason being, that the plant can operate with a molar NH<sub>3</sub>/NO<sub>x</sub> ratio of greater than one, because the surplus NH<sub>3</sub> is removed



**Figure 5.1:** Flow Diagram of the SNOX process (Andreasen et al, 1992)

by the SO<sub>2</sub> converter down stream from the SCR. Conventional SCR uses a ratio of less than one, in order to eliminate ammonia slip, which must be below 5ppm.

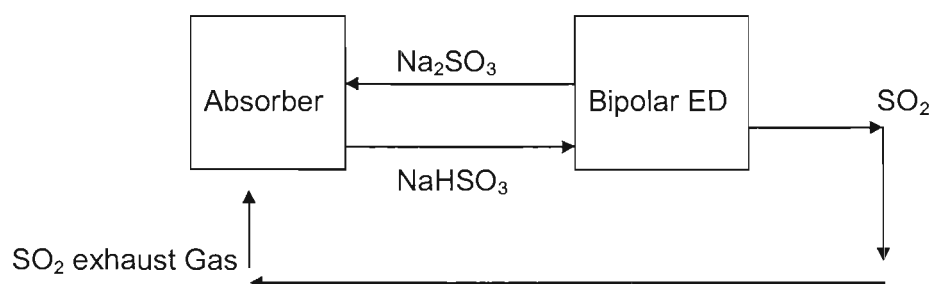
- Scaling problems generally associated with SCR are reduced, due to operation above the dew point temperature of ammonia sulphate or bisulphate.
- Reduced power consumption (0,2% of electricity production).
- Recovery of thermal energy.

## 5.2 ELECTROMEMBRANE DESULPHURISATION

Dr Vladimir Linkov of the Sasol Centre for Chemistry, Potchefstroom University for CHE, conducted investigations into Electromembrane desulphurisation (Tilly and Keir, 1995). The report on the economic evaluation has been included in Appendix 4. The aim of the research was to do a feasibility study of membrane desulphurisation. The economic evaluation has shown that the membrane technology is very cost effective.

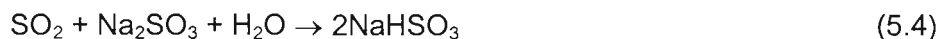
### 5.2.1 Process Description

For this process, the second part of the Wellman-Lord method was modified (Linkov, 1995). The thermal decomposition of sodium bisulphate was replaced by a electromembrane conversion in a double-flow electrodialyser equipped with a bipolar membrane. A schematic diagram of the process is shown in Figure 5.2.



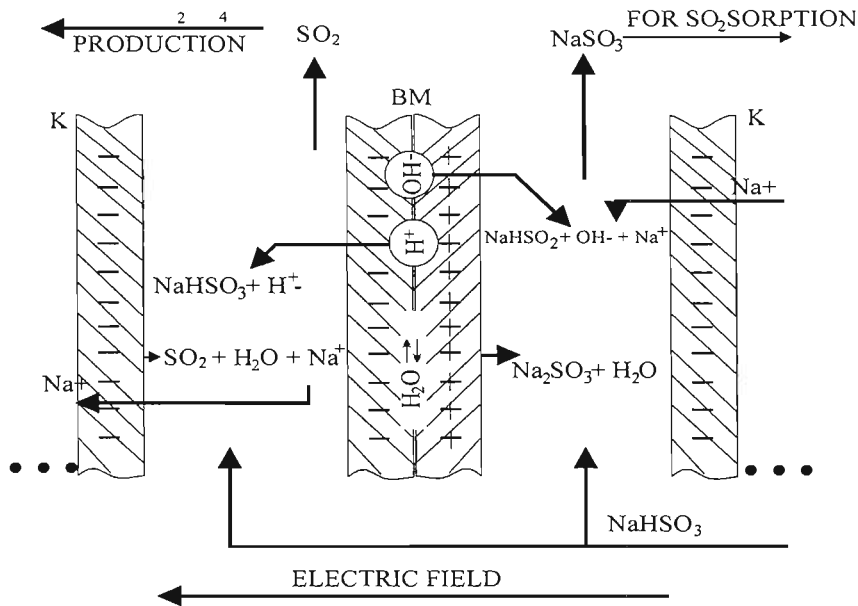
**Figure 5.2:** Schematic diagram of electromembrane sulphur dioxide treatment.

In this process, SO<sub>2</sub> is sorbed by a sodium sulphite solution in an absorber, according to the following equation:



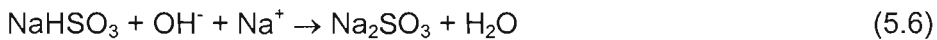
Sodium bisulphate flows to a bipolar electrodialyser (See Figure 5.3). At the cation exchange side of the bipolar membrane, hydrogen ions are generated. SO<sub>2</sub> evolves from the solution:





**Figure 5.3:** Schematic diagram of electrodialysis cell with bipolar membrane.

At the anion-exchange side of the bipolar membrane, hydroxyl ions and sodium bisulphate react to form sodium sulphite:



The cycle is completed by the use of  $\text{Na}_2\text{SO}_3$  for  $\text{SO}_2$  absorption.

The catalyst is used on the boundaries between the cation and anion exchange parts of the bipolar membrane, for the generation of hydrogen and hydroxyl ions.



For current densities higher than  $100\text{A/m}^2$ , the current yield of hydrogen and hydroxyl ions is constant and equal to 0.95:

$$\eta_{\text{H}^-, \text{OH}^+} = 0,95 \quad (5.8)$$

### 5.2.2 Process Equations

The following equations describe the various processes of the bipolar membrane

- The voltage decrease across a MB-3 bipolar membrane can be described by the equation:

$$U_{MB} = 0,66 + i/\kappa_{MB} \quad (5.9)$$

where:  $i$  = current density;  
 $\kappa_{MB}$  = surface electroconductivity of bipolar membrane

- Surface electroconductivity of the bipolar membrane:

$$\kappa_{MB} = 0,012 \exp[(1/T - 1/293)(-14200/R)] \quad (5.10)$$

where:  $T$  = absolute temperature;  
 $R$  = universal gas constant.

- Voltage decrease across a MK-40 cation-exchange

$$U_K = i/\kappa_K \quad (5.11)$$

where:  $\kappa_K$  = surface electroconductivity of a cation-exchange membrane.

- Surface electroconductivity of a cation-exchange membrane:

$$\kappa_K = (0,097 + 0,026C) \exp[(1/T - 1/293)(-33900/R)] \quad (5.12)$$

where:  $C$  = Concentration of sodium hydrosulphite (equivalent/litre or eq/l).

- Electroconductivity of the sodium hydrosulphite solution:

$$\kappa_P = (0,00946 + 3,44 * 10^{-2}C) \exp[(1/293 - 1/T)(11400/R)] \quad (5.13)$$

- For current densities is higher than  $100\text{A/m}^2$ , the electrical energy consumption for the conversion of 1eq of sodium hydrosulphite is described in terms of (W h/eq):

$$W = 1,053\alpha F[U_{KC} + U_{AC} + (U_{BM} + U_K + 2hi/\alpha_P)n]/n \quad (5.14)$$

where:

- $\alpha$  = coefficient taking into account an increase in the electrolyser resistance due to the introduction of gaskets;
- $F$  = Faraday number (26,8A h/eq);
- $U_{KC}, U_{AC}$  = the voltage decrease in cathode and anode chambers respectively
- $h$  = the distance between the membranes;
- $\alpha_P$  = from equations (1.2), (1.4) and (1.6);
- $n$  = number of cells.

### 5.2.3 Conclusions

- The electromembrane method has a great deal of potential because of the three-layer bipolar membranes, which has a catalytic effect in the water dissociation reaction.
- This method is based on the use of standard absorption apparatus for the removal of sulphur dioxide from the flue gas by a sodium bisulphite solution.
- The resulting sodium bisulphite is converted to sulphite and sulphur dioxide by use of standard electrodialysis installations. The sulphur dioxide is processed to sulphuric acid.

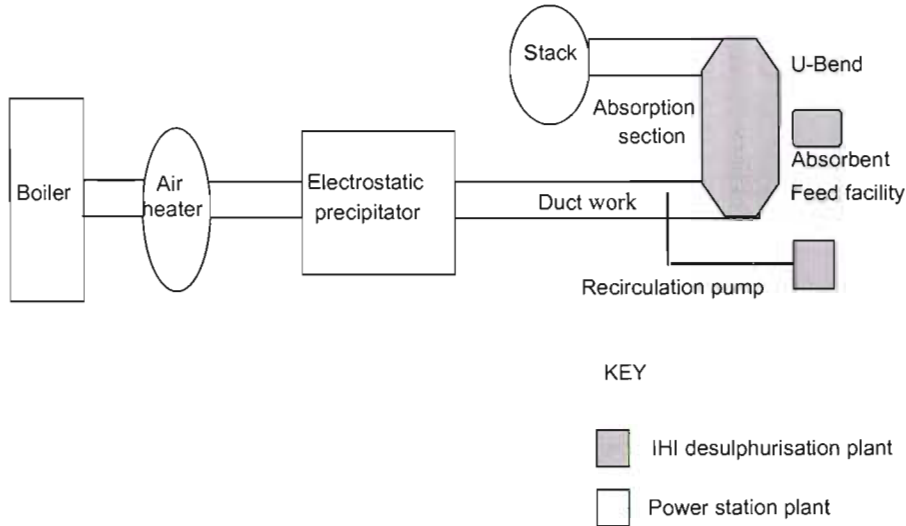
## 5.3 ISHIKAWAJIMA- HARIMA HEAVY INDUSTRIES IN-LINE TYPE FGD

This system was developed as an improvement to the conventional FGD system by Ishikawajima-Harima Heavy Industries (IHI) Company, (Ltd), Tokyo, Japan.

### 5.3.1 Process Description

The flue gas passes through an air heater, electrostatic precipitator (ESP) and an absorption section, which forms part of the duct work. The absorbent ( $\text{CaCO}_3$ ) is fed automatically to the desulphurisation system according to the demand signal from the absorption liquid pH sensor or the  $\text{SO}_x$  concentration sensor at the outlet of the desulphurisation section. The  $\text{SO}_x$  is absorbed by the limestone slurry, which is supplied by a nozzle spray, with an efficiency of up to 70%. The absorbent is oxidised and converted to gypsum ( $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$ ), after which the gypsum is separated from the liquid in the dehydrator.

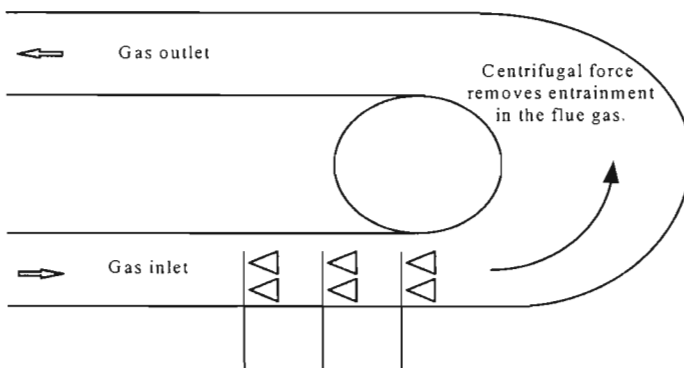
After desulphurisation, the flue gas is introduced into a U-shaped duct where the absorption liquid is separated from the gas by centrifugal force. The clean gas is emitted into the atmosphere. The  $\text{CaSO}_3$  is oxidised by oxygen in a recirculation tank.



**Figure 5.4:** Overview of the IHI plant (IHI Pamphlet, 1994)

### 5.3.2 Advantages

- The construction costs are 30-50% cheaper and operating costs are 50-60% cheaper than conventional FGD systems.
- Lower construction costs and a smaller installation area is required.
- No need for a mist eliminator as reintrainment is removed in the U-shaped duct. If higher removal efficiencies are required, a conventional mist eliminator can be installed.



**Figure 5.5:** Close up view of the U-bend (IHI, 1994)



- Lower operation costs due to the lower liquid-gas ratio and lower pumping head
  - High slurry concentration is used and therefore no thickener is required to enhance the gypsum
  - The system is simple because of the continuous dehydrator.
  - Air is injected into the recirculation tanks to oxidise the  $\text{CaCO}_3$  to gypsum, therefore an oxidation tower is not required. (IHI, 1994)
-

## CHAPTER 6

### MAJUBA POWER STATION

Eskom was established in 1923. It has grown into the world's largest coal burning utility, burning in excess of 80 million tons of coal annually with an installed generating capacity of over 38 000 MW. This constitutes 98% of the country's peak demand and more than 50% of all electricity consumed in Africa (Power-Gen. International, 1998). The biggest problem that Eskom is currently facing, is the air pollution from the stacks.

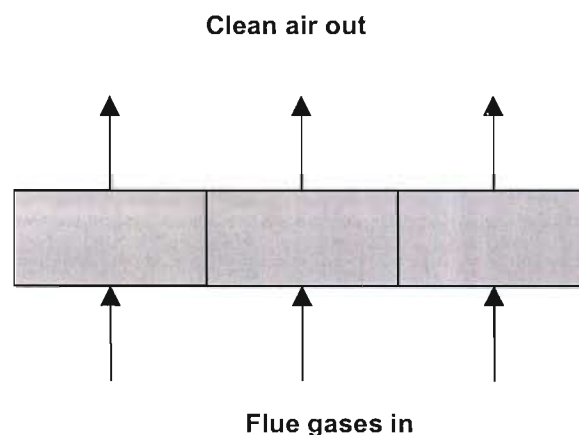
Majuba is situated in the southeastern highveld of Mpumalanga Province at an altitude of 1709 metres above sea level. The power station consists of 3 x 713 MW indirect wet cooled coal fired units totaling 4110 MW generating capacity. The first unit was declared commercial on 1<sup>st</sup> April 1996, with subsequent units being commissioned at yearly intervals (Smith, 1998).

Boiler flue gas particulate removal is achieved by means of a pulse jet fabric filter (PJFF), which when in full commercial operation will be the world's largest of this type (Smith, 1998).

The Chief Air Pollution Control Officer (CAPCO) of the Department of Environmental Affairs and Tourism (DEAT) set the particulate emission level at 50mg/Nm<sup>3</sup> (Presidents Council, 1990). To meet these lowered emissions levels, Eskom decided to employ a fabric filter plant.

#### 6.1 PULSE JET FABRIC FILTER

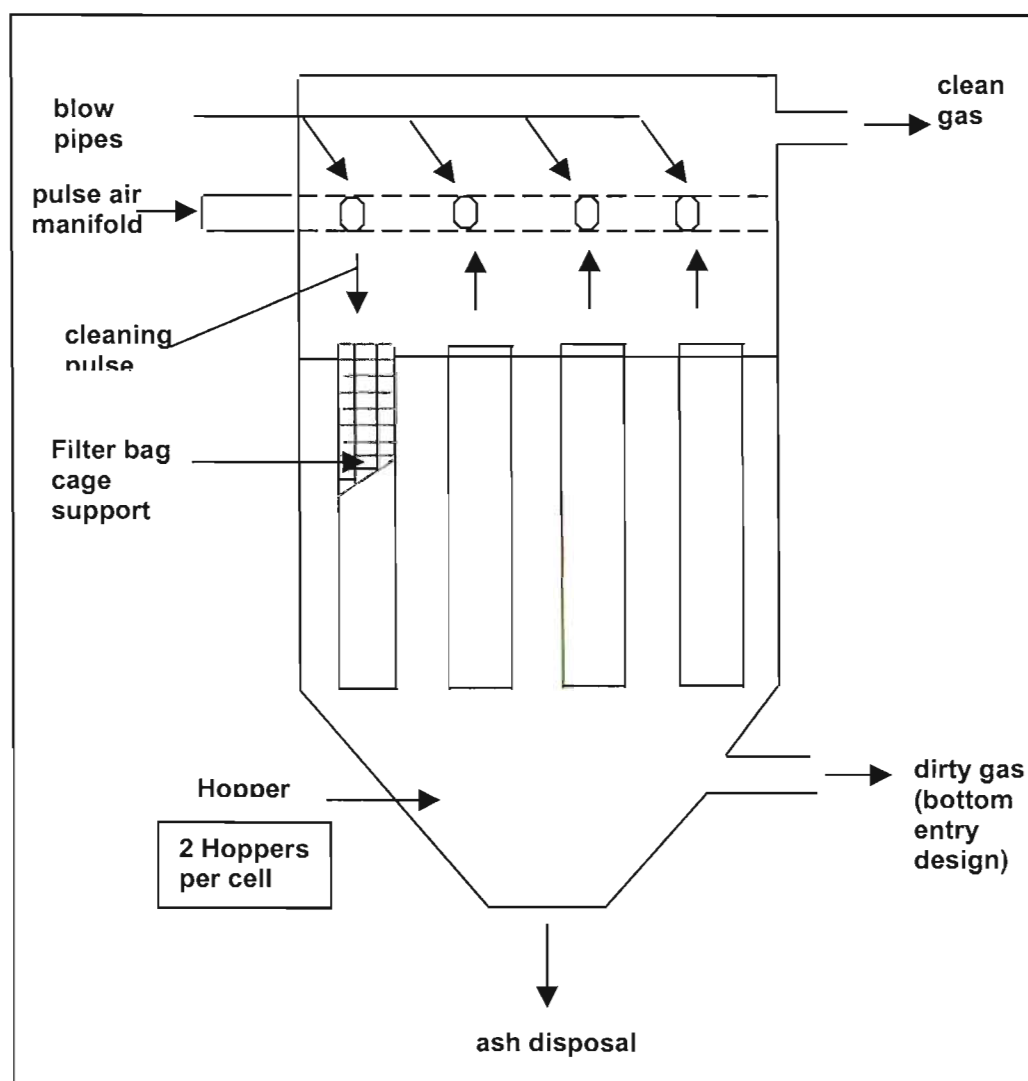
A precipitator can never be 100% efficient, and some dust is discharged via the chimney to the atmosphere. Previously, power stations used a normal shaken bag filter. This consists of multiple cells interconnected as indicated in Figure 6.1.



