Air pollution control costs for coal-fired power stations

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Abstract

This report examines current costs for air pollution control at coal-fired power plants. The costs are highly specific to the site and performance. They depend on technical and financial factors such as degree of retrofit difficulty, unit size, capacity factor, removal efficiency required, labour cost and interest rate, which vary from country to country and from plant to plant. The cost for a specific power plant should be evaluated for site conditions. As a general guide, particulate control can add to the cost of electricity by 2% whilst FGD and SCR may increase the cost of electricity by about 10% and 6% respectively. Major reductions in the capital costs of air pollution control systems have been achieved. This is especially the case with FGD where the cost in the USA has decreased by 75% since the first installation in the early 1970s. The operating and maintenance costs of air pollution control equipment have also generally decreased. The decline in costs is due mainly to continued advances in technology and increased competition in the market.

Acronyms and abbreviations

CFB	circulating fluid bed
CIA	carbon-in-ash
COHPAC	compact hybrid particulate collector
CUECost	coal utility environmental cost
DTI	Department of Trade and Industry (UK)
EBA	electron beam with ammonia
EPA	Environmental Protection Agency (USA)
EPRI	Electric Power Research Institute (USA)
ESP	electrostatic precipitators
ETSU	Energy Technology Support Unit
FGC	flue gas conditioning
FGD	flue gas desulphurisation
FGR	flue gas recirculation
GSA	gas suspension absorption
LIFAC	limestone injection into furnace and activation of unreacted calcium
LSFO	limestone forced oxidation
LSNO	limestone natural oxidation
MCED	membrane contactor electro dialysis
NOx	nitrogen oxide and nitrogen dioxide (NO and NO_2)
PAF	phosphate ammonium fertiliser
SCA	specific collection area
SCR	selective catalytic reduction
SNAP	SO_2/NOx adsorption process
SNCR	selective non-catalytic reduction
SO_2	sulphur dioxide
US DOE	Department of Energy (USA)

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1 Introduction

Coal combustion can cause significant air pollution in countries relying on coal as a major energy source unless properly controlled. Most coal-fired power plants throughout the world are fitted with a type of particulate control device. The majority are using electrostatic precipitators (ESPs) which are highly efficient and cost effective. Flue gas desulphurisation (FGD) is an effective control of sulphur dioxide (SO_2) emissions from coal combustion. It is used in about 25% of the total capacity of coal-fired power plant. Technologies for control of nitrogen oxides (NOx) emissions include combustion modifications and flue gas treatments. These have been developed and applied on a commercial basis in a number of countries, accounting for around 40% of the total world capacity. In the last few decades, control technologies have undergone considerable developments to meet increasingly restrictive emission limits at minimum cost. The details have been reviewed in earlier IEA Coal Research reports (Soud, 1995, 2000; Soud and Fukasawa, 1996).

There have been major reductions in the capital costs of air pollution control systems. This is especially the case with FGD where the cost in the USA is reported to have decreased by 75% since the first installation in the early 1970s (Boward and Brinkmann, 1998). The operating and maintenance costs of air pollution control equipment have also generally decreased. The decline in costs is due mainly to continued advances in technology and increased competition in the market.

This report updates the 1995 report from the Clean Coal Centre on this topic (Takeshita, 1995). It examines the current costs for particulate, SO_2 and NOx emission controls at coal-fired power plants. The technologies discussed include commercially available ones and, where applicable, those under development. The costs for both new installations and retrofits, depending upon the availability of information, are included. Factors which affect capital and/or operating and maintenance costs are also identified. It should be noted, however, that the costs are highly site-specific, varying considerably from country to country and from plant to plant. The purpose of this report is to provide a general guide to the economics of air pollution control. A detailed unit-by-unit analysis is required in order to select the most cost-effective technology best suited to a specific power plant.

The report features five main chapters. Chapter 2 examines the costs for particulate control including electrostatic precipitators and fabric filters. It mainly discusses the costs of ESP upgrades as most existing coal-fired power plants are already fitted with the control device. Chapter 3 reviews the costs for FGD systems. This includes various wet processes, spray dry processes, sorbent injection processes, circulating fluid bed processes, and regenerable processes. In Chapter 4, the costs for NOx control technologies are detailed. It discusses combustion modification-based technologies (low-NOx burners, overfire air, flue gas recirculation, reburn) and flue gas treatments such as selective catalytic reduction (SCR) and selective non-catalytic reduction (SNCR). Combined SO_2/NOx control technologies, such as SO_2/NOx adsorption process and electron beam process, are detailed in Chapter 5. Finally, Chapter 6 examines the effects of the installations of air pollution control technologies on the cost of electricity generation.

2 Particulate control

Particulate emissions from coal combustion can cause significant air pollution in countries relying on coal as a major energy source unless they are properly controlled. In pulverised coal-fired power plants, the particulate emissions are controlled mainly by the use of ESPs, followed by fabric filters. In the last few decades, particulate control technologies have undergone considerable developments. These include improving performance to meet more stringent emission standards and/or reducing costs to compete in the market. Details are discussed in earlier IEA Coal Research reports (Soud, 1995; Wu, 2000a) and a number of recent papers (Crynack and Armiger, 1998; Porle and others, 1998; Grieco, 1997).

The costs for particulate control systems are site-specific and vary from country to country. The choice of technology mainly depends on the required emission limit and type of coal. ESPs are efficient and can meet a wide range of emission limits. The technology is well developed and relatively inexpensive. It is therefore the first choice in most cases. Bag filters offer alternatives to ESPs to meet the most stringent emission limits worldwide – less than 50 mg/m^3. Today's limits are tight enough that cyclones are no longer a viable option and thus are not discussed. The costs for ESPs and fabric filters are discussed first separately. This is then followed by a comparative cost study.

2.1 Electrostatic precipitators

In an ESP, a high-voltage is applied across the discharge electrodes and two earthed parallel collecting plates. The field strength around the discharge electrodes is sufficiently high that the flue gas undergoes an electrical breakdown and hence the corona is generated. The corona discharge produces negative ions which stream across to the collecting electrodes. The particles become electrically charged and are subjected to an electrostatic field. Their migration velocities cause them to deposit on the collecting electrodes. The deposits are then removed by rapping the collecting electrodes and fall by gravity into hoppers (Wu and others, 1997a,b). ESPs at coal-fired power plants can remove between 99–99.7% of the particulates from flue gas, depending on plant operating conditions (Breeze, 1998).

The capital cost for a new ESP is reportedly between 40–60 $/kWe. Higher efficiency ESPs generally have greater costs (Breeze, 1998). The US EPA data show the capital cost for a dry ESP would be 30,000–100,000 $/m^3.s of flue gas treated (1995US$). The operating cost would range from 8,500 to 85,000 $/m^3.s of flue gas treated. The corresponding annualised cost can vary between 10,000–85,000 $/m^3.s of flue gas treated (Hilton, 2000).

The World Bank (2001) estimated the costs of ESPs in China for a 300 MWe unit. It was shown that the capital costs for the 200 mg/m^3 emission level vary between 21–26 $/kWe, depending on the type of coal. If the emission level is 50 mg/m^3, the capital costs will increase to 29–33 $/kWe. The first-year operating cost for an ESP with 80 m^2/m^3/s SCA was estimated to be about 0.27 mills/kWh.

Much of the data relates to the costs for upgrading existing ESPs as the units are already fitted at most coal-fired power plants. Grieco (1997) evaluated the economics of various ESP upgrade technologies for a 500 MWe coal-fired unit (*see* Table 1). The accuracies in the costs are estimated to be within ±20% for most cases, whereas these are closer to ±10% for some less complicated technologies such as transformer/rectifier (T/R) set upgrading, rapper upgrading and sonic horns. Table 1 shows that the capital costs for the older technologies have decreased over the last decade. This is mainly due to a more competitive market and technological advancements. For example, the costs for T/R set and rapper controllers have decreased because of standardisation and the lower cost of electronic components. Magnetic

Table 1 Capital costs for ESP upgrading technologies for a 500 MWe unit (Grieco, 1997)

Technology	Capital cost, $/kWe		Impact on operating cost
	1984 survey	1997 survey	
Gas flow upgrade	0.75	0.70	decreased to unchanged
ESP controls upgrade	1.50	0.25	decreased to unchanged
T/R sets upgrade	3.50	1.75	unchanged to increased
Pulse energisation	10.00	4.50	unchanged to small decrease
Rappers upgrade	2.00	0.25	negligible
SO_3 conditioning	7.00	3.00	increased
Dual conditioning	–	4.00	increased
Granular feedstock conversion	–	0.60	decreased
Native conditioning	–	4.50	small increase
Sonic horns	–	0.75	negligible
Rebuild ESP internals	20.00	7.50	unchanged to increased
Rebuild internals, with 2 new fields	–	20.00	increased

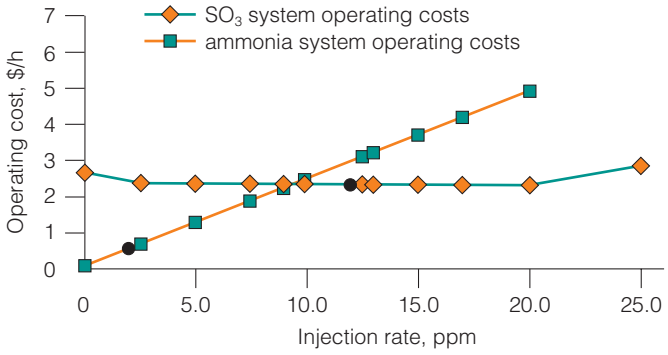
impulse/gravity impact rappers have also undergone dramatic design changes. The most new designs maximise the use of non-metallic materials. ESP rebuilds are less costly today due to greater market competition, the emergence of new construction techniques, and the use of wide plate spacing requiring less collecting plates.

Flue gas conditioning

Flue gas conditioning has proved to be more cost effective than adding new fields (Porle and others, 1997). In a study for a 600 MWe power plant, seven different coals were specified for ESP sizing to achieve an emission of 50 mg/m^3. The coals differ in the sulphur content from 0.3% to 1.1 %, silica plus alumina level from 71% to 93%, and sodium oxide level from 0.1% to 1.5%. Without conditioning, a five-field ESP with SCA in excess of 140 m^2/m^3/s would be required for the most difficult precipitated fly ash. With conditioning, the ESP size could be reduced to three-fields, with a SCA reduction close to 50%. This would result in a 20% reduction in the capital cost. In addition, further savings on the ash handling system and erection are possible. The reduction in ESP size with conditioning also lowers the operating cost. This is because fewer fields and hoppers are used, decreasing the number of heaters and consequently the power consumption.

Adding an anhydrous ammonia conditioning system to an existing SO_3 system would cost about 1 $/kWe for a 500 MWe unit (Grieco, 1997). The operating cost would increase by 50,000 $/y. The complete dual conditioning system would then cost around 4 $/kWe, with an operating cost of 150,000 $/y.

Hankins and Cramer (1996) analysed the costs for upgrading SO_3 conditioning to dual conditioning at Georgia Power Plant Yates Unit 5 (125 MWe). The upgrade included anhydrous ammonia storage and vaporisation equipment, ammonia metering device, an ammonia/air mixer, and ammonia injectors. The capital cost for equipment and installation was less than $100,000. This was low compared to the cost for a stand-alone ammonia conditioning system ($200,000 for equipment, $100,000 for installation). Figure 1 shows the operating cost for the dual conditioning system. The annual operating cost for the ammonia system alone was estimated



● optimised programme injection rates; NH_3 2 ppm; SO_3 12 ppm

Figure 1 Operating cost for the dual FGC system at Plant Yates Unit 5 (Hankins and Cramer, 1996)

to be about $6000. The ammonia delivered to the site costs around 520 $/t. At the normal injection rate of 2 ppm, the ammonia tank is filled every two to three months during peak load periods.

The feedstock conversion for an existing SO_3 conditioning system from molten sulphur to granular sulphur would cost about 0.6 $/kWe for a 500 MWe unit (Grieco, 1997). Granular sulphur is stable and solid. It has many handling and storage advantages but is expensive (155 $/t compared to 75 $/t for molten sulphur). The conversion eliminates the need for auxiliary steam. It is economically favorable when the system has a low utilisation factor, for example, a standby conditioning system or a generating unit on peak load.

A native SO_3 conditioning system for a 500 MWe unit would require a relatively high capital cost of 4.5 $/kWe (Grieco, 1997). This cost does not cover boiler modifications which may be needed to increase the thermal efficiency. However, its operating cost is expected to be lower than the conventional SO_3 conditioning system. The major operating cost is due to the loss in thermal efficiency resulting from a flue gas slipstream around the air heater, and catalyst rejuvenation and replacement.

Replacing internals

Wide plate spacing is one of the most economic and effective approaches to replacing internals (Sankey and Mastropietro, 1997). The cost benefits are driven from: the need for less internal elements; materials and erection savings due to reuse of part of the original casing; the weight savings effects on the existing support structures and foundations. The savings in internal elements are proportional to the ratio of the spacings. As an example, replacement of 23 cm by 41 cm spacing would reduce the number of collecting plates by 44% (1-23/41). The cost for replacement by 41 cm spacing and rigid discharge electrodes would be about 10 $/kWe, and 11 $/kWe for 30 cm spacing.

Enlarging ESP

The cost for enlarging ESP and replacing ESP internals for a 250 MWe unit was reported by Sankey and Mastropietro (1997). The enlargement included addition of fields and increase of collecting plate height. Replacement of internals involved replacement of an original design with wire discharge electrodes and 23 cm plate spacing with rigid discharge electrodes and 41 cm spacing. The total cost for increasing ESP height by 60% and length by 33% and replacement with 41 cm spacing was estimated to be 22.5 $/kWe. If the length is extended by 100% with the same other modifications, the cost will increase to 28.5 $/kWe.

Gaikwad and Boward (1999) estimated the costs for upgrading existing ESPs at a number of coal-fired units in the USA. It was shown that adding one ESP field would cost about 12 $/kWe for a 500 MWe unit. The increased operating cost was estimated to be 0.1 million dollars per year. The technology was expected to reduce the emissions from 0.05 lb/MBtu (62 mg/m^3) to 0.03 lb/MBtu (37 mg/m^3). The combination of adding one field with flue gas conditioning would lower the emissions from 0.07 lb/MBtu (86 mg/m^3)to 0.03 lb/MBtu (37 mg/m^3). This would increase the capital

Table 2 Comparative costs for the three types of bag filter (Hilton, 2000)

Technology	Capital cost	Operating cost	Annualised cost
Reverse gas	19,000–178,000	14,000–58,000	16,000–106,000
Pulse jet	13,100–54,900	11,200–51,700	13,100–83,400
Mechanical shaker	16,000–150,000	9,300–51,000	1,000–95,000

All costs presented as 1998 $/m³.s of flue gas treated

cost to 14–15 $/kWe and operating cost by 0.16–0.21 million dollars per year respectively, depending on unit size.

Sonic horns

Retrofit for sonic horns (acoustic cleaning of electrodes) for a 500 MWe unit may cost around 0.75 $/kWe if adequate compressed air is available (Grieco, 1997). A small dedicated air compressor would only add 0.03 $/kWe to the cost of installation. The operating cost is also relatively low.

2.2 Fabric filters

Fabric filters (bag houses) are tube-shaped filter bags through which the flue gas passes. The particles in the gas stream are captured by the fabric of the bags. The bag filters can be divided into three types according to the bag cleaning methods including reverse gas, pulse jet and mechanical shaker. The units are efficient, removing over 99% of the particulates from flue gas, even in the 0.05–1 µm range (Breeze, 1998).

Bag filters are reported to cost between 50 $/kWe and 70 $/kWe. They are generally more expensive than ESPs for collection efficiencies up to 99.5%. However, the units become more cost effective for higher collection efficiencies (Breeze, 1998). Table 2 summaries the comparative costs for the three types of bag filter according to the US EPA (Hilton, 2000). It shows that the reverse-gas bag filter represents the highest capital and operating costs, 19,000–178,000 and 14,000–58,000 $/m³.s of flue gas treated, respectively. The pulse-jet bag filter has the lowest capital cost and its operating cost is also relatively low.

The World Bank (2001) estimated the costs of bag filters in China for a 300 MWe unit. The capital costs were based on the assumption that bags (and cages for the pulse-jet bag filters) will have to be imported. The costs for three different designs are as follows:

- 57 $/kWe for the reverse-gas;
- 36 $/kWe for the pulse-jet;
- 26 $/kWe for COHPAC.

The first-year operating costs for the reverse-gas were estimated to be about 0.40 mills/kWh, and 0.48 mills/kWh for the pulse-jet and COHPAC.

Gaikwad and Boward (1999) presented a cost analysis for compact hybrid particulate collector (COHPAC) that uses a pulse-jet bag filter following an ESP. The technology was considered for upgrading ESPs at a few coal-fired units with 150–300 MWe. The capital costs were estimated according to Sargent & Lundy's database, vendor input on recent jobs, and previous studies. In estimating the operating costs, a bag life of three years was assumed. The COHPAC was shown to have a capital cost of 57–70 $/kWe and an operating cost of 0.32–0.57 million dollars per year, both depending on the unit size.

2.3 Comparative costs

Figure 2 compares the annualised costs for the ESP, reverse-gas bag filter and pulse-jet bag filter over 15 years (Gyllenspetz and others, 1998). The costs reflected US practices. The comparison was based on a 250 MWe unit firing different coals for an emission of 50 mg/m³ at 6% oxygen. The following assumptions were made: ESP pressure drop of 40 mm wg (392 Pa); bag filter pressure drop of 190 mm wg (1862 Pa); electricity cost at 5 ¢/kWh; bag filter life of four years. The comparative annualised costs tend to favour the ESP.