**Figure 6.1:** Schematic diagram of Normal Shaken Bag Filter

Through each compartment/cell flue gases pass. Upon entering, the bags are shaken to retain particulate matter and clean air is released at the top. These bags have to be periodically cleaned. This method however proved ineffective in complying with the limits as set by the Chief Air Pollution Control Officer. As a result of this, Eskom undertook an investigation to seek the best technological solution for upgrading existing electrostatic precipitators (ESP). The pulse jet fabric filter was subsequently introduced to overcome the problem.

Lurgi SA in partnership with Howden Energy Systems was awarded the contract to design and construct a pulse jet fabric filter that would achieve the new limits set by the CAPCO. Figure 6.2 shows a diagram of the pulse jet fabric filter.



**Figure 6.2:** Schematic diagram of a Pulse-Jet Fabric Filter

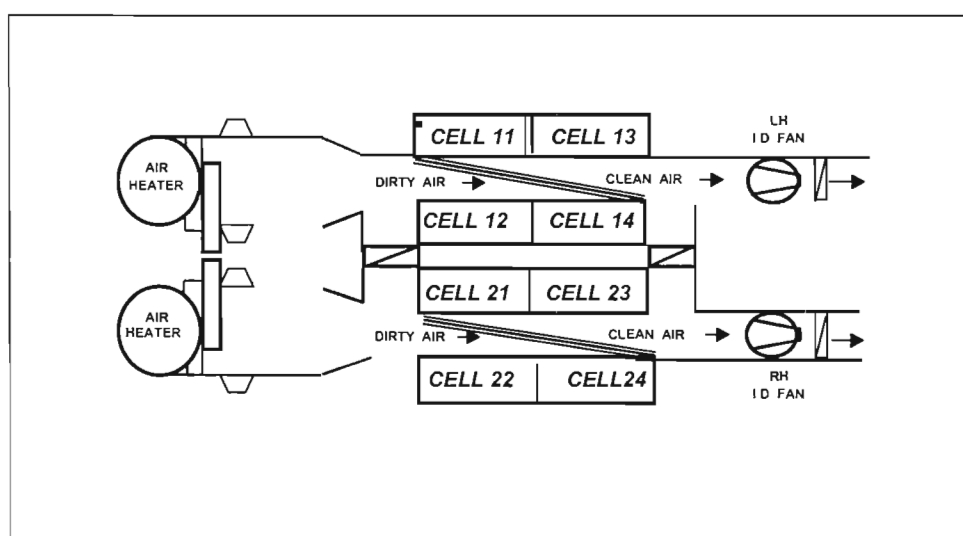
### 6.1.1 Design and Specifications of Pulse Jet Fabric Filter (PJFF)

The PJFF is designed to control particulate emissions from the boiler to below  $50\text{mg}/\text{Nm}^3$  when burning coal at boiler loads between 50 and 100% Maximum Continuous Rating (MCR) with 7/8 of the PJFF in service (Smith, 1998).

The bags for Units 1 to 3 were manufactured from a “low temperature” fabric called Dralon-T® Polyacrylonitrile. This was based on pilot plant tests conducted at Eskom’s Duvha Power Station. The operating temperature of this material is established to be  $125^\circ\text{C}$ . The boiler airheater exit temperature can vary from design of  $135^\circ\text{C}$  to  $150^\circ\text{C}$ . The flue gas is first brought to the required temperature by admitting ambient air in before the PJFF increases the flow to  $848\text{ Nm}^3/\text{s}$  for which the ID fans were upgraded by adding a second stage. A flue gas to cloth ratio (face velocity) not exceeding  $0.02\text{ m/s}$  is specified under maximum flow, the generally accepted value for high ratio filters. These design criteria are to be met with a pressure drop across the PJFF terminal points not exceeding  $2.5\text{ kPa}$  verified by acceptance tests carried out after 8500 operating hours.

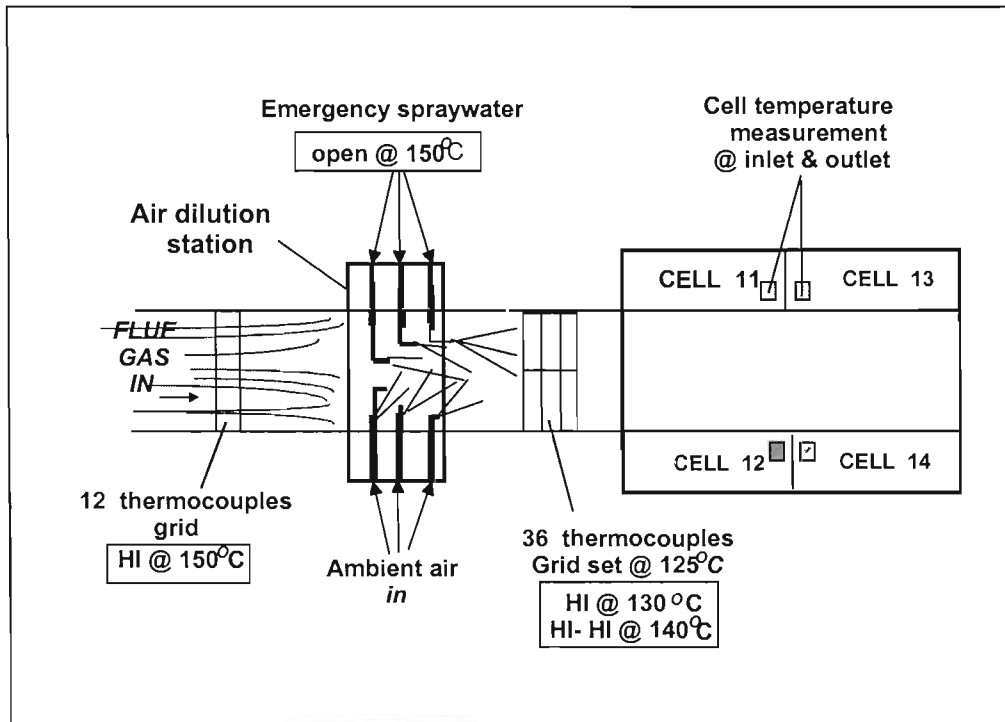
### 6.1.2 Plant Layout and Operation

Two inlet ducts separate the PJFF into the left and right hand casings each housing four cells (Figure 6.3). The cells are separated by balancing crossover ducts situated between the casings at inlet and outlets. The crossover ducts are accurately sized to allow one cell at a time to be removed from service at full load for maintenance. They are provided with pneumatically operated multi-louver dampers allowing isolation of one half of the PJFF for maintenance.



**Figure 6.3:** Schematic Layout of Isolatable Cells (Smith, 1998)

Cells have well insulated walk-in plenum chambers with ventilation fans allowing safe access shortly after removal from service. Each cell has four blade-type inlet dampers and six poppet-type outlet dampers, pneumatically activated, for isolation (Figures 6.4). Pressure sensors and thermocouples are installed at each inlet and outlet to monitor operating temperature and pressure.



**Figure 6.4:** Air System Temperature Control (Smith, 1998)

Each of the filter bags is 7 metres long and elliptical in cross section. They are arranged in 14 non-radially aligned concentric circles above which a manifold rotates delivering an 80 kPa cleaning pulse. There are 484 bags in a module and 8 modules in each cell totaling 30 976 bags per unit, a cloth area of 84 560m<sup>2</sup> (Smith, 1998). Polyacrylonitrile (PAN) bags are installed at Majuba because of their resistance to acidity and alkalinity. Below each cell are two 10 to 12 hour ash storage capacity, inverted pyramid shaped hoppers with thermostatically controlled heaters to maintain the dust at a handleable temperature. Pneumatic level switches indicate ash levels. The fly ash is removed continuously from the hoppers by dust handling plant airslides to a central surge hopper.

The PJFF casings are seal welded and strengthened to withstand internal pressure and external wind loading. They are supported approximately 20 metres above the ground level on open cross-based steel structures. Due to thermal expansion of ducts and casings, provision is made for this by sliding support bearings between casings/hoppers.

The attemperating air system consists of a grid of 12 thermocouples mounted upstream indicating the airheater outlet temperature, and a downstream grid of 36 thermocouples which admit the cooling ambient air. The attemperating system serves to distribute ambient air into the inlet duct assuming there exists fairly laminar flow in the duct. In the case of the temperature exceeding the maximum design specification, an emergency spray water system is provided to cool the flue gases.

The PJFF is designed to operate automatically supervised via touch sensitive VDU's from the Unit Control Room (UCR). A programmable Siemens automation system is used to monitor the various instruments and bag cleaning sub-system. It ensures that the flue gas entering the PJFF remains below the design maximum level and adjusts frequency of bag cleaning to maintain the pre-selected pressure drop across the PJFF. It controls alarms and indications in the UCR.

### 6.1.3 Correlation of Stack Emission Monitors

Majuba has a continuous monitoring system installed in the 250-meter high multi-flue chimneys. Each chimney houses three 8.1 m diameter flues. Dust is measured by Sick® OMD 41 to 3 transmissometers, SO<sub>2</sub> and NO<sub>x</sub> by GM 30 to 2P gas analyzers. Long term results for dust emissions are low (Figure A3.1).

The dust readings in mg/Nm<sup>3</sup> obtained from the Sick® OMD 41 to 3 transmissometers are correlated against the actual measurements to give a unique curve.

$$y = m\{(dust \times 0.16) + 4\} + c \quad (6.1)$$

where:         $y$         = dust in mg/Nm<sup>3</sup>        (dust = % reading)  
                   $m \& c$     = test result constants (for Unit 1:  $m = 5.917$  &  $c = -20.516$ )

Figure A3.1 shows a comparison between dust concentrations recorded at the monitoring station and the actual dust concentrations using equation 6.1 together with a line representing the maximum permissible levels of particulate matter, according to the CAPCO, which would not require particulate control. All dust emission limits are well below the required limit, proving that PJFF is effective in particulate removal.

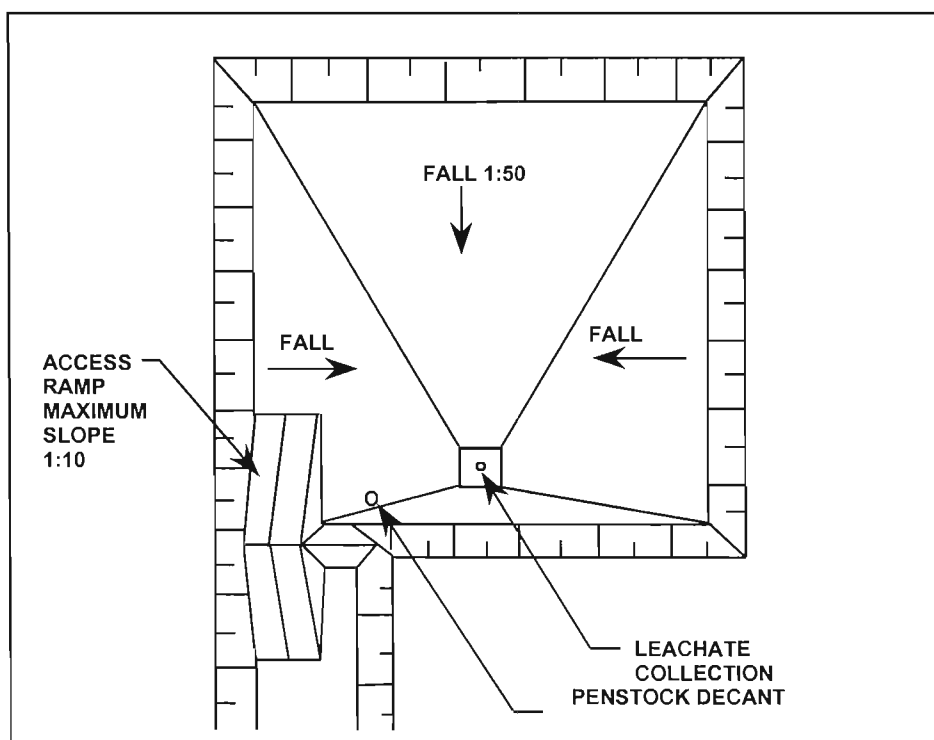
The method employed above is not very accurate because the correlation is affected by the characteristics of the dust in the measuring area, and depends on the operating conditions and the way the materials are handled.

### 6.1.4 Plant operational problems

- Mechanical damage due to bags rubbing against the casing wall and bag cages unclipping. High velocities into the cells push the bags into contact with the casing wall eroding the fabric. This results in an increase in operating cost in terms of bag replacement.
- High velocities into the cells result in high erosion of ductwork, filter bag ashcake and fabric, and also result in ash re-entrainment.
- Potential ash spillages from the PJFF hopper level detection pipework and dust handling plant surge hopper vent valve due an increase in pressure.
- The flue gas temperature must rise above the acid dewpoint as quickly as possible to limit acid degradation of the filter bags. This has been solved by a stepped heating-up rate to prevent sulphuric acid from being deposited on the bags during the periods below the acid dewpoint. Not only does the acid attack the PAN fabric but also causes the ashcake to become more adhesive.
- Fire hazard if the ash is too hot. Rise in plant operating temperature due to faulty thermocouple readings. Both problems are solved by the installation of an emergency spray water system.

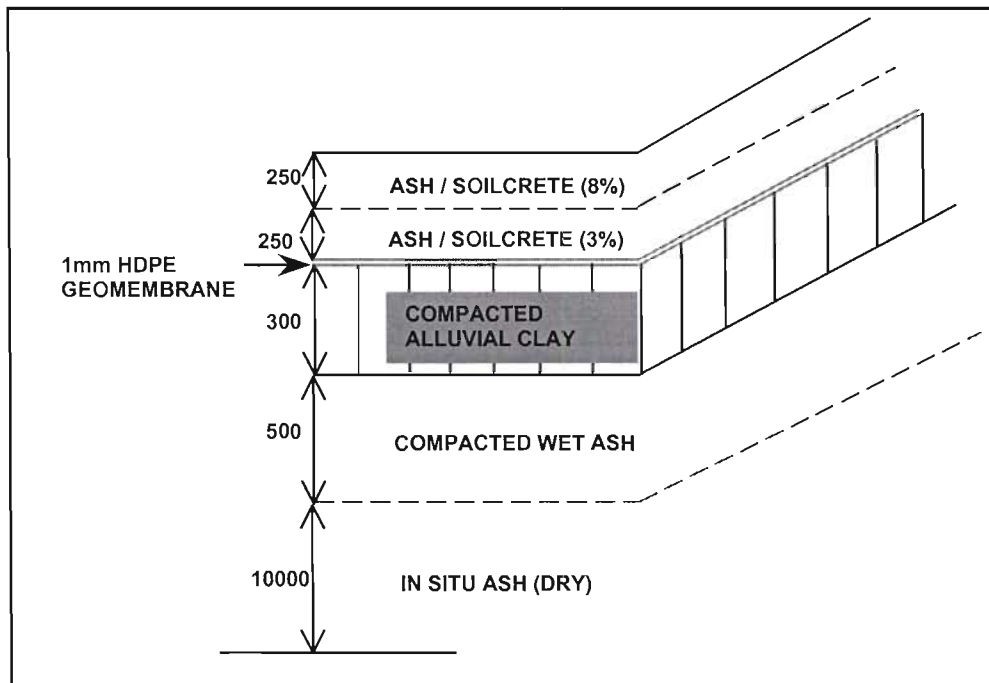
### 6.1.5 Waste Disposal

The dry ash produced is disposed of in a dry ash dump. Majuba power station does not have a toxic waste disposal site, consequently all wastes are either disposed of at Tutuku power station, or via a waste disposal company, e.g. Waste Tech, is employed to dispose of these wastes (Smith, 1998). Appendix 2 provides a list of toxic particulates and gases.



**Figure 6.5:** Plan of Waste Disposal Site at Tutuku Power Station

The wastes from the sludge contain a high concentration of chlorides, sulphates, sodium and magnesium. If these wastes are successful in reaching the groundwater, they will increase the total dissolved solids of the water body, affecting both plants and aquatic species. To prevent the leaching of wastes to the groundwater, the landfill is therefore lined with an impervious clay layer (Figure 6.6).



**Figure 6.6:** Landfill Liner of Toxic Waste Disposal System

### 6.1.6 Advantages

- The particulate emissions from PJFF will be low enough to avoid future retrofit actions that may arise out of more stringent legislation.
- PJFF can operate a higher volume gas for relatively fewer bags and the required number of bags can be retrofitted within the confines of an existing two-field ESP casing.
- Has an efficiency rate of 99%.
- Capable of removing significant portions of sulphur dioxide.
- Ability to capture very fine particulate matter ( $< 1\mu\text{m}$ ) (Dunderdale, 1990).
- Approximate life span of bags is 15 000 hours (Smith, 1998).

### 6.1.7 Future enhancements

- Increasing the number of bags to reduce dust emissions.
- Installation of a bypass damper system to extend bag life.



## 6.2 ESKOM'S ENHANCEMENT PROGRAM

Many of Eskom's older Electrostatic Precipitator's (ESP's) cannot comply with the particulate emission levels currently required by the Chief Air Pollution Control Office (CAPCO). The reasons for this include the following:

- Low design specifications. While these specifications were acceptable when the older power stations were designed and commissioned, they have since been reviewed because of industrial expansion and population growth in the once remote locations.
- Excessive gas flow rates. These are generally due to deterioration of coal quality.
- ESP deterioration. Like any other component of a power plant, an ESP deteriorates over time and between overhauls.
- Low sulphur, high ash coals. The use of low sulphur coals results in reduced levels of sulphur dioxide emission, foregoing the use of FGD equipment. However such coals increase fly ash resistivity into the region where inefficient collection occurs. Additionally, high ash coal increases emissions even though the collection efficiency might not be impaired.
- Maintain effective water utilisation at power stations.

As a consequence of the above, an investigation has been conducted with the objective of seeking the best technological and economic solutions for upgrading existing ESP's at Eskom's coal-fired power stations. The technologies investigated were flue gas conditioning (FGC) and electrical enhancement. FGC is discussed in detail in Chapter 4, Section 4.4.5, pp 4-6.

## CHAPTER 7

### ECONOMIC ANALYSIS

This study was done to show the cost associated with the different detoxification processes. A variety of methods were used to calculate the costs. Some of the costs were obtained directly by contacting suppliers, such as Lurgi South Africa who in turn contacted Lurgi Germany. Other costs were based on existing data from reference plants and assumptions made. The aim is to determine the best process for South Africa to employ for future control measures. All costs were based on a plant with a capacity of approximately 600 MW. When comparing costs of the different processes, it is important that the basic values and assumptions are considered, because their effect on the end result is quite decisive.

#### 7.1 COST COMPONENTS

##### 7.1.1 CAPITAL COSTS

The capital costs consist of direct and indirect costs incurred up to the successful commissioning date of the facility. Direct costs include the costs of various items of equipment and the labour and materials required for installing these items and interconnecting the interconnecting systems. Indirect costs are expenditures for the overall facility that cannot be attributed to specific equipment. They include equipment such as freight and spares. For this analysis, the investment cost of a specific process is a constant regardless of the locality of where the desulphurisation plant is built.

##### 7.1.2 ANNUALIZED COSTS

This refers to operating costs divided into the fixed operating and variable operating costs. The type of process and the complexity associated with it determines fixed and variable operating costs.

###### 7.1.2.1 FIXED OPERATING COSTS

The fixed operating costs of desulphurisation consist of labour, maintenance and repair costs.

###### Labour

Includes costs of supervision and skilled and unskilled labour to operate, monitor and control the FGD process.

---

## Maintenance and repairs

Includes cost of manpower and materials required to keep the unit operating efficiently. The function of maintenance is both preventive and corrective, to keep failure of equipment to a minimum.

### 7.1.2.2 VARIABLE OPERATING COSTS

The variable operating costs depends on the price of electricity, steam and water, and considerably on the price of waste disposal.

## Utilities

Include costs of water for slurries, cooling, and cleaning, electricity for pumps, fans, valves, lighting controls, conveyors, and mixers. Also include costs of fuel for reheating flue gases, and steam for processing.

## 7.3 COST CALCULATIONS

A variety of methods were used to calculate costs. This was due to cost being difficult to obtain.

### 7.3.1 Conversion of Currency

This section has been included in the document as a point of reference for all the costs provided throughout the document.

**Table 7.1:** Exchange rate

Conversion from	Year		
	1988	1994	2000
1 USA dollar (\$)	1.159	1	
1 \$/ R		R3.192	6.050
1 German Mark (DEM)		1	
1 DEM / R		DEM0.472	3.185
1 UK pound (£)		1	
1 £ / R		R5.479	9.915
1 Finland Mark (FIM)			
1 FIM / R			0.526

## 7.4 ECONOMIC EVALUATION FOR A WET SCRUBBING AND SPRAY DRY SCRUBBING PROCESS

The cost of desulphurisation consists of investment and operating costs.

### 7.4.1 INVESTMENT COSTS

The investment costs of desulphurisation increases as a function of heat effect of the power plant according to the functional characteristics of each process. This applies to the formula below:

$$I = a * (b + P)^c \quad (7.1)$$

where:  $P$  = firing rate (MW) and  $a$ ,  $b$  and  $c$  constants.

**Table 7.2:** Values of constants of investment cost function (Valimaki, 1990)

Process	a	b	c
Wet Scrubbing	0.145	180	0.983
Spray-Dry	0.560	60	0.770

### 7.4.2 OPERATING COSTS

The operating costs are divided into the fixed operating and variable operating costs.

#### Fixed Operating Costs

The fixed operating costs of desulphurisation consist of labour and maintenance costs. Labour costs were calculated on the assumption that there were 8 operators, with each operator working for 8 hours per day, 300 days a year. Maintenance costs are estimated to be 2% of the investment/capital costs.

#### Variable operating costs

The waste is assumed to be located in the power plant area

#### Waste Cost

This cost is mainly caused by the treatment of desulphurisation waste. Some of the parameters considered is the distance the waste has to be transported and dumping area costs if necessary.

The waste costs is estimated according to the following function (Valimaki, 1990):

$$W = ((A + B + C) \times D) \times Mw \quad (7.2)$$

Where

- W = waste cost (R/t/day)
- A = loading cost (R/t)
- B = dumping area charge (R/t)
- C = transportation cost (R/t/km)
- D = transportation distance (km)
- Mw = waste quantity (t/day)

### Fly Ash

The commercial value of fly ash is calculated based on the following (Valimaki, 1990):

$$FA = e \times mfa \quad (7.3)$$

Where

- FA = value of fly ash (FIM)
- e = unit price of fly ash (FIM/t)
- mfa = quantity of fly ash (t)

The transportation distance and the fly ash price are of great significance to the profitability of the process. Fly ash may be sold for use in the in the cement industry as a partial replacement for cement in concrete. Fly ash may also be used as a fill material and in soil reclamation. By using fly ash as a fill material, poor load bearing soils and uneven terrain can be converted into useable land. In the soil reclamation process, the fly ash is mixed with very acidic soils. This technique adjusts the pH of the soil and allows vegetation to grow, and is found to be most useful in reclaiming surface-mined areas. This reduces the operating costs.

### Values of Operating Costs

Table 7.3 provides a list of all operating costs with the average current (year 2000) prices.

**Table 7.3:** Operating Costs

Description of Operating Costs	Prices
Average labour cost	R50/hr
Coal sludge	R68/ton
Loading cost of sludge + dumping area charge	R10/ton
Transportation cost of waste	R2.50/ton/1.5km R1.40/ton/0.5km
Transportation distance to dumping area	±1.5km
Unit price of fly ash	R30/ton
Average ash content in coal	25% of coal burned
Quantity of fly ash	80 ton/hr
Lime cost	R296.23/ton + 14% vat
Limestone cost	R44.04/ton + 14% vat
Water (price for 1994)	R1.29/m <sup>3</sup>
Coal	R68/ton
Electricity (price for 1990 estimate, Max & Klaus, 1990)	R0.18/kWh
Steam (price for 1990 estimate, Max & Klaus, 1990)	R16.94/1000lb
Waste (price for 1990 estimate, Valimaki, 1990)	R4.19/t
Waste water (price for 1990 estimate, Valimaki, 1990)	R4.19/t

- Lime/limestone costs were obtained from PPC LIME
- All other variable operating costs were obtained from contact with Bruce Smith (Majuba power station).

**Table 7.4:** Summary of costs for Wet Scrubbing Process

COSTS	
<b>CAPITAL COSTS</b>	
<b>Total capital cost</b>	<b>R555.51M</b>
<b>OPERATING COSTS</b>	
<b>Fixed</b>	
• Labour	2.88
• Maintenance	11.11
<i>Total fixed cost</i>	<i>13.99</i>
<b>Variable</b>	
• Lime	13.25
• Waste	0.74
• Waste water	0.14
• Water	0.90
• Steam + Electricity	112.97
<i>Total variable cost</i>	<i>128.00</i>
Recoverable Costs Operating Cost	
<i>Fly ash</i>	<i>21.04</i>
<b>Total Operating Cost</b>	<b>R120.95M/year</b>

### 7.4.3 Discussion of costs for Wet Scrubbing Process

The wet scrubbing process is the most widely used desulphurisation process for the removal of sulphur dioxide. It does not however facilitate the removal of nitrogen oxides. Table 7.4 shows the capital and operating costs. The wet scrubbing process has the advantage that it produces a high grade byproduct that can be sold to offset some of the operating costs. South Africa's energy/electricity supplier is Eskom. Eskom has a visible NO<sub>x</sub> plume problem at some of the power stations where particulate emissions are reduced by flue gas conditioning or the installation of fabric filter e.g. Majuba power station (Smith, 1998). Therefore, the process cannot be considered as it only facilitates sulphur dioxide removal. The wet scrubbing process was costed to provide a comparison to the other processes.

**Table 7.5:** Summary of costs for Spray Dry Scrubbing Process

<b>COSTS</b>	
<b>CAPITAL COSTS</b>	
<i>Total capital cost</i>	<b><i>R456.75M</i></b>
<b>OPERATING COSTS</b>	
<b>Fixed</b>	
• Labour	2.88
• Maintenance	9.14
<i>Total fixed cost</i>	<i>12.02</i>
<b>Variable</b>	
• Lime	24.65
• Waste	0.74
• Waste water	0.00
• Water	0.90
• Steam + Electricity	112.97
<i>Total variable cost</i>	<i>139.26</i>
Recoverable Costs Operating Cost	
<i>Fly ash</i>	<i>21.04</i>
<b><i>Total Operating Cost</i></b>	<b><i>R130.24M/year</i></b>

### 7.4.4 Discussion of costs for Spray Dry Scrubbing Process

This process is used for sulphur dioxide removal only. This process is generally more suited for small to medium sized boilers (approximately 300MW).

Comparing the results of Table 7.5 to Table 7.4, the spray dry scrubbing process is a more competitive process as compared to the wet scrubbing process. This is due to the fact that both the capital and operating costs are lower making the spray dry scrubbing process the more economical option to consider.

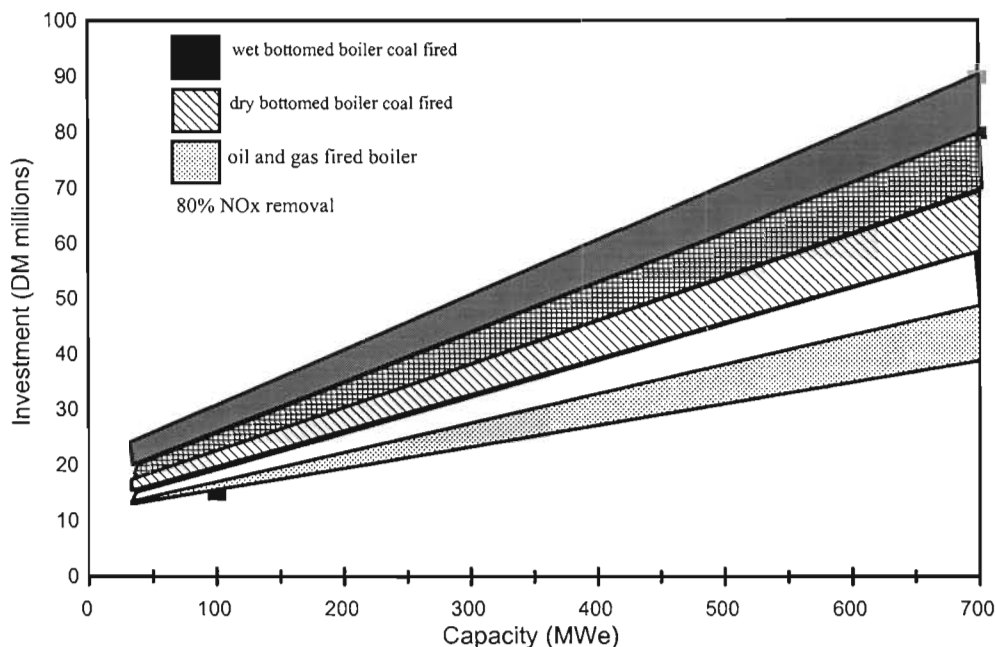
## 7.5 SELECTIVE CATALYTIC REDUCTION

The capital costs of SCR was calculated using the expression below, based on the fact that the costs appeared to be more dependent on the system capacity, rather than design (Lowe and Perlsweig, 1990). The expression is as follows (calculated on the 1988 DEM):

$$C1 = 1.27 \times Mwe^{0.71} \quad (7.4)$$

where:  $C1$  = capital cost in million German Mark (M. DEM)  
 $MWe$  = plant net electrical output in MW

The graph below is an indication of the variability of costs for retrofitted plants. It shows the capital cost estimates for German SCR retrofit installation plants based on 1985 DEM with a catalyst cost of 40 000DM/m<sup>3</sup> (Lowe and Perlsweig, 1990). The graph shows the costs of a wet bottom, dry bottomed and oil and gas fired boilers. In a dry bottomed boiler, the ash that is dislodged from the furnace wall is below the ash melting point and leaves the furnace bottom in a relatively dry condition. All current pulverized-coal reheat units are of the dry bottomed type. The wet-bottom or slagging-bottom furnaces are used for applications, such as furnaces for gasifying coal (Singer, 1981). All current pulverized-coal reheat units are of the dry bottomed type. For one dry bottom boilers of 590 MWe, the costs lies in the range 49 to 59M DEM. This is equivalent to R157.12M to R189.18M at the 2000 exchange rate.



**Figure 7.1:** Capital cost estimates for German SCR retrofit installations.



There was not much information available with regards to running costs of the plant. The costs below are merely estimates (Lowe and Perslweig, 1990).

### Estimated annual running costs

Tail gas installations: 2 – 4c/kWh for 600 MWe  
High dust installations: 4 – 5c/kWe for 600 MWe

## 7.6 CIRCULATING FLUIDISED BED (CFB)

The cost assumptions were obtained from discussions with Lurgi Germany (Tilly, 1996). The capital costs for the CFB are less than for wet FGD-plant (Gypsum process). For a wet FGD process, the operating costs associated with the CFB would be higher because the stoichiometric factor of SO<sub>x</sub> absorption is higher for CFB than for wet FGD. Based on the typical flue gas volume of 532 Sm<sup>3</sup>/s for a 600 MWe Eskom boiler, Lurgi considers a CFB plant to be less economical than wet scrubbing (Tilly, 1996).

Since Lurgi had not built a plant to this size, they have provided an estimated budget price as follows (Tilly, 1996):

**Capital costs:** 75% of equivalent wet FGD  
**Operating costs:** 130% of equivalent wet FGD

The above are estimated costs.

## 7.7 ELECTROMEMBRANE DESULPHURISATION

A detailed economic evaluation is done in Appendix 4. The data obtained during the economic evaluation is shown in Table 7.6. The evaluation was carried out according to standard methodology. The extra costs of construction and maintenance of the power plant were not taken into account. All cost were converted using the current exchange rate of R6.05 per US\$.

**Table 7.6:** Summary of costs for the Electromembrane Desulphurisation Process (Appendix 4, Table A4.2).

<b>I. CAPITAL COSTS</b>	<b>US\$ * 10<sup>3</sup></b>	<b>Rand * 10<sup>3</sup></b>
<i>Building for housing electrodialysis units.</i>	250.40	1514.92
<i>Chemically stable pumps</i> (pumps the working solution into electrodialysers) <ul style="list-style-type: none"> <li>Type X-80-50-200A.</li> <li>Power 30 kW. 1 in operation, 1 reserve.</li> </ul>	5.12	30.98
<i>Assembling of pumps</i> (Construction regulations, IV, 6-82, part 22)	1.004	6.07
<i>Electrodialysers</i> (with 2 reserves) <ul style="list-style-type: none"> <li>Production of Alma-Ata electromechanical plant.</li> <li>Cation-exchange membrane area 453m<sup>2</sup></li> <li>Bipolar membrane area 453m<sup>2</sup></li> </ul>	2595.34	15701.81

<b>I. CAPITAL COSTS</b>	<b>US\$ * 10<sup>3</sup></b>	<b>Rand * 10<sup>3</sup></b>
<i>Assembling of electrodialysers</i>	111.595	675.15
Pumps for solution delivery into absorbers, similar to 2.3	5.12 1.004	30.98 6.07
<i>Absorbers, material</i> <ul style="list-style-type: none"> <li>• Steel 09G2S - 12X18H10T</li> <li>• Mass 45 950 kg (price list 23-03)</li> <li>• Number of working absorbers - 5, reserve - 1.</li> </ul>	1942.74	11753.58
<i>Assembling of absorbers</i> (Construction regulations iv-6.82 p.18-146-1)	134.185	811.82
<i>NaSO<sub>4</sub> required</i>	2.627	15.89
<b>SUB-TOTAL FOR CAPITAL COSTS</b>	<b>\$5.049M</b>	<b>R30.56M</b>
<b>II. OPERATING COSTS</b>		
• <i>Electrical energy consumed by electrodialysers</i>	500.3	3026.82
<i>Electrical energy consumed by all pumps</i>	11.680	70.66
<i>NaSO<sub>3</sub> losses</i>	0.495	2.99
<i>Land usage rates (standard rate at R36/m<sup>2</sup>)</i> (includes all buildings, pumps and absorbers <ul style="list-style-type: none"> <li>• Total area 1000 m<sup>2</sup></li> </ul>	10.00	60.50
<i>Personnel salary (minimum state salaries are given)</i> <ul style="list-style-type: none"> <li>• 5 operators</li> <li>• 2 technicians</li> <li>• 1 construction</li> <li>• 1 labouratory assistant</li> <li>• 7 electrodialyser maintenance</li> </ul>	22.59 7.028 3.514 2.410 29.618	136.67 42.52 21.26 14.58 179.19
<i>Maintenance costs</i> <ul style="list-style-type: none"> <li>• absorbers</li> <li>• electrodialysers</li> </ul>	97.137 122.39	587.68 742.57
<i>Membrane exchange costs</i> (10% total membrane area)	98.241	594.36
<b>SUB-TOTAL FOR OPERATING COSTS</b>	<b>\$0.906M</b>	<b>R5.48M</b>

### 7.8.1 Discussion of costs for Electromembrane Desuplurisation Process

Facilitates the removal of sulphur dioxide. Selectively removes one chemical species, ionically transports it across a membrane via a potential gradient, and then regenerates the chemical on the other side of the membrane. Membrane technology appear to be most promising in that it has low capital and operating costs making this a cost effect option to consider. It also produces a saleable byproduct. However, the process does not facilitate NO<sub>x</sub> removal. Due to the low operating and capital costs, this process should not be discarded on the grounds that it does not reduce NO<sub>x</sub> emissions. The costs of retrofitting existing power stations with additional NO<sub>x</sub> control equipment may be cheaper than using a process that removes both SO<sub>2</sub> and NO<sub>x</sub>.

## CHAPTER 8

### CONCLUSIONS AND RECOMMENDATIONS

#### 8.1 Conclusions

Currently, South Africa's emissions compare favorably with the international legislative requirements with the exception of the German limits, which are currently the strictest. This is because South Africa burns a lower grade quality of coal with low sulphur contents. Although the coal has a low sulphur content, the emissions per terawatt hour are relatively high, because large quantities of coal are burnt due to the lower calorific value.