It is also shown that the costs for the bag filters are constant over a wide range of fly ash resistivity. However, the cost for ESP is found to increase steadily with fly ash resistivity. This is because the bag filter performance is less likely to be affected by the resistivity and other fly ash and flue gas characteristics (such as inlet dust loading and gas temperature). The ESP performance, on the other hand, depends greatly on the fly ash resistivity. High resistivity fly ashes can cause back-corona on the collecting plates, lowering the ESP efficiency. The ESP will then need upgrading in order to achieve the same emission level. This will therefore result in higher capital and operating costs. The graph indicates a possible advantage of using the bag filter for high fly ash resistivity over 10^{13}–10^{14} ohm.cm.

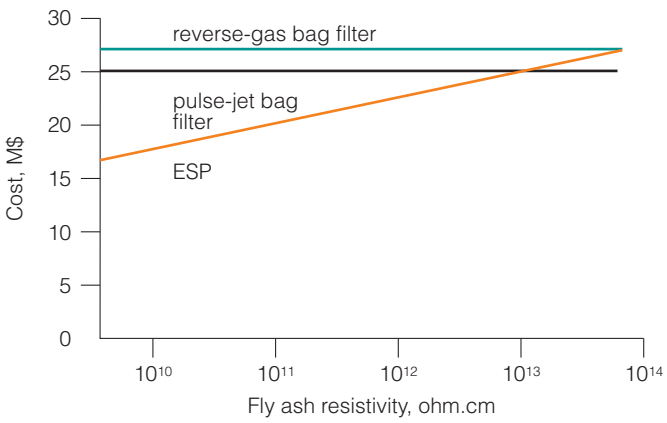


Figure 2 Annualised costs for ESP, reverse-gas bag filter and pulse-jet bag filter over 15 years (Gyllenspetz and others, 1998)

Sloat and Gaikwad (1997) compared the economics of various ESP retrofit technologies, including adding field, SO_3 conditioning, COHPAC, replacement with a pulse-jet bag filter, and adding a wet ESP. The comparison considered a moderate ESP with a specific collection area (SCA) of 275 ft²/1000 acfm (54 m²/m³/s) (Case 1) and a small ESP with an SCA of 150 ft²/1000 acfm (30 m²/m³/s) (Case 2) meeting emission limits of 0.1–0.01 lb/MBtu (123–12.3 mg/m³). In both cases, the capital costs were estimated according to Sargent & Lundy's database, vendor input on recent jobs, and previous studies. The levelised costs were based on a 30-year plant life. Both of the costs were based on December 1997 US dollars.

Figure 3 shows the comparative capital and levelised costs of various options for Case 1. It should be noted that the bag filter and wet ESP options included a 10% and 20% process contingency respectively, to cover their capability at the 0.01 lb/MBtu (12.3 mg/m³) level. It is shown that ESP upgrades (adding field, SO_3 conditioning) will have lower capital costs than the bag filter and wet ESP options for medium/high sulphur Illinois coal with emission levels of 0.1–0.01 lb/MBtu (123–12.3 mg/m³), and for low sulphur PRB coal with emission levels of 0.1–0.03 lb/MBtu (123–37 mg/m³). However, the levelised costs indicate that only the COHPAC option will be an alternative for low sulphur PRB coal to meet the 0.01 lb/MBtu (12.3 mg/m³)

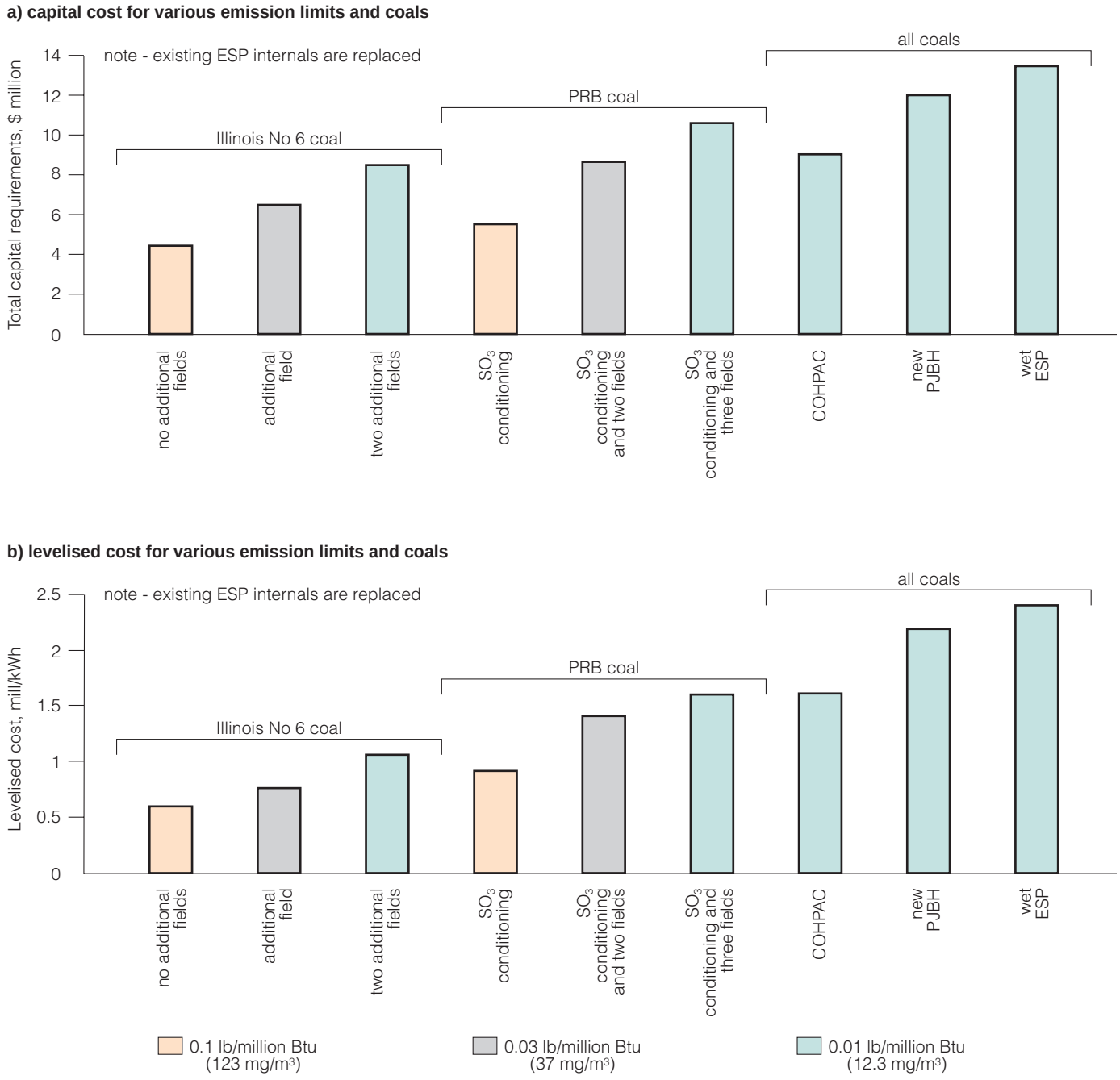


Figure 3 Comparative capital and levelised costs of various options for Case 1 (Sloat and Gaikwad, 1997)

level. This is mainly because the bag filter has a high operating cost due to bag replacement and auxiliary power. While the COHPAC is much smaller than a conventional pulse-jet bag filter, the lower operating cost makes it comparable with the ESP upgrades.

Figure 4 shows the comparative capital and levelised costs of various options for Case 2. It indicates that ESP upgrades (adding field, SO_3 conditioning) have lower capital costs than the bag filter and wet ESP options for medium/high sulphur Illinois coal with emission levels of 0.1–0.03 lb/MBtu (123–37 mg/m³), and for low sulphur PRB coal with an emission level of 0.1 lb/MBtu (123 mg/m³). However, the levelised costs show that only the COHPAC option will be an alternative for both coals to meet the 0.01 lb/MBtu (12.3 mg/m³) emission level. It should be noted that the cost of the outage required for retrofit was not included. If this cost is taken into account, the COHPAC and pulse-jet bag filter options will be more cost effective than the ESP upgrades to meet the 0.01 lb/MBtu (12.3 mg/m³) emission level.

2.4 Summary

The costs for particulate control systems are site-specific and

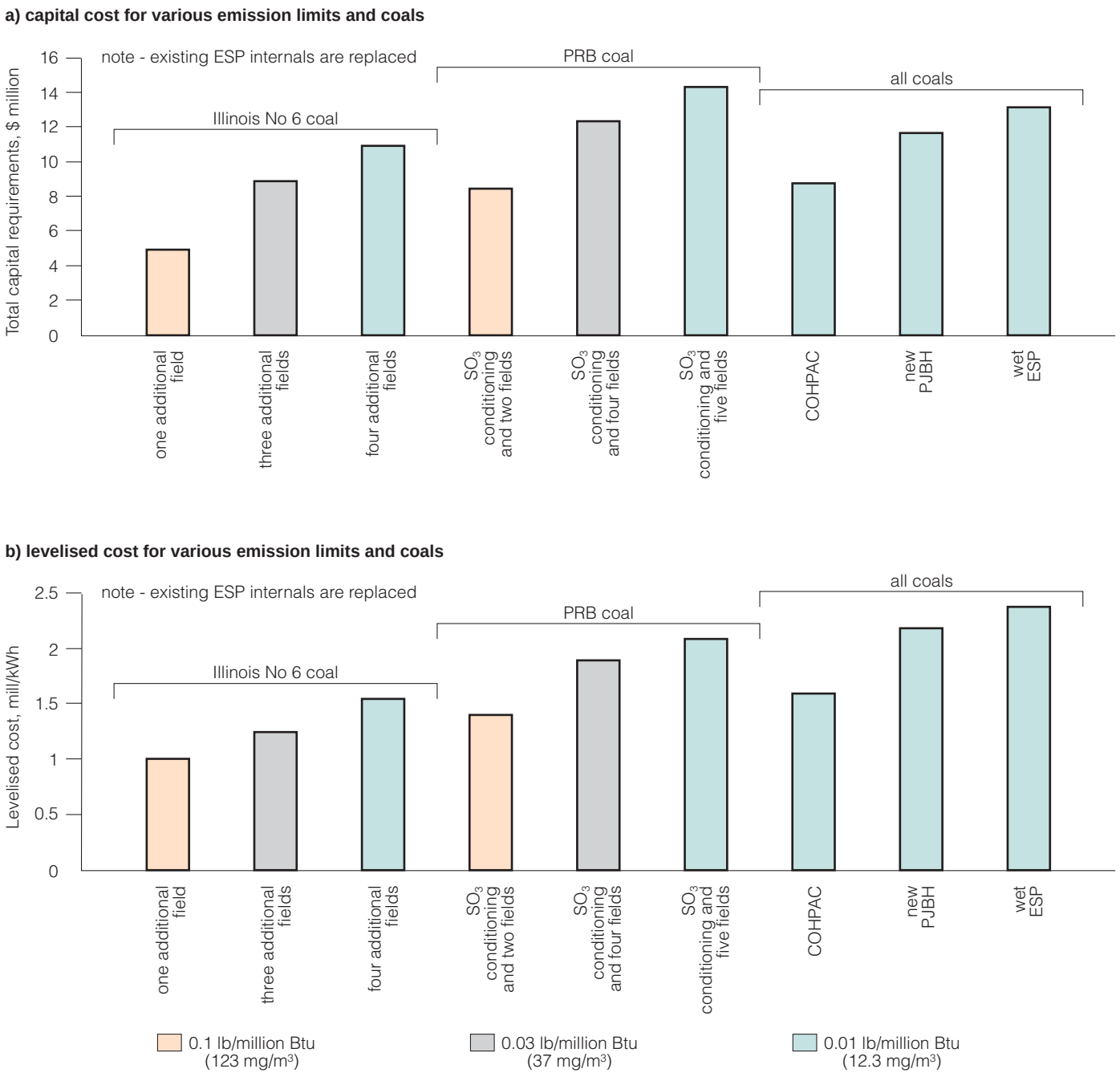


Figure 4 Comparative capital and levelised costs of various options for Case 2 (Sloat and Gaikwad, 1997)

vary from country to country. The capital cost for a new ESP is reportedly between 40–60 $/kWe. The operating cost can vary widely from 8,500–$85,000 $/m3.s of flue gas treated, depending on the plant operating conditions. The annualised cost can range from 10,000 to 85,000 $/m3.s of flue gas treated.

ESP upgrades represent considerably lower capital costs than a new ESP. The capital costs for upgrades have decreased for the last decade, due to a more competitive market and technological advancements. ESP upgrades can cost from 0.25 $/kWe (for rapper upgrade) to 12 $/kWe (for adding one field), depending on the technology used.

Bag filters are reported to cost between 50 $/kWe and 70 $/kWe. The operating costs are also relatively high due to bag replacement and auxiliary power. They are generally more expensive than ESPs for collection efficiencies up to 99.5%. However, the units become more cost effective for higher collection efficiencies.

The choice of technology depends on the coal type and emission level. For medium to high sulphur coals with low fly ash resistivity, ESP upgrades may provide the lowest cost to meet the emission levels of 0.1–0.01 lb/MBtu (123–12.3 mg/m3). For low sulphur coals with high fly ash resistivity, COHPAC and bag filter options offer alternatives to ESP upgrades for the emission level of 0.01 lb/MBtu (12.3 mg/m3). ESP upgrades with SO_3 conditioning provide significant cost advantage for low sulphur coals.

3 SO_2 control

FGD is an effective control of SO_2 emissions from coal-fired power plants. In the last few decades, FGD processes have undergone considerable developments in terms of improved removal efficiency and reliability, as well as reduced costs. Table 3 summaries major FGD processes currently in use and their features. More details are discussed in a recent report of IEA Coal Research – The Clean Coal Centre (Soud, 2000).

There are currently 680 FGD installations at coal-fired power plants worldwide and 140 are under construction or planned. Wet scrubbers are the most commonly used FGD system, accounting for 87% of the total FGD capacity worldwide. This is followed by spray dry scrubbers with 8%, sorbent injection units with 2% and others (circulating fluid bed, regenerable, combined SO_2/NOx) with 3%. The share for wet scrubbers is expected to increase to 95% (Soud, 2000; IEA Coal Research, 2000). Wet FGD processes commonly use lime or limestone as the main sorbent, while other sorbents such as seawater and ammonia have also been employed. Wet limestone is the predominant process, accounting for 82% of all installed wet FGD capacity worldwide (Maller and Hollinden, 2000; Jozewicz and others, 1999).

Factors affecting FGD cost are discussed in Section 3.1. Wet processes are detailed in Section 3.2 and other FGD processes in Sections 3.3–3.6. Comparative costs are then reviewed in Section 3.7.

3.1 Factors affecting cost

The costs of FGD plant include capital and operating costs both of which are site-specific. These are detailed below.

Capital costs

The capital costs of FGD plant are considerably influenced by market conditions and other factors, for example, geographical location and the amount of preparatory site work required. In addition, the costs of FGD plant also depend on technical factors such as (DTI, 2000a)

- volume of flue gas to be treated;
- concentration of SO_2 in the flue gas;
- desulphurisation efficiency required;
- quality of the by-products produced;
- other environmental constraints, for example, permitted waste water discharges;
- the need or otherwise for flue gas reheat;
- the degree of reliability and redundancy required;
- design life.

The capital costs of wet FGD plant have been falling for the 30 years since the first commercial application on utility boilers. This is due to advances in technology, reduction or elimination of equipment redundancy, and a competitive market. For example, many early FGD plants had multiple and/or spare absorber towers whereas this is now seldom the case. The downward trend in FGD capital costs in the USA is shown schematically in Figure 5. It shows an overall reduction of as much as 75% between the 1970s and 2000. The average cost for retrofit FGD in the 1970s was approximately 400 $/kWe. This has now dropped to about 100–125 $/kWe, with further reductions in cost anticipated beyond 2000 (DTI, 2000; Boward and Brinkmann, 1998). Barrett (2000) reported a similar figure for the current wet FGD cost which is in the range 80–120 $/kWe depending on plant size, fuel type and other factors. However, a study by Breeze (1998) shows the cost for a retrofit is 150–270 $/kWe and for a new plant the cost is 120–210 $/kWe.

Operating costs

Operating costs depend on the costs of sorbents, the costs associated with disposal or utilisation of the by-products and power costs, all of which are influenced by location conditions. The costs can be split into variable and fixed costs (DTI, 2000a).

Variable costs cover costs of operation such as reagents, by-product disposal and utilities (steam, power, water). Reagent costs are approximately proportional to the amount of SO_2 removed, although they are also influenced by the chemical stoichiometry, which may vary with the desulphurisation efficiency. Similarly, by-product disposal costs are also proportional to the amount of by-product produced (negative if the by-product is saleable). Costs of steam, power, water depend on both the amount of flue gas treated and the throughput of reagents and by-products. Reagent and by-product disposal costs are considerably influenced by the location of the FGD plant. The total cost of a reagent to the plant includes a transport cost, and therefore the proximity of a limestone mine, for example, can be an important consideration. Similarly, costs of FGD plant can be reduced if there is a local disposal point or user for the by-product.

Fixed costs include costs of operating labour, maintenance (both materials and labour) and administration. Maintenance costs are usually estimated as a percentage of the total plant cost. They are generally taken to be independent of plant operating hours or operating regime. In reality, however, both these affect maintenance costs. The number of start-ups, in particular, can considerably affect the rate of degradation of plant components, due to the mechanical and thermal stresses that the start-up procedure imposes on the plant. Examples include the impacts of thermal cycling on the linings in FGD absorbers and the additional rotational loads on motors and pumps as they are accelerated to operating speed.