Emissions of sulphur dioxide ( $\text{SO}_2$ ), nitrogen oxides ( $\text{NO}_x$ ), and particulate matter are predicted to increase by the end of the century under a high coal scenario (Bosnak, 1997).

A review of the various processes of desulphurisation indicates that the most widely employed technique is the wet scrubbing process. The operating costs of this process is high but it does however produce high-grade saleable byproducts and is characteristic of a high  $\text{SO}_2$  removal efficiency.

Dry absorption processes e.g. spray dry scrubbing, activated coke process, produce no effluent, because the water is completely evaporated in the absorber. It is generally associated with high  $\text{SO}_2$  removal efficiency. Because operation of dry absorption processes occur within the dew point range, plugging of tank and pipe occur. Also common, is the corrosion of electrostatic precipitators and fabric filters.

Catalytic processes achieve very high removal efficiencies of both  $\text{SO}_2$  and  $\text{NO}_x$ . These processes have a very good on-line reliability and are simple from an operating perspective. It is advantageous in that the flue gas energy is generally recovered to the process as medium grade energy and that these processes generally produce saleable byproducts. Catalytic processes require high dust removal from the flue gas to ensure optimum removal of  $\text{SO}_2$  and  $\text{NO}_x$ .

Investigations into future emerging technologies showed membrane desulphurisation to be cost effective, as well as having high sulphur dioxide removal efficiencies.

An investigation was carried out to determine a low cost, reliable method for the removal of particulates from the flue gas. What became evident during the research is that fabric filters/bag filters have high removal efficiencies and despite the high operating and maintenance costs, proves to be more beneficial in the long term because they have the capability to remove very fine particulate matter. They therefore

eliminate the need to install more control equipment to maintain emission levels below the legislative limits. Further, for existing power stations with electrostatic precipitators (ESP), the addition of a chemical conditioning mechanism such as sulphur trioxide would significantly improve the performance of the ESP.

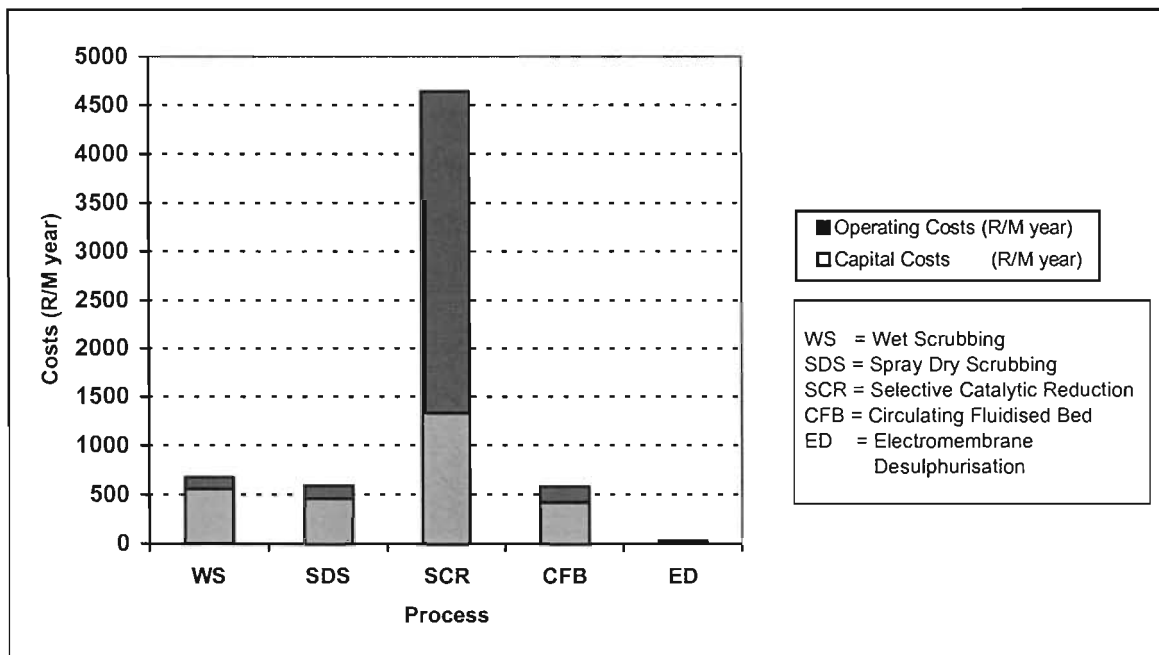
Majuba power plant has an ambient air quality monitoring program. Dust is measured by Sick® OMD 41 – 3 transmissometers, SO<sub>2</sub> and NO<sub>x</sub> by GM 30 to 2P gas analyzers. Data were obtained from the chief environmental officer for the years 1998 and 1999. A trend analysis was conducted on the data collected for this period. In general, the SO<sub>2</sub> trends show a decrease over the monitoring period. Long term results for dust emissions are low (Figure A3.3). The actual concentrations of SO<sub>2</sub>, NO<sub>x</sub> and dust emissions, however, have been below the Department of National Health and Population Development (DNHPD) limits for the entire monitoring period.

With more attention being focused on the environment, South Africa is likely to introduce gaseous emission legislations in the foreseeable future. It is therefore important to know what the most cost-effective options are for gas emissions in order to access the impact on the business. Based on this, a cost analysis was done on various detoxification technologies. A variety of methods were used due to costs being difficult to obtain. All of the costs were converted using the January 2000 exchange rates.

The processes being compared are a typical wet scrubbing, spray dry scrubbing, selective catalytic recovery, circulating fluidised bed and electromembrane desulphurisation process. As the outcome of this comparison depends on the basic data, limitations and assumptions made for each case, the results of the cost analysis should not be over-generalised. The results have been summarized below.

**Table 8.1:** Capital and Operating cost comparisons

Process	Gases Removed	Capital Cost R M/year	Operating Cost R M/year
Wet Scrubbing	SO <sub>2</sub>	555.51	120.95
Spray Dry Scrubbing	SO <sub>2</sub>	456.75	130.24
Selective Catalytic Recovery	NO <sub>x</sub>	1336.78	3300.00
Circulating Fluidised Bed	SO <sub>2</sub> , NO <sub>x</sub>	416.33	157.24
Electromembrane	SO <sub>2</sub>	30.56	5.48



**Figure 8.1:** Total costs divided into capital and operating costs

Figure 8.1 shows the electromembrane process to be the most promising technology in that it has the lowest operating and capital costs. The disadvantage is that this process facilitates  $\text{SO}_2$  removal only. One of the problems experienced by Eskom power station is a visible  $\text{NO}_x$  plume problem from stations using fabric filters and flue gas conditioning for particulate removal. Therefore, a more appropriate technology would be one that removes both  $\text{SO}_2$  and  $\text{NO}_x$  emissions.

Wet scrubbing process has the highest capital costs next to the selective catalytic reduction process making it an uneconomical option to consider. Further, this process facilitates  $\text{SO}_2$  removal only.

According to the figures in Table 8.1, the spray dry scrubbing process is not the most competitive process. While capital costs are lower than the wet scrubbing process, the operating costs are still higher. Used for  $\text{SO}_2$  removal only.

The circulating fluidised bed process has lower capital and operating costs than the wet scrubbing, spray dry scrubbing or selective catalytic reduction process. This process removes both  $\text{SO}_2$  and  $\text{NO}_x$ .

Therefore the best available technology that is also cost effective is the circulating fluidised bed.

## 8.2 RECOMMENDATIONS

- During the installation of desulphurisation in existing power plants, extra costs will arise. This is because the space required in the power plant area by the desulphurisation equipment has not been taken into account at the plant building stage. It is therefore recommended that when examining an existing power plant, extra investment costs must be taken into account to accommodate for FGD.
  - More research should be done on technologies that are both technically viable and cost effective. The possibility of the simultaneous removal of SO<sub>2</sub>, NO<sub>x</sub> and particulate matter should be investigated.
  - Eskom has a visible NO<sub>x</sub> plume problem at some of the power stations where particulate emissions are reduced by flue gas conditioning or by the installation of fabric filters. It is therefore recommended that they consider the cost of NO<sub>x</sub> removal when evaluating the capital costs of FGD process.
  - Further investigations on electromembrane desulphurisation should be done in order to achieve the correct efficiency.
-

## REFERENCES

- Andreasen, J, Laursen, J K and Rud Bendixen O**, "SNOX Plant in Full Scale Operation", Haldor Topsoe A/S, Nymollevej 55-DK-2800, Lyngby, Denmark, 1992.
- Amulya, K N, Robert, H R and Thomas, B W**, "Energy after Rio: Prospects and Challenges", UNDP, United Nations Plaza, New York, 1997.
- Ballinger Energy Series**, " Direct Use of Coal-Prospects and Problems of Production and Combustion", Ballinger Publishing Company, Cambridge, Massachusetts, 1979.
- Bechtel Corporation**, "Ocean Disposal of Coal Ash", San Francisco Printers, California, 1986.
- Bhabha Atomic Research Centre**, "Draft report by Radio Tracer Study carried out at Ennore (Madras) by Istope Divisions", Bhabha Atomic Research Centre, Trombay, 1971.
- Bogucki, R P and Pragnell, R J**, "Flue Gas Desulphurisation for Industrial Scale Coal-Fired Combustors - A UK Perspective", Sheffield, 19-21 April 1989.
- Bosch FJ and Hansen R S**, "Gaseous Detoxification Technologies and their Cost Implications," Eskom, Technology Research and Investigations, Eskom, Generation Engineering Division, South Africa, Johannesburg, 1997.
- Bosnak, A D**, "Coal and its Effects on the Environment", Eastman Kodak Company, 1997.
- Brevik, T J and Bisset, R**, "Shaping the Environmental Agenda of the 21st Century", UNEP Governing Council, 15 January 1997.
- Burnett, T, Patterson, S and Dail, R**, "Solutions for the Fossil Power Industry - South Africa's Eskom Licences Tube Alert Technology", Houston, Texas, 1997.
- Price List No. 23-03**, "Prices of Chemical Equipment - Part I", Standard Chemical Equipment Book Company, 1994.
- Coal Advisory Board**, "Pulverised Fuel-Fired Boiler Plants - Environmental Performance", International Energy Agency, 1996.
- Cock, J and Koch, E**, "Going Green: People, Politics and the Environment in South Africa", Oxford University Press, Cape Town, 1991.
- Consolidated Edison Corporation**, " Designation of an Ocean Disposal Site for Coal Ash from New York - New Jersey Metropolitan Area Utilities", New York, July 1983.
- Council of European Communities Directive**, "The Limitation of Emissions of Certain Pollutants into the Air from Large Combustion Plants", 88/609/EEC, 24 November 1988.
- Dacey, P and Cope, D R**, "Particulate Controls for Coal-Fired Boilers", I E A Coal Research Working Paper 69, London, January 1986.
- Davidson, B and Bradshaw, R W**, "Thermal Pollution of Water Systems", Environmental Science Technology 1, p 619, 1967.
- Davies iii, J C**, "The Politics of Pollution", 1996, p 3, 4.
- Dulka, J J and Risby, T H**, "Analytic Chemistry", United States of America, 1986.
- Dunderdale, J**, "Energy and the Environment", Royal Society of Chemistry Publishing Company, Great Britain, 1990.
- Financial Times**, Germany, January 14, 1994



- Friberg, L, Piscator, M, Nordberg, G F and Kjellstrom, T**, "Cadmium in the Environment", 2nd edition, CRC Press, Cleveland, 1974.
- Fuggle, R F, and Rabie, M A**, "Environmental Concerns in South Africa", Juta and Company, Ltd, 1989.
- Guilcher, M and Clarke, J D**, "Water Pollution Control Programs", Canada, 1997.
- Huntley, B, Siegfried, R and Sunter, C**, "South African Environments into the 21st Century", Human, Rousseau & Tafelberg, Cape Town, 1989.
- Hutton, C A and Allen, R C**, "Cleaning Up Coal", Ballinger Publishing Company, Cambridge, Massachusetts, 1982.
- Ishikawajima-harima Heavy Industries (IHI) Co (Ltd) Pamphlet**, "IHI In-line Type Flue Gas Desulphurisation System", November 1994.
- Jacobs, H L**, "Fly Ash Disposal", Sewage Industrial Wastes, 1986.
- Khan, S R, Desai, M S and Gavin, A F**, "NO<sub>x</sub> Control Technologies: An Overview of the Effectiveness for Stationary Combustion Sources", Proceedings from the American Power Conference, Pennsylvania, 1990.
- Klingspor, J S and Cope, D R**, "FGD Handbook - Flue Gas Desulphurisation Systems", IEA Coal Research, London, 1987.
- Kohl, A L and Riesenfeld, F C**, "Gas Purification 4<sup>th</sup> Edition", Gulf Publishing Company, Houston, Texas, 1985.
- Lave, L B and Seskin, E P**, "Toxic Chemicals: Health and Environment", Baltimore: John Hopkins, University Press, 1995.
- Lerner, L E**, "Conference on Smokeless Charging of Coke Ovens", Atmospheric Pollution Bill, 1956-1957.
- Linkov, V**, "Electrochemical Membrane Technologies", Sasol Centre for Separation Technology, Potchefstroom University for CHE, 1995.
- Livengood, C D and Markussen, J M**, "Emerging NO<sub>x</sub> and SO<sub>2</sub> Control Technologies", Argonne National Lab, 1 L Sponsor: Department of Energy, Washington, DC. Report No: ANL/Cp-75635, Conf-920294-1, February 1992.
- Lowe, P A and Perlswieg, M**, "Recent Experience for SCR Systems at Coal-Fired Utility Boilers", Proceedings of the American Power Conference, 1990.
- Ludwig, H F, Evans, J W, Brockman, W Y and Lohani, B N**, "Manual of Environmental Technology in Developing Countries - Volume 11", South Asian Publishers, Pvt. Ltd., New Delhi, 1991.
- Mackenzie, G A**, "The Newsletter of the UNEP Collaborating Centre on Energy and the Environment", Riso National Laboratory, Denmark, April 1996.
- Makansi, J**, "Seek The Best From World Class SO<sub>2</sub>, NO<sub>x</sub> Control Techniques", Electric Power International, June 1995. pp 23-33.
- Marten, J H and Lloyd Jr. G M**, "Coal-Fired Power Plant with Pollution Control and Useful By-products", Florida Institute, 1996.
- McKetta, J J**, "Acid Rain: The Whole Story to Date", National Acid Precipitation Assessment Project, Rational Readings on Environmental Concerns, Van Nostrand Reinhold, 1992, pp 44-58.
- Minns, S A, Sendlein, L V A, Dinger, J S and Currens, J C**, "Effect of Coal Ash Disposal on the Hydrological Environment", United Kingdom, 1995-1996.
- Naude, D P**, "Fluidised Bed Combustion Research", Energy Research Institute, University of Cape Town, Report No. In 021/C, February 1991.
-



- Nemerow, N L**, " Liquid Wastes of Industry-Theories, Practices and Treatment", Addison-Wesley Publishing Company Inc., Philippines, 1971.
- Pachauri, R K, Srivastava, L and Thukral, K**, "Energy- Environmental Development", Volume 1 and 2, Har-Anand Publications, New Delhi, 1991.
- Perry M S and Timmerhaus K D**, "Plant Design and Economics for Chemical Engineers", Fourth Edition, McGraw Hill Book Company, 1991.
- Perry R H and Green D**, "Perry's Chemical Engineers Handbook – 6<sup>th</sup> Edition", McGraw Hill Book Company, 1984.
- Portney, P**, "Does Environmental Policy Conflict with Economic Growth?", Resources for the Future, No. 115, Spring 1994, p 23.
- Powell, S T**, "Power Generating Stations Can Develop Stream Pollution Problems", Ind. Eng. Chem., 12A, 1954.
- Power-Gen. International**, "The Largest, Most Comprehensive Gathering of Power Generation Professionals in the World", Orange County Conference Centre, Orlando, Florida, December 9-11, 1998.
- Presidents Council**, "Report of the Three Committee's of the President's Council on a National Environmental Management System", Government Printers, South Africa, 1991.
- Ray, D L**, "Environmental Overkill: Whatever Happened to Common Sense?" Regnery Gateway Publishers, United States, Washington, D. C., 1993.
- Rinaldi, N U**, "Wet scrubbers: Choose the Best Chemical Reagent", Environmental Engineering World, March April 1995.
- Sauer, H and Anders, R**, "Operating Experience with a Dry FGC Plant (System CFB)", VGB Kraftwerkstechnik 69(10), October 1989, pp 887-892.
- Sauer, H, Weisweiler, W and Riley, J O**, "Simultaneous SO<sub>2</sub> and NO<sub>x</sub> Removal in a Circulating Fluidised Bed", Presented by LURGI, The First International Power Technology Conference, Chicago, Oct 31- Nov 2, 1989.
- Savage, W H D**, "Average Analyses of Product Samples taken by the Fuel Research Institute during 1978 and Analyses of Ashes of Selected Coal Products", Fuel Research Institute of South Africa, Bulletin No. 90, 1978.
- Schoeberlein, D**, "Electricity Environmental Issues", Energy Information Administration/Annual Energy Outlook, 1996.
- Shannon, R H**, "Handbook of coal – based Electric Power Generation; The Technology, Utilisation, Application and Economics of Coal, for Generating Electric Power", Noyes Publ., Park Ridge, New Jersey, USA, 1982.
- Sheahan, R T**, "Evaluating the FBC Option", Government Institutes, USA, 1983.
- Singer, J G**, "Combustion: Fossil Power Systems 3<sup>rd</sup> Edition", Cape Town, Combustion Engineering Incorporated, 1981.
- Smith, B**, "Air Quality Impacts at Majuba Power Station", Environmental Officer, October 1998.
- Stenby, E W, Hoskins, W W and Donaldson, A M**, "Environmental Control Impacts of Selected Alternate Fuels on Existing Plants," Stearns-Roger Engineering Corporation, Hemisphere Publishers, USA, 1990.
- Takeshita, M**, "Air Pollution Control Costs for Coal-Fired Power Stations", Centre for Applied Energy Research, January 1995.
-