3.2 Wet processes

Figure 6 compares the capital and operating costs for various wet processes. It shows that the limestone process has the highest capital costs, twice as high as those for the magnesium and sodium processes. However, it has the lowest

Table 3 FGD processes (Wu, 2000b)

FGD process	Main features/comments
Wet lime/limestone process	90–98% SO_2 removal efficiency; over 99% of operation reliability with forced oxidation, 95–9% with natural oxidation
Other wet processes	expensive reagent, limited market for by-products, used in early installations and in relatively small plants
Dual alkali	seldom used in new plants
Sodium carbonate	lower capital costs, less volume of reagent required, 74–95% SO_2 removal efficiency, used for <1% sulphur coals
Seawater	use of natural alkalinity, low capital and operating costs
Magnesium	low capital costs, suitable for plants located near the coast
Ammonia	high SO_2 removal efficiency, marketable by-product, no waste water, no other waste by-product requiring disposal
Passamaquoddy recovery	92% SO_2 removal efficiency, marketable by-product
Spray dry processes	97–100% of operation reliability, lower capital costs but higher and more expensive sorbent use compared with wet processes, used mostly for small to medium plants firing low to medium sulphur coals, preferable for retrofits
Sorbent injection processes	simple process, moderate SO_2 removal efficiency, lower capital costs
Furnace sorbent injection	50% SO_2 removal efficiency with limestone injection, poor sorbent utilisation
Duct sorbent injection	50% SO_2 removal efficiency, low power consumption, easy to install
Hybrid sorbent injection	over 90% SO_2 removal efficiency
Circulating fluid bed	up to 99% SO_2 removal efficiency, easy to operate and maintain
Moving bed	90% SO_2 removal efficiency
Regenerable processes	sorbent is regenerated and reused, saleable by-products (elemental sulphur or sulphuric acid), higher capital costs
Wellman-Lord	95–98% SO_2 removal efficiency
Magnesium oxide	95–98% SO_2 removal efficiency
Amine based sorbent	90–99% SO_2 removal efficiency
Combined SO_2/NOx removal processes	complex and costly
Solid adsorption/regeneration	95–98% SO_2 and 80-90% NOx removal efficiency
Gas/solid catalytic operation	over 90% SO_2 and NOx removal efficiency
Electron beam irradiation	over 95% SO_2 and 80% NOx removal efficiency
Alkali injection	80–98% SO_2 and 60–80% NOx removal efficiency, under bench/pilot/demonstration scale
Wet scrubbing	90% SO_2 and 50% NOx removal efficiency with the Fe(II) EDTA process, 90% SO_2 and 30–80% NOx removal efficiency with the ClO_2 process

operating costs mainly because limestone is cheaper than other reagents.

Wet limestone/lime process

Table 4 shows the capital costs for some recent wet limestone forced oxidation (LSFO) retrofits ranging in size from 316 to 1700 MWe. The costs include published ones and those predicted from the Coal Utility Environmental Cost (CUECost) model which was recently developed by the National Risk Management Research Laboratory of the US EPA. The published costs vary from 180 to 348 $/kWe. The model predictions range from 179 to 362 $/kWe, within +/- 10% of the published ones.

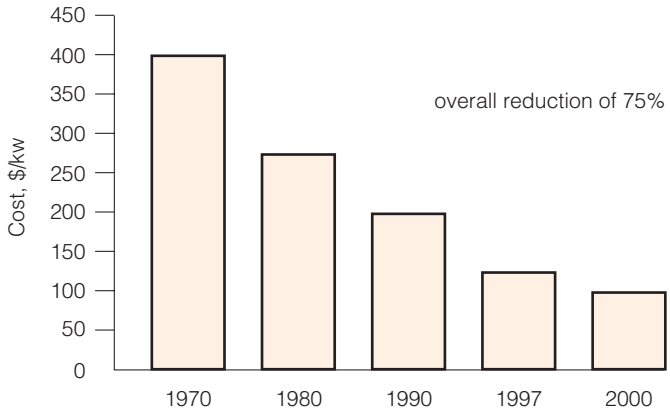


Figure 5 Reductions in the capital costs of retrofit wet FGD plants in the USA (Boward and Brinkmann, 1998)

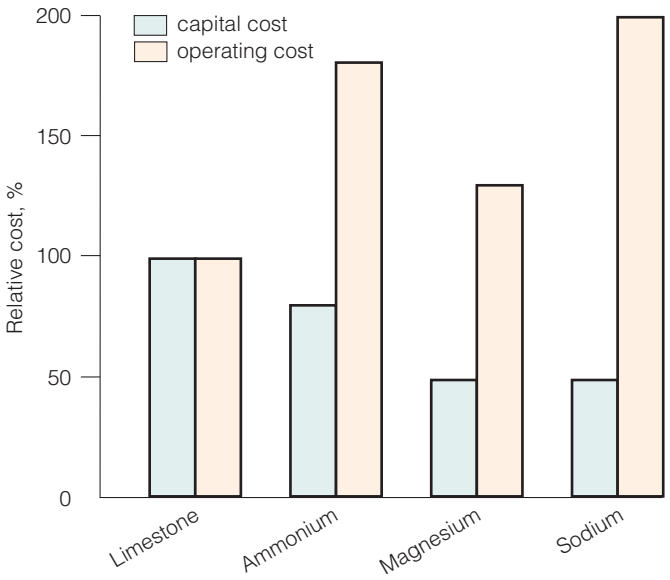


Figure 6 Cost comparison for various wet FGD processes (Nara, 2000)

The operating costs for wet limestone/lime processes in the USA were recently surveyed (Blythe and others, 1999; Weilert and Dyer, 1999). Table 5 summarises the results of the survey for different coal sulphur levels which are discussed below.

High-sulphur coal – The 43 high-sulphur-coal FGD systems included LSFO, limestone natural oxidation (LSNO), and magnesium-lime systems. Table 5 shows that on average the LSFO systems had the lowest operating costs, but a higher auxiliary power consumption than the other two system types. The higher power consumption for the LSFO systems would add only $0.10–0.20 to the costs for this system type relative to the costs for the other two, assuming a typical utility internal power consumption cost of 20 $/MWh. The overall operating costs for the LSFO systems are about 50–60% of the costs for limestone natural oxidation, and about 40% of those for the magnesium-lime systems.

Factors affecting the operating costs should be considered. The survey showed the lower total operating costs for newer FGD systems. The LSFO systems are on average the newest, and this has contributed to the lower operating costs of the systems. The system types examined fire different sulphur content coals but there was no clear trend for increasing operating costs with the coal sulphur. The survey also showed lower total operating costs ($/MWh) for larger units over the range of 200 to 800 MWe. The magnesium-lime systems are on average installed on larger units than the limestone systems. This would give even higher operating costs for the magnesium-lime systems for the same unit size.

Medium-sulphur coal – In the survey there were only five LSFO systems and seven LSNO systems. The average operating costs for the two system types are shown in Table 5. The relatively small number of systems makes the comparison difficult. The costs for the LSFO systems are shown to be higher than those for LSNO systems although the LSFO systems are relatively new. This is affected by one small LSFO system that has high labour and waste disposal costs (as $/MWh). The high labour costs are due to the small plant size and the high waste disposal costs are because the plant is relatively far from a wallboard plant.

Table 4 Capital costs for some recent LSFO retrofits (FGD and DeNOx, 2000a)

Power plant	LSFO system capacity, MWe	Published, $/kWe	CUECost predicted, $/kWe	Variation, %
Petersburg	657	317	291	-8.20
Cumberland	1300	200	187	-6.50
Conemaugh	1700	195	179	-8.21
Ghent	511	215	229	+6.51
Bailly	668	180	196	+8.89
Milliken	316	348	362	+4.02
Navajo	750	236	213	-9.75

Table 5 Operating costs for wet lime/limestone FGD systems with different sulphur-content coals (Blythe and others, 1999)

High-sulphur-coal Cost category	Limestone forced oxidation, $/MWh*	Limestone natural oxidation, $/MWh*	Magnesium-lime $/MWh*
Feed materials and chemicals	0.246	0.518	1.435
Labour and supervision	0.341	0.431	0.353
Waste disposal	0.002	0.275	0.461
Maintenance, materials and other	0.239	0.606	0.411
Total	0.828	1.830	2.473
Average unit size, MWe	612	500	747
Average capacity factor, %	63.3	65.9	65.2
Average sulphur, lb SO_2/MBtu (mg/m³)	5.4 (6642)	5.6 (6888)	6.0 (7380)
Average in-service date	Dec 1989	Mar 1983	May 1985
Auxiliary power, % of net generation	2.5	1.7	2.0
Medium-sulphur-coal Cost category	Limestone forced oxidation, $/MWh*	Limestone natural oxidation, $/MWh*	
Feed materials and chemicals	0.481	0.496	
Labour and supervision	0.719	0.089	
Waste disposal	0.186	0.213	
Maintenance, materials and other	0.349	0.178	
Total	1.436	0.859	
Average unit size, MWe	480	570	
Average capacity factor, %	68.1	72.7	
Average sulphur, lb SO_2/MBtu (mg/m³)	3.6 (4428)	3.4 (4182)	
Average in-service date	Mar 1995	Oct 1983	
Auxiliary power, % of net generation	1.5	1.0	
Low-sulphur-coal Cost category	Limestone forced oxidation, $/MWh*	Limestone natural oxidation, $/MWh*	Wet lime $/MWh*
Feed materials and chemicals	0.086	0.098	0.333
Labour and supervision	0.256	0.214	0.356
Waste disposal	0.061	0.049	0.023
Maintenance, materials and other	0.331	0.292	0.207
Total	0.695	0.610	0.919
Average unit size, MWe	475	505	406
Average capacity factor, %	66.5	67.2	67.0
Average sulphur, lb SO_2/MBtu (mg/m³)	1.7 (2091)	1.1 (1353)	1.2 (1476)
Average in-service date	Mar 1988	May 1982	May 1978
Auxiliary power, % of net generation	1.1	1.2	1.7

* $/MWh of annual net generation by scrubbed units

Low-sulphur coal – In the survey there were five LSFO systems, 25 LSNO systems and 12 wet lime reagent systems. Table 5 shows that the wet lime system has the highest total operating costs and the LSNO system has the lowest total costs. A large percentage of the difference is due to the effect of reagent cost. It is noted that the LSFO systems are on average the newest, but treat the highest sulphur coals. The wet lime systems are on average the oldest, smallest, and consume the highest percentage of the net generation.