- Technical Information**, "Kendal and Lethabo Power Stations Pamphlet", Communications Department, Megawatt Park, 1994.
- Tilly, H**, "Gaseous Detoxification Technologies", Technology Research Report Number TRR/P96/016, South Africa, September, 1996.
- Tilly, H A and Keir J**, "Keirs Environmental Database (KED)", Updated information for 1982 to 1995, 1995.
- Tyson, P D, Kruger, F J and Louw, C W**, "Atmospheric Pollution and its Implication in the Eastern Transvaal Highveld", South African National Scientific Programmes Report No. 150, CSIR, Pretoria, 1988.
- Valimaki, E J**, "Cost of the Lifac Desulphurisation", The Institute of Energy, Tampella Ltd, London, 1990
- Vernon, J**, "Sulphur Control: The Regulatory Framework", KPMG Peat Marwick Management Consultants, Institute of Energy, London, 1990.
- Vernon, J L**, "Emission Standards for Coal-Fired Plants: Air Pollutant Control Policies", IEACR/11, London, UK, IEA Coal Research, 52pp, August 1989.
- Veziroglu, T N**, "Environmental Problems and Solutions", Hemisphere Publishing Corporation, United States of America, 1990.
- Weiler, H**, "Transport, Storage and Handling of Limestone and Gypsum in FGD Plants", Bulk Handling of Limestone and Gypsum in FGD Plants Seminar, Birdcage Walk, Westminster, 1992.
- Wilson, R, Colome, S D, Spengler, J D and Wilson, D G**, "Health Effects of Fossil Fuel Burning", Ballinger Publishing Company, Cambridge, Massachusetts, 1980.
- World Energy Council Report**, "International Energy Data", Mediaprint, London, January 1993.
-

## APPENDIX 1

## SIGNATORIES OF THE SULPHUR DIOXIDE PROTOCOL

**Table A1.1:** Signatories of the sulphur dioxide protocol to the UN Convention on Long-Range transboundary air pollution (Vernon, 1990)

No.	List of Countries	No.	List of Countries
1	Austria	11	Hungary
2	Belgium	12	Italy
3	Byelorussian SSR	13	Leichtenstein
4	Canada	14	Luxembourg
5	Czechoslovakia	15	Netherlands
6	Denmark	16	Norway
7	Finland	17	Sweden
8	France	18	Switzerland
9	Federal Republic of Germany	19	Ukrainian SSR
10	German Democratic Republic	20	USSR

## A.1 CURRENT LEGISLATION LIMITS WORLD WIDE

**Table A1.2:** Large Combustion Plant Ordinance (LCPO) European Community Limits for Bituminous Coal (Bosch and Hansen, 1997)

FURNACE OUTPUT	TYPE OF PLANT		NO <sub>x</sub> LIMIT	NO <sub>x</sub> LIMIT	SO <sub>2</sub> LIMIT	SO <sub>2</sub> LIMIT	DUST
			mg/Nm <sup>3</sup>	ppm	mg/Nm <sup>3</sup>	ppm	mg/Nm <sup>3</sup>
50 – 100	new	dry	800	596	2000	685	50
		wet	1800	1343	2000	685	50
	old	dry	1300	970	2000	685	125
		wet	2000	1493	2000	685	125
100 – 300	new	dry	800	596	2000	685	50
		wet	1800	1343	2000	685	50
	old	dry	1300	970	2000	685	125
		wet	2000	1493	2000	685	125
> 300	new	dry	800	596	400	136	50
		wet	1800	1343	400	136	50
	old	dry	1300	970	400	136	125
		wet	2000	1493	400	136	125

<b>Where:</b>	<b>old</b>	Refers to old plant (from 1980 -1993)
	<b>new</b>	Refers to new plant (from 1993 - 1999)
	<b>wet</b>	Refers to wet bottomed boilers
	<b>dry</b>	Refers to dry bottomed boilers (Eskom boilers are dry bottomed)

**Table A1.3:** Provisions of the European Communities Large Combustion Plant Ordinance (LCPO): emission targets for existing plants (Vernon, 1989)

European Communities	Emission targets compared with 1980 emissions		
	1993	1998	2003
Belgium, France, Germany, Netherlands	-40	-60	-70
Denmark	-34	-56	-67
Greece	+6	+6	+6
Spain	0	-24	-37
Ireland	+25	+25	+25
Italy	-27	-39	-63
Luxemborg	-40	-50	-60
Portugal	+102	+135	+79
UK	-20	-40	-60

**Table A1.4:** Current First World Limits for Sulphur Dioxide (Makansi, 1995)

Country	Legislation/Comments	SO <sub>2</sub> Limit mg/Sm <sup>3</sup>	SO <sub>2</sub> Limit ppm
Germany	Most stringent world wide	400	136
USA	Phase 2 Clean Air Act	1480	1100
USA	New Stationary Source Act	740	552

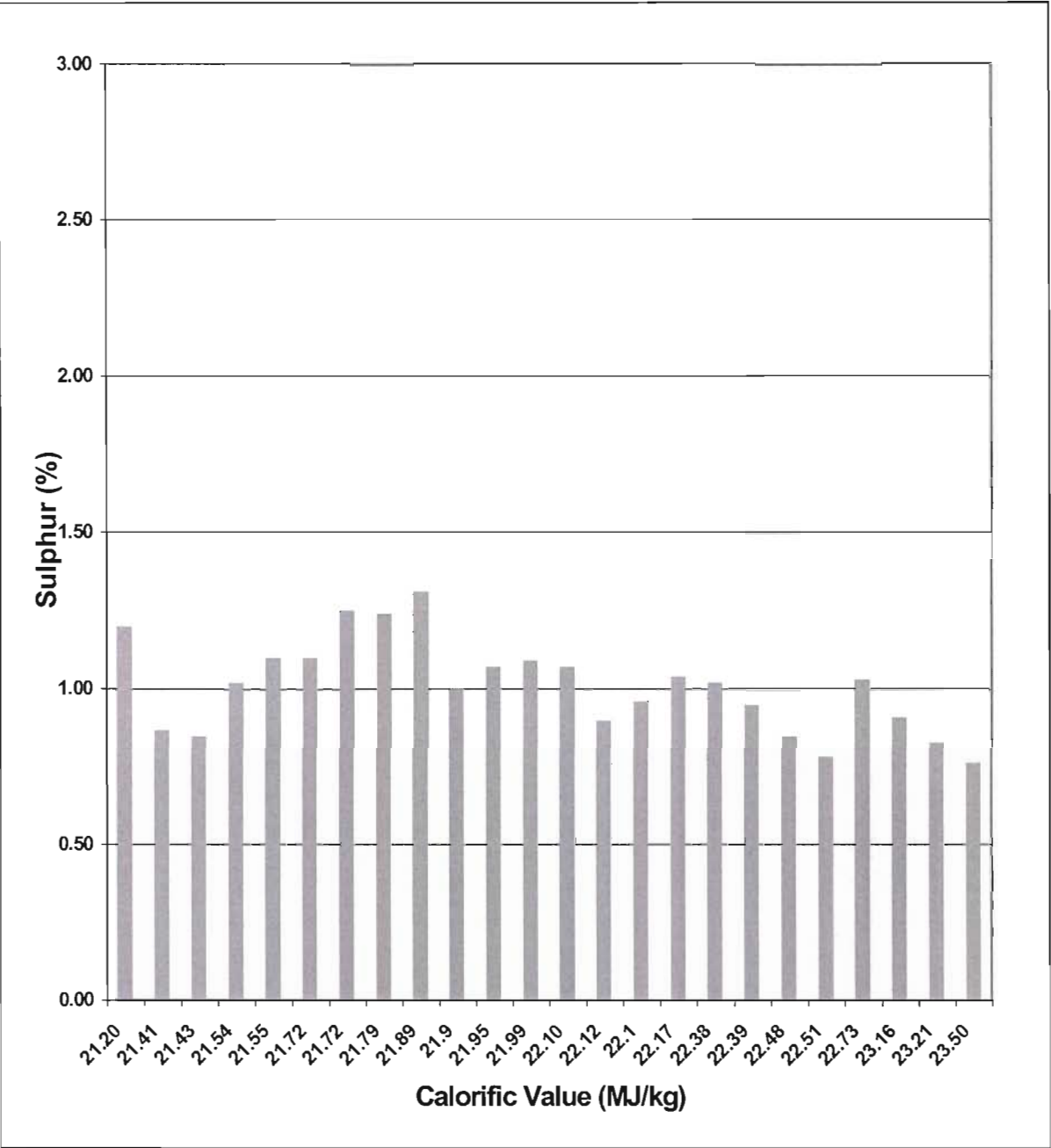
## A.2 COMPARISON BETWEEN SOUTH AFRICAN AND WORLD ENERGY COUNCIL MEMBER SO<sub>x</sub> AND NO<sub>x</sub> EMISSIONS

Table A1.5 is a comparison of South African SO<sub>x</sub> and NO<sub>x</sub> emissions per terawatt hour (tons/TWh) to other World Energy Council (WEC) member countries. The figures are for hard coal power generation emissions and electricity supply from coal sources.

**Table A1.5:** Table of comparison between WEC Members SO<sub>x</sub> emissions, NO<sub>x</sub> emissions and electricity generated from hard coal power plant (Bosch and Hansen, 1997).

	1990 SO <sub>x</sub> emissions	1990 NO <sub>x</sub> emissions	1990 electricity	1990 SO <sub>x</sub> emissions	1990 NO <sub>x</sub> emission s
WEC Member Country	'000 tons	'000 tons	TWh	tons SO <sub>x</sub> /TWh	tons NO <sub>x</sub> /TWh
Germany	210	287	318	0,660	0,903
Netherlands	49	52	28	1,767	1,857
Korea	65	57	20	3,260	2,850
Italy	176	138	35	5,016	3,943
Finland	52	43	10	5,200	4,300
Israel	54	38	10	5,429	3,810
France	210	88	32	6,550	2,750
United States	13600	6400	1668	8,154	3,837
South Africa	1250	435	138	9,058	3,152
Poland	1447	475	131	11,044	3,626
United Kingdom	2346	687	207	11,333	3,319
Ireland	74	30	6	12,400	5,000

**Source:** SO<sub>x</sub> emissions, NO<sub>x</sub> emissions, Electricity generation: World Energy Council Report, 1993)



**Figure A1.1:** Relationship between the Calorific Value and Sulphur Contents of various South African Coals (Tilly, 1996)

## APPENDIX 2

## GASEOUS AND PARTICULATE EMISSIONS

**Table A2.1:** Some Gaseous and Particulate Substances from Coal Combustion  
(Stenby, Hoskins and Donaldson, 1990)

Substance	Toxicity		Sources of Pollution	Comments
	Acute	Chronic		
Particulates				
Nitrates, nitrites	Increases infant susceptibility to lower respiratory infection due to conversion of nitrates to nitrites.	May combine with amines to form carcinogenic nitrosamines, also mutagenic and teragenic. Nitrate is a direct animal carcinogen.	Conversion of NO <sub>2</sub> to nitrates and nitrites, therefore primary sources are NO <sub>x</sub> emissions from fossil fuel combustion, fertiliser production, chemical plants, auto and industrial emissions.	
Organic matter	Unknown for many compounds. Specific toxicity for others.	Long-term is potentially carcinogenic and mutagenic.	Fossil fuel direct and indirect use in combustion, refining, plastics, tars, coking and chemical production.	Higher concentrations likely to be associated with non-direct combustion of fuels.
Arsenic (oxide forms)	Effects large to small depending on form and route of exposure.	Carcinogens, and teratogenic cumulative poison.	Weathering, mining and smelting. Coal combustion. Pesticides and detergents.	

Substance	Toxicity		Sources of Pollution	Comments
	Acute	Chronic		
Particulates				
Beryllium	Short term poison at high concentrations, especially toxic by inhalation.	Long term synthetic poison at low concentrations, carcinogenic in experimental animals.	Industrial: combustion of coal, rocket fuels.	
Cadmium	Very toxic at high concentrations to animals and aquatic life. Toxic by all routes of exposure.	Possible carcinogen. Cumulative poison. It is associated with hypertension, cardiovascular disease and kidney damage.	Weathering, mining and smelting especially of zinc; iron and steel industry; coal combustion; urban runoff; Phosphate fertilisers.	Chronic cadmium poisoning resulting in illness and death has occurred in Japan, where cadmium mobilised by mining contaminated daily diet.
Chromium	Hexavalent form is most harmful: skin and respiratory tract irritant.	Carcinogenic: workers engaged in manufacture of chromium chemicals have incidence of lung cancer, there is no evidence of risk in non-occupational exposure.	Emissions from industrial processes, including coal combustion, electroplating, tanning and dyes.	



Substance	Toxicity		Sources of Pollution	Comments
	Acute	Chronic		
Particulates				
Mercury	Methyl mercury and mercury fumes are very toxic; other forms are of variable toxicity.	Methyl mercury is a very toxic, cumulative poison; it affects the central nervous system.	Coal combustion, mining and smelting; weathering; volcanoes; sewage sludge; urban runoff; fungicides.	Environmental pollution leading to contamination of fish and shell fish caused illness and death in Japan; contamination of fish in the U. S. has caused closure of waters to commercial fishing.
Selenium	Soluble compounds are highly toxic.	Probable carcinogen; it is also found to be essential for life.	Coal combustion; industrial processes, mining and smelting; natural.	It interacts with other metals, increasing or decreasing toxicity.
Gases				
Sulphur dioxide	Increased respiratory impairment, morbidity and mortality in combination with particulates increases.	Increased respiratory diseases and decreased respiratory function with particulates.	Sulphur is contained in fossil fuels, smelters, and volcanoes.	Coal combustion presently represents between 60 % and 70 % of Uranium, solid waste, SO <sub>2</sub> emissions. In combination with particulates and humidity, the concentrations exceed the concentrations set by the EPA.

Substance	Toxicity		Sources of Pollution	Comments
	Acute	Chronic		
Gases				
Nitrogen dioxide	Increase respiratory infections i.e. irritation of respiratory tract.	Changes suspected in lung function; Emphysema.	Nitrogen contained in fossil fuels: coal, oil, gasoline combustion.	Effects on health questionable at EPA standards. Organically bound fuel nitrogen is a more important for coal NO emissions than for other fossil fuels.
Carbon monoxide	Behaviour changes, nausea, drowsiness, headaches, coma, death	Increased risk of coronary heart disease-arterial sclerosis suspected.	Incomplete combustion of fossil fuels, coal, oil, gas, gasoline combustion.	Evidence is not conclusive; effects short-term, reversible; increased tolerance with prolonged exposure; suspected long term effects.
Aromatic hydrocarbon	Fatigue , weakness, skin paresthesias (> 100 ppm)	Irritation, leukopenia and anemia. Certain compounds are mutagens and carcinogens.	A broad class of compounds naturally evolved from organic material, and from the evaporation and combustion of fossil fuels and other organic industrial chemicals.	Higher concentrations likely from less efficient and smaller boiler operation. Higher concentrations possible proximate to coal conversion facilities. Standard designed for photochemical oxidant control.

Substance	Toxicity		Sources of Pollution	Comments
	Acute	Chronic		
Gases				
Ozone	Increased respiratory infection, eye irritation, headaches, chest pain, impaired pulmonary function.	Unknown	Photochemical reactants involving hydrocarbons, nitrogen dioxide and other compounds in lower atmosphere. Reaction of atomic oxygen and oxygen in the upper atmosphere.	

**Table A2.2:** Characteristics of Particulate Control Technologies (Dacey and Cope, 1986)

Characteristics	Electrostatic Precipitator	Baghouse	Cyclone/ Multicyclone
<b>Technical</b>			
Efficiency	High	High	Moderate
Pressure drop	Low	Moderate	Moderate
Maximum operating temperature	High (up to 450°C)	Moderate (up to 260°C with conventional fabrics)	High (up to 800°C)
Sensitivity to gas temperature excursions	Performance possible affected, but no damage	Potential damage if fabric temperature limit is exceeded	Minor effects on performance
Effect of coal type	ESP performance best with high sulphur coal	Coal sulphur content is found to affect performance but is less than for ESP	Ineffective for collecting particles in the size range (5 to 10 µm)
Effect of ash type	Found to improve general performance of ESP	Minor to no effects on performance	Abrasive ashes can cause corrosion
Effect of moisture in ash	Can improve performance	Can cause caking and adversely affect performance	Leads to blockage, especially in multicyclones
Potential safety hazard	Electrical or fires	Fire hazard if the ash is too hot	none
Space requirements	None	Moderate to high	Moderate to high for cyclones, low to moderate for multicyclones
Aqueous waste stream	None	None	None
<b>Economic</b>			
Power requirement	Low for ESP	Low to moderate	Low to moderate
Capital cost	High	Moderate to high	Low
Operating and maintenance cost	Low	Moderate	Low

**APPENDIX 3****AIR QUALITY RESULTS FOR MAJUBA POWER STATION****Table A3.1:** Majuba Power Station Statistics

<b>Generating Capacity</b>	2139 MW	3 × 713 MW
	2007 MW on order	3 × 699 MW on order
<b>Cooling System</b>	Dry cooled Forced draft with direct cooling	48 fans per unit Total 288 fans Fan diameter - 9.1 m 45 × 80 × 85 m
<b>Emission Control</b>	Fabric filters	Design efficiency - 99.99%
<b>Stack Height</b>	250 m	3-flue with windshield

**A3.1 AIR QUALITY RESULTS**

All air quality data were obtained from Majuba's senior environmental officer, Bruce Smith. These results are tabulated below for the last two years from the month of January 1998 to December 1999. In addition, graphs were drawn to show how the various parameters varied on a daily basis, and comparisons were made between the emissions from the power station and the limits as set by the Chief Air Pollution Control Officer (CAPCO).

**Table A3.2:** Dust Emissions from Majuba from January 1998 to December 1999

Units Online	Date	Dust Conc.	Daily Emission	Total Month Acc.Emiss	Mass Ton/month (metric)	
	month	mg/Sm <sup>3</sup>	Avg. Ton/day (metric)	Ton (metric)	Target Ton/unit	Limit Ton/unit
2	Jan-98	10.90	1.74	1.74	85.31	170.6
	Feb-98	3.69	2.03	3.78	85.31	170.6
	Mar-98	5.11	2.82	6.59	85.31	170.6
3	Apr-98	8.43	5.49	12.08	85.31	255.9
	May-98	10.51	8.23	20.31	85.31	255.9
	Jun-98	9.78	9.29	29.60	85.31	255.9
	Jul-98	10.54	10.38	39.98	85.31	255.9
	Aug-98	11.22	5.76	45.75	85.31	255.9
	Sep-98	7.81	7.84	53.59	85.31	255.9
	Oct-98	10.19	2.14	55.72	85.31	255.9
	Nov-98	8.45	5.53	61.25	85.31	255.9
	Dec-98	7.67	2.03	63.28	85.31	255.9
	Jan-99	6.91	6.24	69.52	85.31	255.9
	Feb-99	7.53	4.42	73.94	85.31	255.9
	Mar-99	9.47	6.20	80.14	85.31	255.9
	Apr-99	10.12	2.01	82.15	85.31	255.9
4	May-99	11.09	4.72	86.87	85.31	341.2
	Jun-99	11.49	5.13	92.00	85.31	341.2
	Jul-99	9.42	4.04	96.04	85.31	341.2
	Aug-99	9.08	2.51	98.55	85.31	341.2
	Sep-99	9.77	4.09	102.64	85.31	341.2
	Oct-99	8.63	2.96	105.61	85.31	341.2
	Nov-99	8.91	4.82	110.43	85.31	341.2
	Dec-99	10.50	4.34	114.77	85.31	341.2

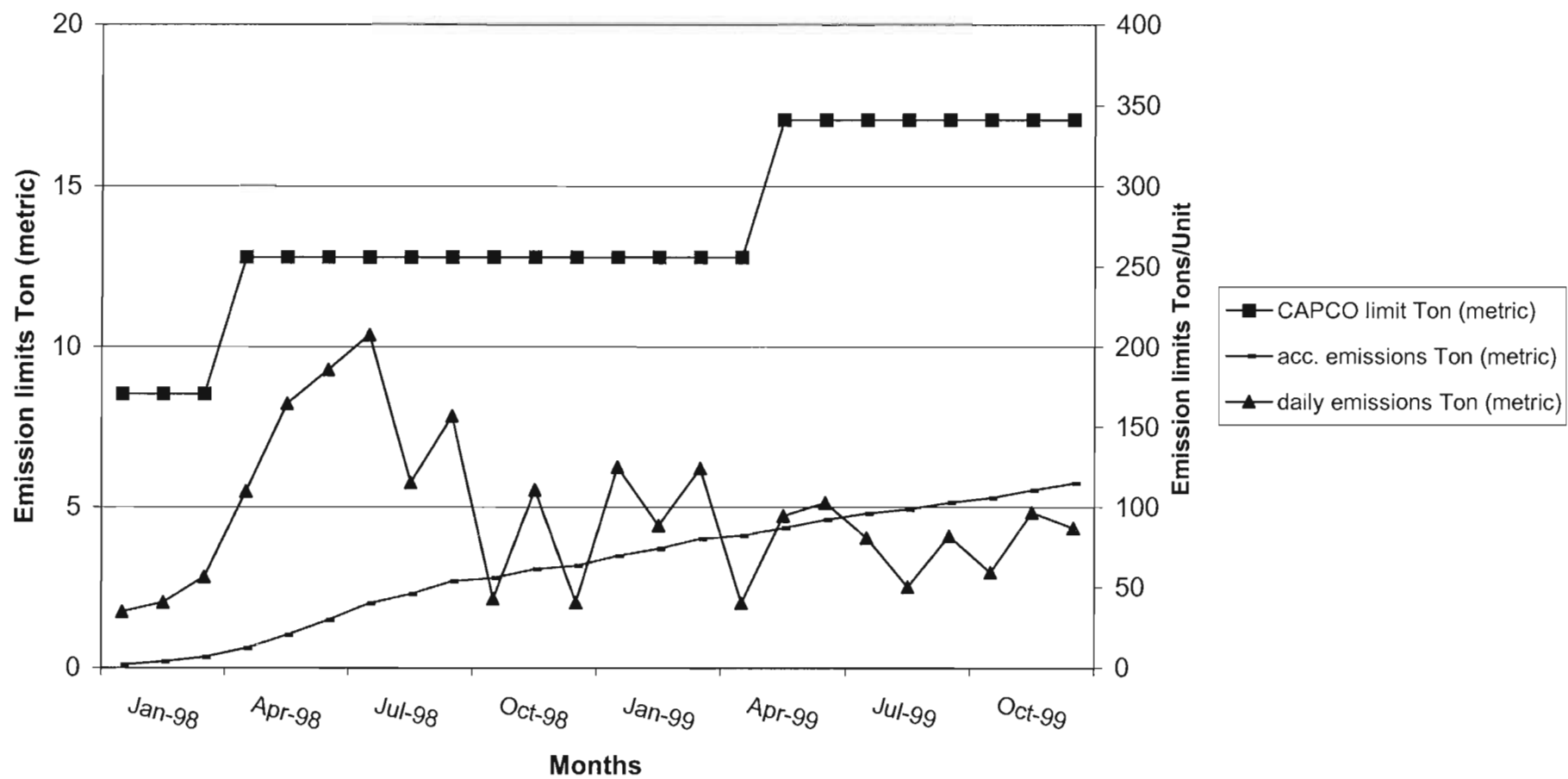


Figure A3.1: Dust emission limits for 1998 in comparison with 1999

- The average opacity is summed to as follows:

$$\text{Average Opacity} = \frac{\sum \text{Daily Emission}}{24} = 4.78 \text{ mg/Sm}^3$$

- Relative emissions (mg/Sm<sup>3</sup>) – denotes the emission limits and targets as set by the Chief Air Pollution and Control Officer (CAPCO).
- Given a volumetric flowrate of 637.02 m<sup>3</sup>/s for one unit, the target mass is calculated:

$$\begin{aligned}\text{Target mass} &= 637.02 \text{ m}^3/\text{s} \times 50 \text{ mg/Sm}^3 \\ &= 31851.105 \text{ mg/s} \\ &= 85.31 \text{ Ton (metric)/month}\end{aligned}$$

- The above data is based on emission limits for units 1, 2, 3, 4, 5 and 6 (Table A3.2).  
Limit (for 1 units) = 1 × 85.31 = 85.31 Ton (metric)/month  
Limit (for 2 units) = 2 × 85.31 = 170.72 Ton (metric)/month  
Limit (for 3 units) = 3 × 85.31 = 255.93 Ton (metric)/month  
Limit (for 4 units) = 4 × 85.31 = 341.23 Ton (metric)/month  
Limit (for 5 units) = 5 × 85.31 = 426.55 Ton (metric)/month  
Limit (for 6 units) = 6 × 85.31 = 511.86 Ton (metric)/month
-



**Table A3.3:** Comparison of dust concentrations from emission monitors (Sick® OMD 41 to 3 transmissometers) are correlated against the actual measurements

Date	Dust Concentration from Emission Monitors	Dust Concentrations	Actual Dust Concentrations $y=m\{(dust \times 0.16)+4\}+c$	CAPCO Limit
month	mg/Sm <sup>3</sup>	%	mg/Sm <sup>3</sup>	mg/Sm <sup>3</sup>
Jan-98	10.90	5.02	7.90	100
Feb-98	3.69	1.70	4.76	100
Mar-98	5.11	2.35	5.38	100
Apr-98	8.43	3.88	6.83	100
May-98	10.51	4.84	7.73	100
Jun-98	9.78	4.50	7.41	100
Jul-98	10.54	4.85	7.75	100
Aug-98	11.22	5.17	8.04	100
Sep-98	7.81	3.60	6.56	100
Oct-98	10.19	4.69	7.59	100
Nov-98	8.45	3.89	6.83	100
Dec-98	7.67	3.53	6.49	100
Jan-99	6.91	3.18	6.16	100
Feb-99	7.53	3.47	6.43	100
Mar-99	9.47	4.36	7.28	100
Apr-99	10.12	4.66	7.56	100
May-99	11.09	5.11	7.99	100
Jun-99	11.49	5.29	8.16	100
Jul-99	9.42	4.34	7.26	100
Aug-99	9.08	4.18	7.11	100
Sep-99	9.77	4.50	7.41	100
Oct-99	8.63	3.97	6.91	100
Nov-99	8.91	4.10	7.04	100
Dec-99	10.50	4.83	7.73	100
Sum =	217.22			

The dust readings in mg/Nm<sup>3</sup> is obtained from the Sick® OMD 41 to 3 transmissometers are correlated against the actual measurements to give a unique curve.

$$y = m\{(dust \times 0.16) + 4\} + c \quad (A3.1)$$

where:  $y$  = dust in mg/Nm<sup>3</sup> (dust = % reading)  
 $m \& c$  = test result constants (for Unit 1:  $m = 5.917$  &  $c = -20.516$ )

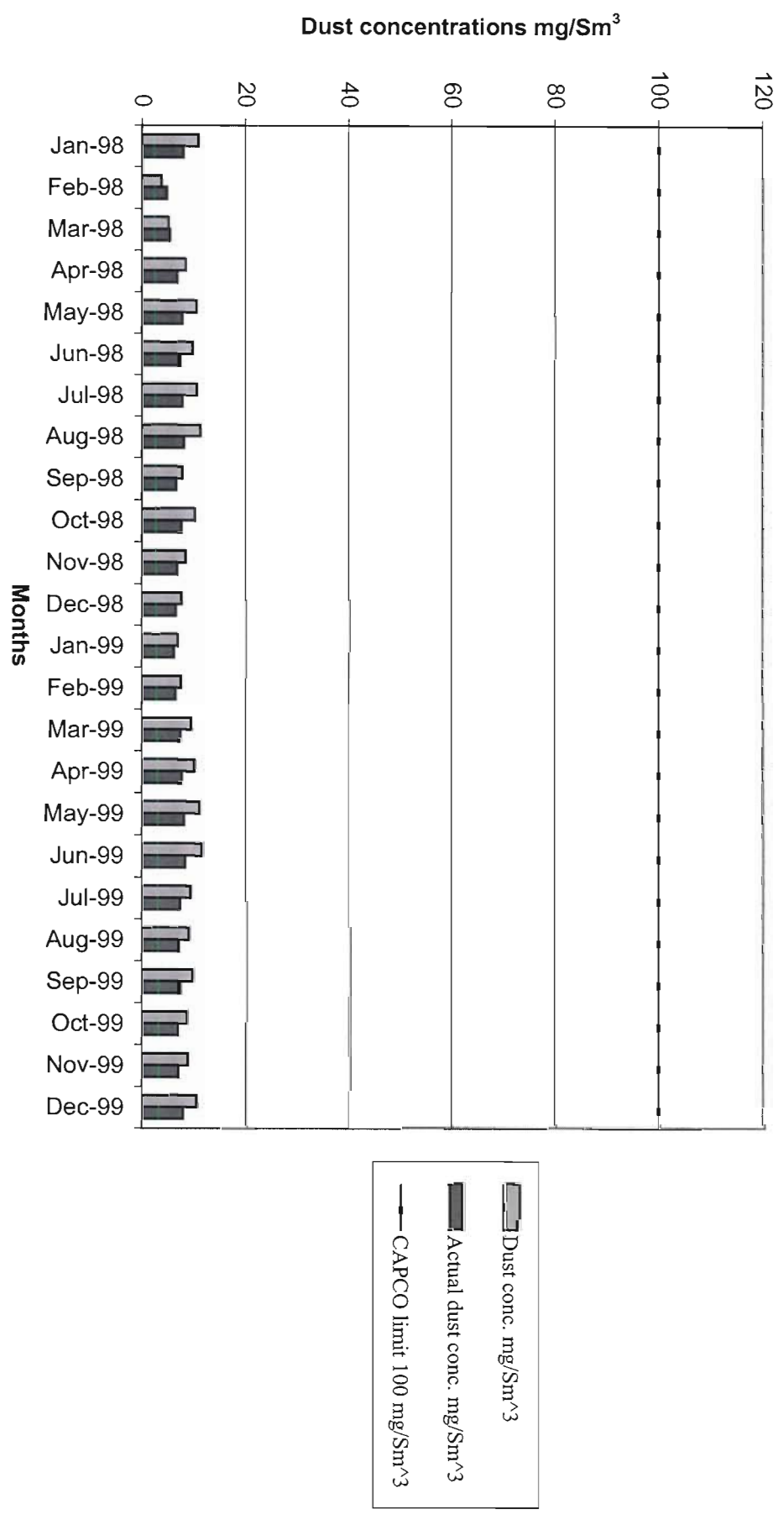


Figure A3.2: Dust concentrations from Majuba for 1998 and 1999

**A3.2 GASEOUS EMISSIONS****Table A3.4:** Ambient Sulphur Dioxide Concentrations

Month	Actual SO <sub>2</sub> Concentrations ( ppm )	Calculated SO <sub>2</sub> Concentrations (100 ppb)
Jan-98	621	8.63
Feb-98	598	8.31
Mar-98	692	9.61
Apr-98	668	9.28
May-98	702	9.74
Jun-98	836	11.62
Jul-98	560	7.78
Aug-98	762	10.59
Sep-98	756	10.50
Oct-98	800	11.12
Nov-98	826	11.47
Dec-98	663	9.21
Jan-99	810	11.25
Feb-99	761	10.57
Mar-99	803	11.15
Apr-99	815	11.32
May-99	806	11.19
Jun-99	797	11.06
Jul-99	731	10.15
Aug-99	766	10.64
Sep-99	822	11.42
Oct-99	656	9.11
Nov-99	660	9.16
Dec-99	679	9.43

- The actual sulphur dioxide concentration is calculated using the following formula:

$$Actual\ SO_2\ conc = \frac{Average\ SO_2}{number\ of\ units\ that\ were\ online}$$

Number of units online = 4

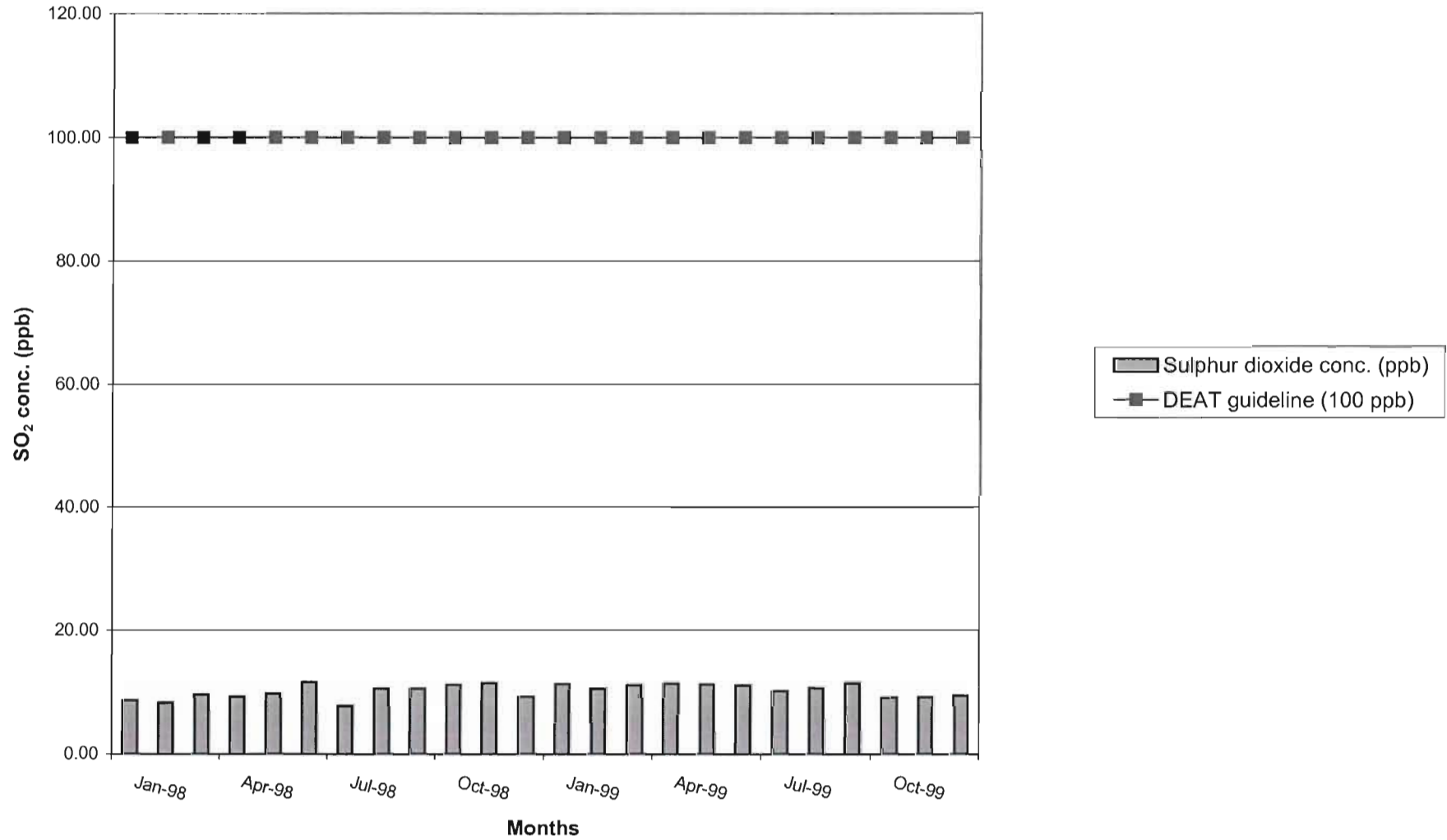


Figure A3.3: Ambient sulphur dioxide

The following table shows the relevant guideline values for sulphur dioxide as set by the CAPCO with the Department of Environmental Affairs and Tourism (DEAT).