Comparing data in Table 5, it can be seen that higher sulphur coals generally result in higher operating costs. This trend would in particular be driven by reagent and by-product disposal costs. It is noted, however, that the operating costs for the high-sulphur LSFO systems are only 20% higher than the low-sulphur system costs. This is apparently because the by-products from the high-sulphur systems are sold while the low-sulphur systems produce smaller amounts of by-products which make the sale less worthwhile. In addition, the high-sulphur systems are on average newer and larger.

The **Chiyoda CT-121 wet limestone process** was reported to have a lower operating cost than most other wet limestone processes (Ogi and others, 1999). The process allows the use

of coarser limestone particles with 90% passing through a 200 mesh screen, compared to 95% passing through a 325 mesh screen for most other processes. This reduces the operating cost of the limestone grinding system by about 20%, lowering the overall FGD operating cost. The Chiyoda process uses an advanced control system to maintain a constant SO_2 concentration at the FGD outlet. This minimises the power and limestone consumption, reducing the variable operating cost. In addition, the process permits an increased particulate loading at the FGD inlet. The ESP system can therefore be designed for a higher outlet particulate loading, reducing the ESP operating cost possibly by 20–30%.

An economic analysis of the wet limestone and **wet Thiosorbic magnesium-enhanced lime** (developed by Dravo Lime Company, USA) processes was conducted by Smith and others (1998). The Thiosorbic process has a higher SO_2 removal efficiency. This allows the firing of lower cost and higher sulphur coals, reducing the fuel cost while maintaining SO_2 emission compliance. The fuel and FGD variable operating costs were compared for the two processes. This shows that the Thiosorbic process has a higher variable operating cost which results from the higher reagent cost. However, the lower fuel cost with the process reduces the overall plant operating cost by 1.4 million US dollars per year.

Ammonia process

In contrast to Figure 6, EPRI's survey shows the ammonia process has higher capital costs but lower operating costs than the limestone process (DTI, 2000a). It is claimed, however, that both the capital and operating costs of the Clean & Green ammonia process are less than the costs of the LSFO process (Benetech Inc, 1999). The ammonia process benefits from the sale of ammonium sulfate fertiliser. The cost of the process was estimated for a hypothetical installation at a 300 MWe new unit firing 2.6% sulphur bituminous coal. The operating cost is shown to be approximately 322 \$/t SO_2 removal (1990 value), only 75% of the costs for the LSFO process.

Borio and others (1999) presented a case study comparing the capital and operating costs of the ammonia and limestone processes for a hypothetical installation. The case study considers a plant with 2x500 MWe coal-fired units operating at 85% capacity factor. The plant is located in the mid-western USA with reasonable access to major agricultural areas. The SO_2 concentration of the flue gas is assumed to be 2.44 kg/GJ (6588 mg/m^3). Both processes have a removal efficiency of 95%.

The capital costs are estimated on a lump sum, turnkey basis excluding interest during construction and the owner's internal costs. The main cost items include:

- firm, lump sum and turnkey;
- moderately difficult retrofit;
- all process equipment (absorber vessel, reagent preparation, by-product processing);
- reagent and by-product storage facilities;
- ductwork, dampers, new stack;
- control system;
- civil (piles, foundations, buildings);
- electrical (such as switchgear, cable, lighting, fire detection, motors).

The estimated capital cost for the ammonia process is 140 million US dollars, higher than the cost for the limestone process which is 120 million US dollars.

The operating costs are estimated including the costs for financing, operating and maintenance labour, power, pressure drop, natural gas (only for ammonia process), reagent, and

Table 6 Comparison of the seawater and wet limestone processes (Ziemann-Nöthe and Küper, 1997)

	Wet limestone	Seawater
SO_2 removal efficiency	up to 95% (sulphur content in coal >1%)	up to 98% (limited to sulphur content in coal <1%)
By-product	170,000 tonnes of gypsum per year	sulphate ions (dissolved in the seawater)
Absorbent	100,000 tonnes of limestone per year	40,000 m^3/y of seawater
Power consumption	0.5% of total electric capacity	0.5% of total electric capacity
Process water consumption	700,000 m^3/y	1000 m^3/y
Waste water quality	complies with common standards slight increase of suspended solids	complies with common standards slight increase of sulfate ions and heavy metals
Space requirement/civil works	high absorber + auxiliary systems	high absorber + seawater supply
Capital cost	100%	80–85%
Operating cost	medium	low

by-product. The results show that the ammonia process, with the sale of by-product, has a substantial economic advantage over the limestone process.

Seawater process

The seawater process has economic advantages over the wet limestone process. Dighe and others (1992) reported that the Alstom seawater process has a relatively low operating cost. This is because there are no costs for sorbent and by-product disposal. In addition, the process only requires a negligible amount of process water.

Ziemann-Nöthe and Küper (1997) compared the seawater process with the wet limestone process for a 2x660 MWe plant firing 1% sulphur coal. The comparison is given in Table 6. It shows that the capital cost of the seawater process is 80–85% that of the wet limestone process. The operating cost of the seawater process is also relatively low.

3.3 Spray dry processes

The spray dry process typically uses lime or calcium oxide as sorbent. It generally has a lower capital cost but higher and more expensive sorbent use compared with wet processes. The process is used mostly for small to medium plants firing low to medium sulphur coals, and is preferable for retrofits (Wu, 2000b).

In a recent survey in the USA, the operating costs for the lime spray dry process were reported (Blythe and others, 1999). Table 7 summarises the results together with those for the wet LSFO process. It shows that the spray dry process has a higher operating cost than the wet process due to the higher cost of sorbent.

3.4 Sorbent injection processes

The sorbent injection process, with a moderate SO_2 removal efficiency, has a relatively low capital cost. The capital cost for furnace sorbent injection is around 70–120 $/kWe (Breeze, 1998).

Korhonen and Ekman (1996) reported on the economics of the LIFAC (limestone injection into furnace and activation of unreacted calcium) process. The process consists of three steps. The first step (limestone furnace injection) provides 25–35% SO_2 removal and represents about 10% of the total capital cost. The second step (flue gas humidification with dry ash recycle) brings the SO_2 removal up to 75%. It accounts for 85% of the total capital cost. The addition of slurry ash recycle increases the SO_2 removal to 85%, representing about 5% of the total cost. The LIFAC process is considered simple and easy to operate. It requires little maintenance compared with the wet limestone process.

SaskPower (Canada) estimated the lifetime costs of the wet limestone and LIFAC processes. A comparison was made for the 300 MWe Shand Power Plant burning local low sulphur and low heating value lignite (*see* Table 8). The cost of pulverised limestone is relatively high (about 50 US$/t) due to a long transport distance to the power plant. It is shown that the capital cost of the LIFAC is only one fifth that of the wet process. The total lifetime cost for the LIFAC is 52% lower in comparison with the wet limestone process while the operating cost is substantially higher.

3.5 Circulating fluid bed processes

The circulating fluid bed (CFB) process is considered easy to operate and maintain. EPRI's survey (DTI, 2000a) shows that the CFB has a relatively low capital cost, similar to that for the spray dry process. The process also has a low/moderate fixed operating cost but its variable operating cost is relatively high. Toher (1994) compared the costs of Lurgi CFB, spray dry and wet limestone processes (*see* Table 9). The Lurgi CFB and spray dry processes are shown to have lower fixed operating costs and capital charges. The variable operating costs for the two processes are higher due to the expensive sorbent but lower sorbent utilisation compared with the wet limestone process. The larger plant size with higher sorbent utilisation favours the wet limestone process. However, for a smaller plant size, the fixed operating cost and capital charge account for a higher proportion of the total levelised cost. The CFB and spray dry processes are better suited for small plants over any sulphur content coal and also for larger plants with lower sulphur coal (Toher, 1994; Takeshita, 1995).