**Table A3.5:** Maximum Allowable Concentration for Atmospheric Sulphur Dioxide  
(Source: Department of Environmental Affairs and Tourism in Pretoria)

	3-Minute average ppb	1 Hour average ppb	24 Hour average ppb	48 Hour average ppb	1 Month average ppb	Annual average ppb
National limits	600	300	100		50	30
Steering Committee Guideline values	267	191	57	38		

**Note:** Shaded area implies no data was available

For all calculation purposes, the hour average National Limits will be used.

Using a 1 hour average = 300 ppb

## APPENDIX 4

### COST ANALYSIS

#### A4. COST CALCULATIONS

A variety of methods was used to calculate costs. This was due to cost being difficult to obtain. Costs were converted using the exchange rate at a specific date.

#### A4.1 ECONOMIC EVALUATION FOR A WET SCRUBBING AND SPRAY DRY SCRUBBING PROCESS

The power plant chosen is a typical condensing power plant with a capacity of 600MW using coal as the fuel. The waste is assumed to be located in the power plant area. For this analysis the investment cost of a specific process is a constant regardless of the locality of where the desulphurisation plant is built. The operating costs depend on the price of electricity, steam and water, labour and considerably on the price of waste disposal.

##### A4.1.1 Basic values and assumptions

The basic values and assumptions applied to attain costs are found in Table A4.1.

**Table A4.1:** Basic values and assumptions

Description	Values/Prices
Capacity of the power plant	600MW
Gas emission	995m <sup>3</sup> /sec
Atmospheric pressure (P)	81kPa
Temperature (t)	145°C
Gas emission (V)	983m <sup>3</sup> /sec
Sulphur content of coal	1.08%
SO <sub>2</sub> – absorption in ash in the boiler and ESP	5%
SO <sub>2</sub> – emission unit	mg/MJ
SO <sub>2</sub> concentration in flue gas (C <sub>SO2</sub> )	799ppm
Turnover tax	not calculated
Building time interest	15%
Calculatory interest	15%
Life span of investment	50 years
Current (2000) exchange rate	R6.05 per US\$

The gas emission under normal conditions can be calculated as:

$$V_o = pV_{to}/(p_o t) = 81 * 995 * 293/(101.3 * 418) = 551 \text{ (m}^3\text{/sec)} \quad (\text{A4.1})$$

At STP ( $p_o = 101.3\text{kPa}$ ,  $t_o = 273^\circ\text{C}$ )

The density ( $\rho$ ) of flue gases under normal conditions is  $1.293\text{kg/m}^3$ . Therefore the mass of flue gas can be calculated as follows:

$$Q = V_o \rho = 551 * 1.293 = 712.44 \text{ (kg/sec)} \quad (\text{A4.2})$$

The mass of  $\text{SO}_2$  emitted is therefore:

$$\begin{aligned} Q_{\text{SO}_2} &= Q * C_{\text{SO}_2} * 10^{-6} = 712.44 * 799 * 10^{-6} \\ &= 0.5689 \text{ kg/sec} \end{aligned} \quad (\text{A4.3})$$

which is equivalent to  $8.89 \text{ g-mol/sec}$ .

The economic evaluation was carried out using USA Dollars and the current exchange rate of R6.05 per US\$.

#### A4.2 WET SCRUBBING (LIME) FGD

The removal coefficient of  $\text{SO}_2$  ( $f_{\text{SO}_2}$ ) was taken as 90% (0,9). The amount of  $\text{SO}_2$  removed is calculated by the following equation:

$$Q'_{\text{SO}_2} = Q_{\text{SO}_2} * f_{\text{SO}_2} = 8.89 * 0,9 = 8.001 \text{ g-mol/sec} \quad (\text{A4.4})$$

which is equivalent to  $44.17 \text{ tonnes of } \text{SO}_2 \text{ removed per day}$ .

According to the reaction:



the removal of  $1\text{g-mol}$  of  $\text{SO}_2$  requires  $1\text{g-mol}$  of calcium oxide ( $\text{CaO}$ ). The molecular weight ( $\text{mw}$ ) of  $\text{CaO}$  is 56, therefore the stoichiometric amount of  $\text{CaO}$  required is  $8.001 \text{ g-mol/sec}$  or  $38.71 \text{ tonnes/day}$ .

The conversion from  $\text{g-mol/s}$  to  $\text{tonnes/day}$  is:

$$\text{g-mol} (8.001) * 3600 * 24 * \text{mw} * 10^{-6}$$

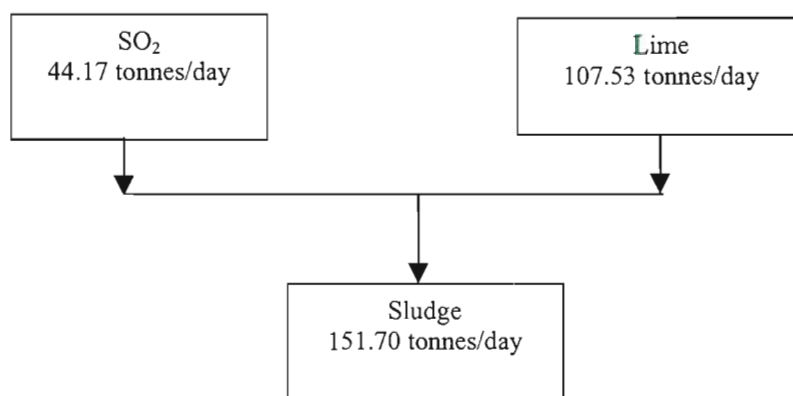
If the excess coefficient of  $\text{CaO}$  for the removal of 90% of  $\text{SO}_2$  is taken as 2, (the actual content of  $\text{CaO}$  in lime is 72 %), the daily lime consumption can be determined by:

$$Q_l^{\text{day}} = 2 Q_{\text{CaO}} / 0.72 = 2 * 38.71 / 0,72 = 107.53 \text{ tonnes/day} \quad (\text{A4.6})$$

The total amount of the dry sludge produced is:

$$Q_s = Q_{\text{SO}_2} + Q_l = 44.17 + 107.53 = 151.70 \text{ tonnes/day} \quad (\text{A4.7})$$

The mass-flow diagram for lime desulfurisation is shown in Figure A4.1.



**Figure A4.1:** Mass-flow diagram of lime sulphur dioxide removal method.

The evaluation was carried out, taking into account the above-mentioned items (Tilly, 1996). The extra costs of construction and maintenance of the power plant were not taken into account.

#### A4.2.1 INVESTMENT COSTS

The investment costs is calculated using the equation 7.1, Section 7.4.1, pp 7-3.

$$I = a * (b + P)^c$$

where:  $P$  = firing rate (MW) and  $a$ ,  $b$  and  $c$  constants.

Values for constants  $a$ ,  $b$  and  $c$  were taken from Table 7.2, Chapter 7, pp 7-3.

Using an exchange rate of 1FIM = \$0.173 and 1R = \$6.05

$$\begin{aligned}
 \text{Investment Cost of Wet Scrubber} &= \text{R}105.69/\text{MW} \\
 &= \text{R}105.69/\text{MW} * 600\text{MW} * 24 * 365 \\
 &= \text{R } 555.51 \text{ M}
 \end{aligned}$$

#### A4.2.2 OPERATING COSTS

The operating costs are divided into the fixed operating and variable operating costs.

##### FIXED OPERATING COSTS INCLUDE:

##### Labor

- 8 operators/year at R50/hr
- Assumed each operator worked for 8 hours/day, 300 days/year
- Same labor rate for all processes



$$\begin{aligned}\text{Labor Cost} &= 8 * \text{R}50/\text{hr} * 24\text{hr}/\text{day} * 300 \text{ day}/\text{year} = \text{R}2880\,000/\text{year} \\ &= \text{R}2.88 \text{ M}/\text{year}\end{aligned}$$

### Maintenance

- 2% of Investment cost.

$$\text{Wet scrubbing} = 0.02 * 555.51 = \text{R}11.11\text{M}/\text{year}$$

## VARIABLE OPERATING COSTS

### Lime Cost

$$\begin{aligned}\text{Lime cost} &= \text{R } 107.53 \text{ tonnes}/\text{day} * \text{R } 337.70 / \text{tonne} * 365 \text{ days}/\text{year} \\ &= \text{R } 13.25 \text{ M}/\text{year}\end{aligned}$$

### Waste Cost

This costs is mainly caused by the treatment of desulphurisation waste. Some of the parameters considered is the distance the waste has to be transported and dumping area costs if necessary.

The waste costs is estimated using equation 7.2, section 7.4.2, pp 7-3.

$$\begin{aligned}W &= ((A + B + C) \times D) \times M_w \\ W &= ( (10 + 2.50) * 1.5) * 107.53 \text{ t}/\text{day} * 365 \text{ day}/\text{year} = \text{R}0.74 \text{ M}/\text{t}/\text{year}\end{aligned}$$

### Fly Ash

The commercial value of fly ash is calculated using equation 7.3, Section 7.4.2, pp 7-4.

$$\begin{aligned}\text{FA} &= e \times m_{\text{fa}} \\ \text{FA} &= \text{R}30/\text{ton} * 80 \text{ ton}/\text{hr} * 24 \text{ hr}/\text{day} * 365 \text{ day}/\text{year} = \text{R}21.04 \text{ M}/\text{year}\end{aligned}$$

### Water

- Approximately  $80\text{m}^3/\text{hr}$  water is required at the 1994 average  $\$0.213/\text{m}^3$

$$\text{Water cost} = 80\text{m}^3/\text{hr} * \$0.213/\text{m}^3 * \text{R}6.05/\$ * 24 * 365 = \text{R}0.90 \text{ M}/\text{year}$$

### Electricity

- Assumed a price of  $\$0.03/\text{kWh}$  (Max and Klaus, 1990, pp 815). Therefore electricity price is using  $1\$ = \text{R}6.05$ :

$$(\$0.03/\text{kWh} * \text{R}6.05/\$ * 24 * 365 * 600) = \text{R}0.95 \text{ M}/\text{year}$$

## Steam

- For the design specifications given above, the cost of steam lies in the range of \$1.50 – 3.20/1000lb, (Max and Klaus, pp815).
- Assumed a steam cost of \$2.8/1000lb. This is equivalent to:

$$(\$2.8/1000\text{lb} * R6.05/\$ * 1\text{lb}/453.6\text{g} * 1 \times 10^6 \text{ g/ton}) = R37.34/\text{ton}$$

- Amount of coal used is approximately 3 000 000 tonnes/year, therefore steam cost is:

$$(3 \times 10^6 \text{ tonnes} * R37.34) = R112.02 \text{ M/year}$$

Total steam and electricity cost = R112.97 M/year

## Waste Water

The lime is initially mixed with water to form slaked lime, also known as calcium hydroxide ( $\text{Ca}(\text{OH})_2$ ). The basic chemical reaction is:



the removal of 1g-mol of  $\text{SO}_2$  requires 1g-mol of water ( $\text{H}_2\text{O}$ ). The molecular weight (mw) of  $\text{H}_2\text{O}$  is 18, therefore the stoichiometric amount of  $\text{H}_2\text{O}$  required is 8.001 g-mol/sec or 12.44 tonnes/day. The conversion from g-mol to tonnes/day is:

$$\text{g-mol} (8.001) * 3600 * 24 * \text{mw} * 10^{-6}$$

If the excess coefficient of  $\text{H}_2\text{O}$  for the removal of 90% of  $\text{SO}_2$  is taken as 2, (the actual content of  $\text{H}_2\text{O}$  in lime is 28 % since the percentage of  $\text{CaO}$  in lime is 72%), the daily lime consumption can be determined by:

$$\begin{aligned} Q_l^{\text{day}} &= 2 Q_{\text{H}_2\text{O}} / 0.28 = 2 * 12.44 / 0.28 = 88.88 \text{ tonnes/day} \\ &= 32\,441.2 \text{ tonnes/year} \end{aligned}$$

Waste water cost = \$0.692/t, based on 1990 price estimate (Valimaki,1990). This price has been escalated to using the present day conversions, 1\$ = R6.05

$$\text{Waste water cost/ton} = \$0.692 * 6.05 = R\,4.187/\text{tonnes}$$

For 32 441.2 tonnes/year

$$\begin{aligned} \text{Waste water cost} &= R4.187/\text{tonnes} * 32\,441.2 \text{ tonnes/year} \\ &= R\,135818.33/\text{year} \\ &= R0.14 \text{ M/year} \end{aligned}$$

### A4.3 SPRAY DRY SCRUBBING

#### A4.3.1 INVESTMENT COST

The investment costs is calculated using the equation 7.1, Section 7.4.1, pp 7-3.

$$I = a * (b + P)^c$$

where:  $P$  = firing rate (MW) and  $a$ ,  $b$  and  $c$  constants.

Values for constants  $a$ ,  $b$  and  $c$  were taken from Table 7.2, Chapter 7, pp 7-3.

Using an exchange rate of 1FIM = \$0.173 and 1R = \$6.05

$$\begin{aligned} I &= a * (b + P)^c \\ I &= 0.560 * (60 + 600)^{0.770} = 83.03 \text{ FIM/MWh} \\ &= \text{R } 86.90 / \text{MWh} * 600 \text{ MW} * 24 * 365 \\ &= \text{R } 456.75 \text{ M} \end{aligned}$$

#### A4.3.2 OPERATING COST

The operating costs are divided into the fixed operating and variable operating costs.

##### FIXED OPERATING:

##### Labor

- 8 operators/year at R50/hr.
- Assumed each operator worked for 8 hours/day, 300 days/year
- Same labor rate for all processes

$$\begin{aligned} \text{Labor Cost} &= 8 * \text{R}50/\text{hr} * 24\text{hr}/\text{day} * 300 \text{ day}/\text{year} = \text{R}2880 \text{ 000}/\text{year} \\ &= \text{R}2.88 \text{ M}/\text{year} \end{aligned}$$

##### Maintenance

- 2% of Investment cost.

$$\text{Spray dry scrubbing} = 0.02 * \text{R } 456.75 \text{ M} = \text{R } 9.14 \text{ M}/\text{year}$$

##### VARIABLE OPERATING COSTS

- Assumed all variable costs to be the same as the wet scrubbing process except the sorbent and wastewater cost.

##### Lime Cost

Assumed 200 tonnes/day of lime were utilized.

$$\begin{aligned}\text{Lime cost} &= 200 \text{ tonnes/day} * \text{R}337.70/\text{tonne} * 365 \text{ days} \\ &= \text{R } 24.65 \text{ M/year}\end{aligned}$$

**Wastewater**

There is no wastewater treatment because the water is completely evaporated in the absorber.

**A4.4 SELECTIVE CATALYTIC REDUCTION**

The capital costs of SCR was calculated using equation 7.4, section 7.5, pp 7-6.

$$C1 = 1.27 \times MWe^{0.71}$$

Based on this expression, the estimated capital cost of six 600 MWe units would be (calculated on the 2000 DEM):

$$C1 = 1.27 \times 3600^{0.71}$$

$$C1 = 425.38 \text{M. DEM}$$

Using exchange rates of 1DEM/0.53\$ and R6.05/\$ (2000)

$$C1 = \$222.78 \text{M}$$

$$C1 = \text{R}1347.82 \text{M}$$

**Capital Cost = R1336.78M**

There was not much information available with regards to running costs of the plant. The costs below are merely estimates (Lowe and Perslweig, 1990).

**Estimated running costs**

Tail gas installations: **2 – 4c/kWh for approximately 600 MWe**

High dust installations: **4 – 5c/kWe for approximately 600 MWe**

Assuming we are using a high dust installation with an operating cost of 5,5M/kWh. This translates to approximately 2,04c/kWh.

**A4.5 CIRCULATING FLUIDISED BED (CFB)**

Since Lurgi had not built a plant to this size, they have provided an estimated budget price as follows (Tilly, 1996):

Capital costs: 75% of equivalent wet FGD

Operating costs: 130% of equivalent wet FGD

---

Using the costs calculated earlier for a wet FGD (Section A4.2.1, pp A4.3), the costs of a CFB plant are:

Capital costs =  $0.75 \times \text{R } 555.51 \text{ M/year}$

**Capital costs = R 416.33 M/year**

Operating costs =  $1.30 \times 120.95$

**Operating costs = R 157.24 M/year**

## A4.6 ECONOMIC EVALUATION OF ELECTROMEMBRANE DESULPHURISATION

### A4.6.1 Assumptions for Economic Calculations

Due to the fact that this is a new technology, very limited data was available for calculation purposes. Hence an economics evaluation was done. Dr Vladimir Linkov of the Sasol Centre for Chemistry, Potchefstroom University for CHE, conducted investigations into Electromembrane desulphurisation.

The following assumption were used for the economic evaluation:

- Power of the unit (N) 600 MW
- Temperature (t) 141 °C
- Atmospheric pressure (P) 83 kPa
- Gas emission (V) 995 m<sup>3</sup>/sec
- SO<sub>2</sub> concentration in flue gas (C<sub>SO2</sub>) 797 ppm

The gas emission under normal conditions can be calculated as:

$$V_o = pV_t/(p_o t) = 83 * 995 * 293/(101.3 * 414) = 539 \text{ (m}^3\text{/sec)} \quad (\text{A4.9})$$

At STP ( $p_o = 101.3 \text{ kPa}$ ,  $t_o = 273^\circ\text{C}$ )

Considering that the density ( $\rho$ ) of flue gases under normal conditions is  $1.293 \text{ kg/m}^3$  (Perry, 1984) the mass of flue gas can be calculated as follows:

$$Q = V_o \rho = 539 * 1.293 = 697 \text{ (kg/sec)} \quad (\text{A4.10})$$

The mass of SO<sub>2</sub> emitted is therefore:

$$\begin{aligned} Q_{\text{SO}_2} &= Q * C_{\text{SO}_2} * 10^{-6} = 697 * 797 * 10^{-6} \\ &= 0.5556 \text{ kg/sec} \end{aligned} \quad (\text{A4.11})$$

which is equivalent to  $8.68 \text{ g-mol/sec}$ .

The economic evaluation was carried out using USA Dollars and the 1995 exchange rate of R3.6 per US\$. This was updated with the current (January 2000) exchange rate of R6.05 per US\$.

The removal coefficient of  $\text{SO}_2$  ( $f_{\text{SO}_2}$ ) was taken as 90% (0.9). The amount of  $\text{SO}_2$  removed is calculated by the following equation:

$$Q'_{\text{SO}_2} = Q_{\text{SO}_2} \cdot f_{\text{SO}_2} = 8.68 \cdot 0.9 = 7.812 \text{ g-mol/sec} \quad (\text{A4.12})$$

which is equivalent to 44.17 tonnes of  $\text{SO}_2$  removed per day.

The removal coefficient of  $\text{SO}_2$  ( $f_{\text{SO}_2}$ ) is taken as 0.9 (90%). According to the following reaction,



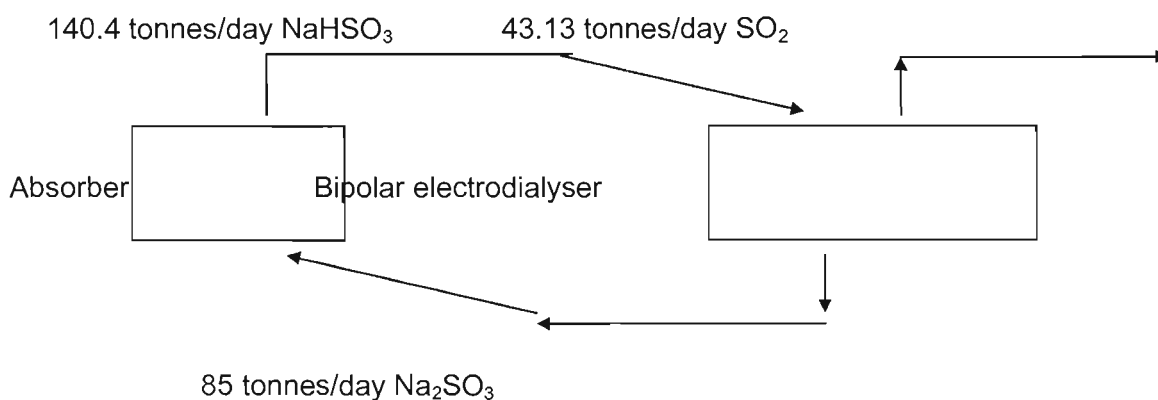
the removal of 1mol of  $\text{SO}_2$  requires 1mol sodium sulphite. The  $\text{Na}_2\text{SO}_3$  consumption can be determined as follows:

$$Q_c = Q_{\text{SO}_2} = 7.812 \text{ g-mol/sec} \quad (\text{A4.14})$$

$$\text{or } 7.812 \cdot \text{mw Na}_2\text{SO}_3 = 984.3 \text{ g/sec}$$

$$\text{or } 3\,543.5 \text{ kg/h.}$$

The mass-flow diagram for the electromembrane method of  $\text{SO}_2$  removal is shown in Figure A4.2.



**Figure A4.2:** Mass-flow diagram for electromembrane method of  $\text{SO}_2$  removal.

After an analysis of the methods and apparatus used for gas adsorption had been carried out, it was decided that mesh plate absorbers would be the most effective in the sulphur dioxide removal process due to their low costs and effectiveness. Also, these absorbers require minimum running costs.

#### A4.6.2 Evaluation of the Plate Absorber Efficiency

The following equations are used to evaluate the plate absorber efficiency.

- *Limiting gas flow rate* ( $W_{lim}$ ):

$$W_{lim} = 0.05 \sqrt{\rho_l / \rho_g}, \text{ m/sec} \quad (\text{A4.15})$$

where:  $\rho_l$  = liquid density, kg/m<sup>3</sup>,  
 $\rho_g$  = gas density, kg/m<sup>3</sup>

If it is accepted that at 60°C,  $\rho_l$  is 990kg/m<sup>3</sup> and at 70°C,  $\rho_g$  is 1.01kg/m<sup>3</sup>, the  $W_{lim}$  is 1.5m/sec. The operating gas flow ( $W_w$ ) in the column is therefore:

$$W_w = 0.9 * 1.5 = 1.35\text{m/sec} \quad (\text{A4.16})$$

- The *total free cross-section* required of the column absorbers:

$$f_{ta} = v_{sec} / W_w, \text{ m}^2 \quad (\text{A4.17})$$

where  $v_{sec}$  is the total flue gas emission per second. If this value is taken as 995m<sup>3</sup>/sec, then

$$f_{ta} = 995/1.35 = 77.3\text{m}^2.$$

If the free cross-section of one absorber has a diameter (D) of 3.6m, then:

$$f_a = \pi D^2 / 4 = 10.2\text{m}^2 \quad (\text{A4.18})$$

- The *required number of elements* with the mesh overflow plates:

$$Z_e = f_{ta} / f_a = 77.3 / 10.2 = 76 \text{ elements} \quad (\text{A4.19})$$

- The *distance between* the absorber plates:

$$h_p = (1.8 - 2.2) * \Delta P_p, \text{ mm} \quad (\text{A4.20})$$

where:  $\Delta P_p$  = hydraulic resistance of one mesh plate, Pa

- The *hydraulic resistance* of one mesh plate is calculated from a number of components according to :

$$\Delta P_p = \Delta P_d + \Delta P_s + \Delta P_\delta, \text{ Pa} \quad (\text{A4.21})$$

a)  $\Delta P_d$  = hydraulic resistance of a dry plate:

$$\Delta P_d = \Sigma W_g^2 \rho_g / 2 F_d^3$$

$$\begin{aligned}
 &= 1.5 * 2.85 * 0.65 / 2 * 0.2^2 \\
 &= 50.1 \text{ Pa}
 \end{aligned}
 \tag{A4.22}$$

b)  $\Delta P_s$  = hydraulic resistance of the gas-liquid layer on the plate

$$\Delta P_s = h_s \rho_l g = 0.03 * 990 * 9.8 = 295 \text{ Pa} \tag{A4.23}$$

c)  $\Delta P_\delta$  = hydraulic resistance resulted from surface tension forces:

$$\begin{aligned}
 \Delta P_\delta &= 4 \sigma / d_{eq} = 4 * 2 * 10^{-3} / 0.012 = 6.7 \text{ Pa} \\
 \text{Therefore: } \Delta P_p &= 50.1 + 295 + 6.7 \approx 353 \text{ Pa}
 \end{aligned}
 \tag{A4.24}$$

- The *minimum distance* between plates is:

$$h_{min} = 2 * 353 \approx 700 \text{ mm} \tag{A4.25}$$

Based on the above calculations, the absorber has the following features:

- Number of elements 15
  - Height (15 \* 0.7) 10.5m.
  - Hydraulic resistance of the absorber ( $\Delta P_a$ ) 353Pa
  - Number of absorbers 6 including 1 reserve
- ( $n = z/n_a = 76/15$ )

#### A4.6.3 CALCULATION OF THE ABSORBER AND MAINTENANCE COSTS

Because of the new process of sorption of  $\text{SO}_2$  from flue gas by  $\text{Na}_2\text{SO}_3$ , there is no standard equipment available for this process.

An absorber with the following specification was chosen (Price List No. 23-03, 1994):

*Absorber No. 02-0480 (Code OKP 36 1211 3029)*

*Type 101.2199.00000*

*Technical Conditions 26-15-003-78)*

- Material of main construction units: Steel 09G2S, 12X18H10T
- Mass: 45950 kg
- Life expectancy: 20 years
- Total cost for 1 absorber: \$0.324M (R1.936M)
- Total price for 6 absorbers: \$1.944M (R11.761M)

The supply of  $\text{Na}_2\text{SO}_3$  solution requires 2 pumps with a productivity of  $5.0 \text{ m}^3/\text{h}$  (1 reserve) (Tilly, 1996).

*Model - X-80-50-200A*

- Productivity:  $5 \text{ m}^3/\text{h}$
- Working distance: 50m



- Dimensions: 1290x504x570
- Mass: 360kg
- Power Consumption: 30kW
- Total cost of 1 unit: \$2 560.20
- Total cost of 2 units: \$5 120.40 (approximately R30 978.42)

The services of the following personnel are required for operating the absorbers and the pumps:

- operator
- 2 technicians
- laboratory chemist

#### A4.6.4 EVALUATION OF THE ELECTRODIALYSIS UNIT

Data used for Initial Calculations:

The data used for the calculations is summarised below:

Quantity of removed SO <sub>2</sub> (Q <sub>SO2</sub> )	31.25kg-mol/h
Removal degree (f <sub>SO2</sub> )	0.9
Current yield (η)	0.95
Current density (A/m <sup>2</sup> )	100A/m <sup>2</sup>

According to Faradays Law, the surface area (S<sub>m</sub>) of coupled membranes required for the conversion of Na<sub>2</sub>SO<sub>3</sub>, is given by the following equation:

$$S_m = Q_{SO_2} * F * f_{SO_2} / \eta I \quad (A4.26)$$

$$S_m = 7934m^2$$

Taking into account the number of membranes, the following choice of electro dialysis unit was made:

*Type EDA - 1500x1000 (horizontal electrical field)*

- Dimensions: 1980x1616x1900mm
- Mass: 3125kg
- Electrodes: platinated titanium
- Membrane dimensions: 1492 x 992mm
- Number of membrane pairs: 300
- Flow-through mesh spacers with a separator net
- total cost of the standard unit: \$0.1104M (approximately R0.396M)

After further research, a different electro dialysis unit was chosen, as it was already equipped with pumps, power sources and all necessary automation systems. The specifications are summarised as follows:

Type - EDU1-400x6

- Total membrane area:  $(S_{cm}) 453m^2$
- Dimensions: 5500x2300x1900
- total cost of this unit: \$0.112M (approximately R0.402M).

The required number of EDU1-400x6 electro dialysis units for the process is:

$$N_e = S_m / S_{cm} = 7934 / 453 = 18 \text{ pieces.} \quad (A4.27)$$

By including 2 reserve installations, the total number units required are 20.

#### 5.7.4.2 Calculation of $Na_2SO_3$ requirements.

The quantity of  $Na_2SO_3$  ( $q_{Na_2SO_3}$ ) required for the operation of absorbers can be determined from the height ( $h_p$ ) of the liquid layer on a plate, and the total area of the free cross-section ( $f_{ta}$ ) of all absorbers. So if  $h_p$  is 0.03m and  $f_{ta}$  is  $77.3m^2$ , then

$$q_s = h_p f_{ta} = 0.03m * 77.3m^2 = 2.2 m^3 \quad (A4.28)$$

The quantity of  $Na_2SO_3$  in electro dialysers,  $q^e$ , is determined by

$$q^e = 2 * S_m * \delta_m * f_m \quad (4.29)$$

$$q^e = 2 * 7934 * 0.001 * 0.68 = 10.8 m^3,$$

$$\begin{aligned} \text{where: } S_m &= \text{surface area} \\ \delta_m &= \text{distance between membranes} = 0.001 \text{ m} \\ f_m &= \text{membrane use coefficient} = 0.68 \end{aligned}$$

If the volume of the solution ( $q'$ ) in pipes, pumps and vessels is  $2m^3$ , then the total quantity of  $Na_2SO_3$  solution required is  $15.1 m^3$  based on the following equation.

$$q_t = + q^e + q' = 15.1 m^3 \quad (A4.30)$$

Because the concentration of the solution is 1M, and the molecular weight of  $Na_2SO_3$  is 126, the quantity of dry  $Na_2SO_3$  required for the system is:

$$q^{dry} = 15.1 * 1 * 126 = 1904kg. \quad (A4.31)$$

#### A4.6.5 CALCULATION OF ELECTRICAL ENERGY REQUIREMENTS.

The specific consumption of electrical energy (W) with units of W-h/g-eq, can be calculated from the following equation:

$$= U * F / \eta \quad (A4.32)$$

$$\begin{aligned} \text{Where: } U &= \text{voltage drop on the elemental cell} \\ \eta &= \text{current yield} \end{aligned}$$

The voltage drop is calculated as follows:

$$U = U_{mb} + U_k + U_s, \quad (A4.33)$$

where:  $U_{mb}$  = voltage drop across the bipolar membrane  
 $U_k$  = voltage drop across cation-exchange membrane  
 $U_s$  = voltage drop across the solution.

- The *voltage drop across the bipolar membrane* is determined as follows:

$$U_{mb} = 0.66 + i/\kappa_{mb}, \quad (A4.34)$$

The  $\kappa_{mb}$  is the surface conductivity of bipolar membrane with units of  $\text{Ohm}^{-1}/\text{cm}^2$ . It is determined as follows:

$$\kappa_{mb} = 0.012 \exp[(1/T - 1/293)(14200/R)] \quad (A4.35)$$

- The *voltage drop across the cation exchange membrane*, if the temperature is  $313^\circ\text{K}$  and  $U_{mb}$  is  $1.26\text{V}$ , will be as follows:

$$U_k = i / \kappa_k, \quad (A4.36)$$

The  $\kappa_k$  is the surface conductivity of the cation-exchange membrane is determined by the following equation.

$$\kappa_k = (0.097 + 0.026C) \exp [(1/T - 1/293)(-33900/R)] \quad (A4.37)$$

Therefore  $U_k$  is  $0.06\text{V}$ .

- The *voltage decrease in the solution* is determined by:

$$U_s = 2 i h / \kappa_s \quad (A4.38)$$

The specific electroconductivity ( $\kappa_s$ ) of  $1\text{M Na}_2\text{SO}_3$  is  $0.062\text{Ohm}^{-1}/\text{cm}^2$  and  $U_s$  is therefore  $0.032\text{V}$ .

The total decrease in voltage ( $U$ ) across an elemental cell is  $1.62\text{V}$  (from A4.33)

The specific consumption of electrical energy is  $45.7\text{W-h/g-mol}$  (from A4.31)

The total power consumption of all the units is  $1\,285\text{kW}$  based on the following equation.

$$N = W Q_{\text{SO}_2} f_{\text{SO}_2} = 45.7 * 31.25 * 0.9 = 1285 \text{ kW}.$$

**Table A4.2** Itemised cost analysis for sulphur dioxide removal from flue gasses by Electromembrane Desulphurisation Method

<b>I. CAPITAL COSTS</b>	<b>US\$ * 10<sup>3</sup></b>	<b>Rand * 10<sup>3</sup></b>
<i>Building for housing electro dialysis units.</i> <ul style="list-style-type: none"> <li>Number of units 20.</li> <li>Total area of the building 810m<sup>2</sup>,</li> <li>standard analogue T.P. 709-4-22.91.</li> <li>Equipped with crane Q=2 tonnes.</li> <li>Building height 7.8 m taking the crane into account.</li> </ul>	250.40	1514.92
<i>Chemically stable pumps</i> (pumps the working solution into electro dialysers) <ul style="list-style-type: none"> <li>Type X-80-50-200A.</li> <li>Power 30 kW. 1 in operation, 1 reserve.</li> </ul>	5.12	30.98
<i>Assembling of pumps</i> (Construction regulations, IV, 6-82, part 22)	1.004	6.07
<i>Electro dialysers</i> (with 2 reserves) <ul style="list-style-type: none"> <li>Production of Alma-Ata electromechanical plant Dimensions 5500x2300x1900.</li> <li>Cation-exchange membrane area 453m<sup>2</sup></li> <li>Bipolar membrane area 453m<sup>2</sup></li> </ul>	2595.34	15701.81
<i>Assembling of electro dialysers</i>	111.595	675.15
Pumps for solution delivery into absorbers, similar to 2.3	5.12	30.98
	1.004	6.07
<i>Absorbers, material</i> <ul style="list-style-type: none"> <li>Steel 09G2S - 12X18H10T</li> <li>Mass 45 950 kg (price list 23-03)</li> <li>Number of working absorbers - 5, reserve 1.</li> </ul>	1942.74	11753.58
<i>Assembling of absorbers</i>	134.185	811.82
<i>NaSO<sub>4</sub> required :</i> 1 904kg at \$1.38/kg (R4.97/kg)	2.627	15.89
<b>SUB-TOTAL FOR CAPITAL COSTS</b>	<b>\$5.049M</b>	<b>R30.56M</b>
<b>II. OPERATING COSTS</b>		
<i>Electrical energy consumed by electro dialysers: at 0.16c/kWh<sup>(1)</sup></i> <ul style="list-style-type: none"> <li>285 kW * 24 hours * 365 days = 11256600kW/h</li> </ul>	500.3	3026.82
<i>Electrical energy consumed by all pumps:</i> <ul style="list-style-type: none"> <li>30 kW * 24 hours * 365 days = 262 800kW/h</li> </ul>	11.680	70.66
<i>NaSO<sub>3</sub> losses</i> <ul style="list-style-type: none"> <li>10% of total quantity of Na<sub>2</sub>SO<sub>3</sub> therefore: 359kg x \$1.38/kg (R4.97/kg)</li> </ul>	0.495	2.99
<i>Land usage rates (standard rate at R36/m<sup>2</sup>)</i> (includes all buildings, pumps and absorbers) <ul style="list-style-type: none"> <li>Total area 1000 m<sup>2</sup></li> </ul>	10.00	60.50
<i>Personnel salary</i> (minimum state salaries are given) <ul style="list-style-type: none"> <li>5 operators</li> <li>2 technicians</li> <li>1 construction</li> <li>1 laboratory assistant</li> <li>7 electro dialyser maintenance</li> </ul>	22.59 7.028 3.514 2.410 29.618	136.67 42.52 21.26 14.58 179.19

<b>II. OPERATING COSTS</b>		
<i>Maintenance costs</i>		
• absorbers	97.137	587.68
• electrolysers	122.39	742.57
<i>Membrane exchange costs</i> (10% total membrane area)	98.241	594.36
<b>SUB-TOTAL FOR OPERATING COSTS</b>	<b>\$0.906M</b>	<b>R5.48M</b>
<b>III. AMORTISATION PAYMENTS</b>		
• building for housing electrolysers (operation time 50 years)	4.769	28.858
• absorbers (operation time 20 years)	97.137	587.68
• electrolysers (operation time 10 years)	259.534	1570.18
<b>SUB-TOTAL AMORTISATION</b>	<b>\$0.362M</b>	<b>R2.19M</b>