The gas suspension absorption (GSA) system was developed by F L Smidth miljø a/s (Denmark). The system has a more than 90% of SO_2 removal efficiency, while achieving a relatively high utilisation of reagent lime. The economics of the GSA were evaluated for a moderately difficult retrofit to a

Table 7 Operating costs for wet and spray dry processes with low-sulphur coals (Blythe and others, 1999)

Cost category	Limestone forced oxidation, $/MWh*	Lime spray dry process, $/MWh*
Feed materials and chemicals	0.086	0.334
Labour and supervision	0.256	0.212
Waste disposal	0.061	0.056
Maintenance, materials and other	0.331	0.247
Total	0.695	0.836

* $/MWh of annual net generation by scrubbed units

Table 8 SaskPower's lifetime cost estimates for wet limestone and LIFAC processes (Korhonen and Ekman, 1996)

Costs	Wet limestone	LIFAC	LIFAC-slurry recycle
Capital cost (million Canadian dollar)			
Reactor & auxiliaries	75	10.90	10.90
Sorbent injection	inclusive	0.90	0.90
Slurry equipment	inclusive	0	1.50
Delay ESP duct erection	inclusive	0.50	0.50
Electrical installation	inclusive	0.60	0.60
Miscellaneous mechanical installation	inclusive	0.10	0.10
Air compressor	inclusive	0.30	0.30
Foundations	inclusive	0.40	0.40
Controls	inclusive	0.20	0.20
Ash system	inclusive	0.30	0.30
Total	75	14.20	15.70
Operating cost (million Canadian dollar)			
Sorbent	9.10	43	26.50
Auxiliaryelectrical power	3.40	0.90	1
Labour & maintenance	10.40	1.20	1.90
Fuel for calcination	0	2.30	1.40
Ash disposal	base	0.50	0.30
Total	23	47.90	31.10
Total lifetime cost	98	62.10	46.80

Table 9 Comparative economics of Lurgi CFB, spray dry and wet limestone processes (1994 $/t of SO_2 removed) (Toher, 1994)

Plant size, MWe	Coal sulphur, %	FGD process	Fixed operating cost	Variable operating cost	Capital charge	Total levelised cost
200	2.6	Wet limestone	79	90	183	352
200	2.6	Spray dry	40	143	134	317
200	2.6	Lurgi CFB	15	140	122	277
500	4.3	Wet limestone	20	76	83	178
500	4.3	Spray dry	11	131	69	211
500	4.3	Lurgi CFB	3	130	63	196

30 years levelised cost, 90% of SO_2 removal efficiency

300 MWe boiler burning a 2.6% sulphur coal (Hsu and others, 1997). The design SO_2 removal efficiency was 90%. Table 10 compares the capital and levelised costs for the GSA and wet limestone forced oxidation systems. The capital cost for the GSA was estimated to be 149 $/kWe, which is substantially lower than the 216 $//kWe for the wet limestone system. The significant reduction in capital is primarily due to the lower cost in SO_2 absorption area. The GSA also has a lower levelised cost than the wet limestone system. The principal operating cost for the GSA system is the cost of pebble lime.

Another process, offered by Alstom Power, is called Flash Dry Absorber Dry FGD (Windsor, 2001; Rader and others, 2000). The process is based on the absorption of SO_2 by a dry absorbent containing lime or dry hydrated lime. It has a high recycle rate, which maximises the utilisation of the absorbent. A 90–95% of SO_2 removal efficiency can be achieved. The performance is reported to be equal or better than conventional dry FGD systems at a greatly reduced cost.

3.6 Regenerable processes

In a regenerable process, sorbent is regenerated and reused. A few regenerable processes such as the Wellman-Lord and MgO processes have been in commercial use since the 1970s. There are some new processes being developed or demonstrated (Soud, 2000; Larsen and others, 1999; Klaasen and others, 1998).

Regenerable processes generally have a high capital cost and power consumption. The net variable operating cost is moderate because the processes produce saleable by-products

(elemental sulphur or sulphuric acid). However, the fixed operating cost is substantially high (DTI, 2000a). Table 11 shows the comparative costs of some regenerable processes. The estimates were based on a 300 MWe plant firing 2.6% sulphur coal with a moderate retrofit factor of 1.3.

Table 10 Comparative economics of the GSA and wet limestone forced oxidation processes (Hsu and others, 1997)

Cost	GSA	Wet limestone
Capital cost (1990$/kWe)		
Reagent feed	25	37
SO_2 removal	38	71
Flue gas dandling	18	24
Solids handing	5	7
General support	1	2
Additional equipment	4	4
Process capital	91	145
Total capital cost	149	216
Levelised cost (mills/kWh)		
Fixed cost		
Operating labour	0.52	0.66
Maintenance	1.49	1.74
Administrative & support labour	0.34	0.41
Subtotal	2.35	2.81
Variable cost		
Raw material	1.82	0.65
Solids disposal	0.86	0.57
Water	0.01	–
Steam	–	0.55
Electricity	0.47	1.16
Subtotal	3.16	2.93
Capital charge	5.40	7.30
Total levelised cost	10.91	13.04

The membrane contactor electro dialysis (MCED) process is a new technology developed by FLS miljø a/s (Denmark) (Larsen and others, 1999). In the process, SO_2 is absorbed in a membrane contactor and absorbent is regenerated by electro dialysis with bipolar membranes. The process produces saleable SO_2 or SO_2 derivatives as sulphur by-product. The economics of the process were evaluated. In the evaluation, the scope of supply for the MCED plant included the inlet and outlet ducts, the bypass duct and the FGD unit. It excluded civil works, fans, power supply, process control system, and by-product handling system. Table 12 summarises the costs of the process in comparison with the conventional wet limestone process. The comparison was based on a plant lifetime of 20 years and a 6500 hours per year load. The capital cost for the wet process was reportedly about 19 millions US dollars. The corresponding operating cost was found to be 27 millions US dollars. This gave an overall cost of 0.94 mills/kWh. In comparison with this, the capital cost for the MCED process was estimated to be 17.8 million US dollars and the operating cost 13.2 million US dollars. The overall cost was therefore found to be 0.64 mills/kWh, approximately two thirds of the cost of the wet process. The MCED process is claimed to become even more competitive with an expected reduction in the electro dialysis membrane price.

3.7 Comparative costs

The capital costs of different FGD processes vary according to the degree of complexity of the process, the amount of engineering required and other factors. EPRI's study shows that processes in which the sorbent is processed to produce a saleable by-product (such as limestone and ammonia processes) or the sorbent is regenerated (for example,

Table 11 Comparative costs of some regenerable processes (1990$) (Takeshita, 1995)

Process	Capital cost, $/kWe	Levelised cost (no inflation), mills/kWh	Levelised cost (no inflation), $/t of SO_2 removed	Levelised cost (5% inflation), mills/kWh	Levelised cost (5% inflation), $/t of SO_2 removed
SOXAL	241	8.5	551	13.6	877
Wellman-Lord	278	8.4	520	13.3	823
MgO	283	8.7	533	13.7	844

Table12 Comparative costs of the MCED and wet limestone processes (Larsen and others, 1999)

Cost	Wet limestone	MCED
Capital cost, $(000)	19,111	17,805
Levelised capital cost, $/kWe	51	47
Annual operating cost, k$/y	1965	967
Total operating cost, $(000)	26,916	13,246
Overall cost, $(000)	46,027	31,051
Levelised overall cost, mills/kWh	0.94	0.64

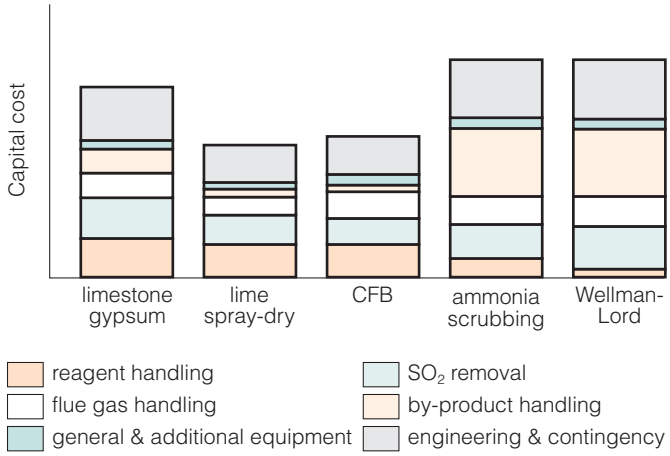


Figure 7 Relative capital costs of FGD processes (reproduced from DTI, 2000a)

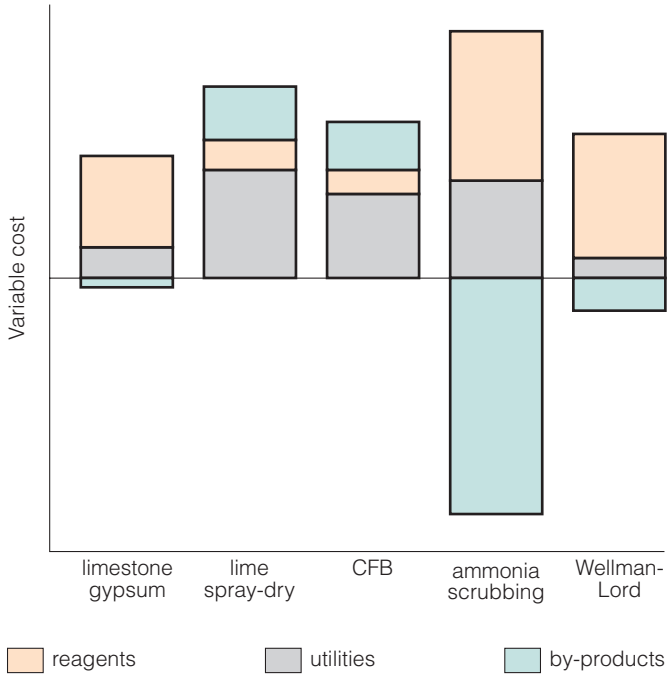


Figure 8 Relative variable operating costs for FGD processes (reproduced from DTI, 2000a)

Wellman-Lord process) have higher capital costs than other processes (*see* Figure 7).

Figure 8 shows the relative variable operating costs for FGD processes (DTI, 2000a). In general, processes that produce a saleable by-product (such as limestone and ammonia processes) have lower net operating costs than those that incur by-product disposal costs. Variable operating costs can generally be in the order of 3 $/MWh. The actual costs can depend critically on the market conditions for sorbents, by-products, and on disposal costs all of which can be location-specific.

Figure 9 shows the relative fixed operating costs for several FGD processes from EPRI's study. It can been seen that

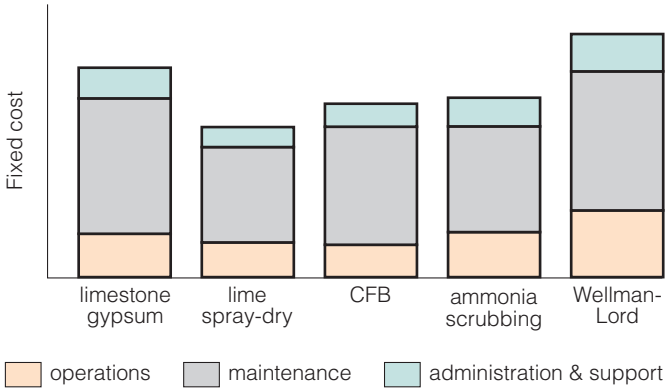


Figure 9 Relative fixed operating costs for FGD processes (reproduced from DTI, 2000a)

maintenance costs depend on the type of unit operation. Operations such as solids handling, or those involving high temperatures or pressures, require more maintenance than a process that involves liquids and gases at ambient conditions.

The costs for FGD plants in China are discussed in a number of papers (Wang, 2000; Xu and others, 1999; Chen and others, 1999; Bo, 1999). The economics of various FGD processes were evaluated according to demonstration projects which are detailed elsewhere (Hao and Lu, 1999; Soud and Wu, 1998; Jiang, 1997). Table 13 compares the costs of five different processes including the wet limestone process at Luohuang Power Plant, spray dry process at Baima Power Plant, LIFAC process at Xiaguan Power Plant, EBA process at Chengdu Power Plant, and the phosphate ammonium fertiliser (PAF) process at Douba Power Plant. The comparative costs differ from those in EPRI's study. It should noted that the costs reflect the Chinese circumstances and depend on the specific conditions of a power plant.

The capital cost for the wet limestone process (436-yuan/kWe, or 48 $/kWe) is not as high as expected. It is similar to the cost for the spray dry process (476 yuan/kWe, or 53 $/kWe) and even lower than the cost for the LIFAC process (600 yuan/kWe, or 67 $/kWe). The LIFAC process requires a high calcium/sulphur ratio and a limestone size of less than 40 µm. This results in a high cost for the sorbent. The levelised cost is as high as 1500 yuan/t of SO_2 removed. The test at Xiaguan Power Plant has shown that a calcium/sulphur ratio of more than 10 was required in order to achieve a 80% SO_2 removal efficiency. The process is considered only suitable for relatively small units, requiring a moderate SO_2 removal efficiency. The EBA process is shown to be expensive. The capital cost for the process (1290 yuan/kWe, or 143 $/kWe) is triple the cost for the wet limestone process. The levelised cost is 2410 yuan/t of SO_2 removed (268 $/t of SO_2 removed) which is four times the cost for the wet process. The capital and levelised costs for the PAF process (developed in China) were found to be 498 yuan/kWe (55 $/kWe) and 570 yuan/t of SO_2 removed (63 $/t of SO_2 removed) respectively. The figures are similar to the costs for the wet limestone process.

Table 13 Comparative costs for FGD processes in China (in Chinese currency) (Wang, 2000)

Process	Wet limestone	Spray dry	LIFAC	EBA	PAF
Operating conditions					
Plant	Luohuang	Baima	Xiaguan	Chengdu	Douba
Unit, MWe	2×360	200	2×125	200	100
Coal sulphur, %	4.0	3.5	0.92 (designed) 0.57 (actual)	–	–
Flue gas treated, m^3/h	2×1080000	820000	2×450000	300000	100000
SO_2 removal efficiency, %	95	80	75	85	70–75
Ca/S ratio	1.1	1.4–1.8	2.5	–	–
Costs					
Capital cost, 104 yuan	31400	9520	15000	9430	1245
Capital cost, yuan/kWe	436	476	600	1290	498
Operating cost, 104 yuan/y	8000	1634	1000	900–1100	154
Levelised cost,					
yuan/t of SO_2 removed	600	770	1500	2410	570
fen/kWh	1.85	2.17	2	2.10	1.03

1 US$ = approximately 9 yuan; 1 yuan = 100 fen

3.8 Summary

The costs of FGD plant are site-specific. The capital costs of FGD plant are considerably influenced by market conditions and other factors, for example, geographical location and the amount of preparatory site work required. In addition, they also depend on technical factors such as volume of flue gas to be treated, SO_2 concentration, and removal efficiency required. The capital costs of different FGD processes vary according to the degree of complexity of the process, the amount of engineering required and other factors. Generally, processes in which the sorbent is processed to produce a saleable by-product or the sorbent is regenerated have higher capital costs than other processes.

The capital costs of wet FGD plant have been falling for the last 30 years, with an overall reduction of as much as 75% in the USA. This is due to advances in technology, reduction or elimination of equipment redundancy, and a competitive market. The average cost for retrofit FGD in the 1970s was approximately 400 $/kWe. This has now dropped to about 100–125 $/kWe, with further reductions anticipated beyond 2000.

Variable operating costs depend on the costs of operation such as sorbents, by-product disposal and utilities (steam, power, water). Fixed operating costs include the costs of labour, maintenance and administration. In general, processes that produce a saleable by-product have lower net operating costs than those that incur by-product disposal costs. The actual costs can depend critically on the market conditions for sorbents, by-products, and on disposal costs all of which can be location-specific. Maintenance costs depend on the type of unit operation. Operations such as solids handling, or those involving high temperatures or pressures, require more maintenance than a process that involves liquids and gases at ambient conditions.

The comparative costs for various processes are generally as follows:

- the wet limestone process has a relatively high capital cost, for example, this can be twice the costs for the magnesium and sodium processes. However, it has a low operating cost mainly because limestone is relatively cheap;
- in comparison with the wet limestone process, the ammonia process has a high capital costs but low operating costs. Its economics improve with increasing coal sulphur level. The capital and operating costs for seawater process are lower than the costs for wet limestone process;
- compared with wet processes, spray dry, sorbent injection and CFB processes represent a lower capital cost but a higher variable operating cost due to the higher and more expensive sorbent use;
- regenerable processes generally have a high capital cost and power consumption. The net variable operating cost is moderate because the processes produce saleable by-products. However, the fixed operating cost is rather high.

4 NOx control

During pulverised coal combustion, some of the nitrogen present in both coal and combustion air is converted into nitric oxide (NO) and nitrogen dioxide (NO_2). These oxides of nitrogen, together commonly referred to as NOx, have been associated with the formation of acid rain and photochemical smog, and the depletion of the earth's protective ozone layer. As a result, legislation has been introduced in recent years in many countries limiting NOx emissions from the combustion processes.

Technologies for control of NOx emissions from pulverised coal-fired power plants can be broadly divided into two groups:

- combustion modifications, effectively constraining the production of NOx during the combustion processes;
- flue gas treatments which remove the NOx from flue gases following its formation.

Additionally, several advanced control systems have been developed in the USA and UK in recent years. These allow optimisation of the various operating parameters to control NOx formation and maximise the boiler performance (DTI, 2000).

The costs for NOx control technologies are again site and performance specific. They depend on various factors including (ETSU, 1997):

- degree of mechanical retrofit difficulty;
- unit size;
- capacity factor;
- life of the boiler;
- uncontrolled NOx levels;
- required NOx reduction;
- average annual heat rate and the impact of NOx control on heat rate;
- financial discount and inflation rates;
- auxiliary power requirements;
- increased fuel costs.

The costs for each control technology are reviewed below. This is then followed by a comparative cost study.

4.1 Combustion modifications

Combustion modifications are used as primary measures to meet NOx emission standards. Optimisation of burners is commonly used to reduce NOx formation. Low-NOx burners and furnace air-staging have been developed and applied on a commercial basis in a number of countries. Furnace fuel-staging using either natural gas or coal as reburn fuel, intelligent control systems, and advanced low-NOx burners with integrated reburn or gas cofiring are also at an advanced stage of development. Combustion modification-based technologies can achieve up to 50% NOx reduction when applied individually. A combination of the techniques provides the potential for even higher NOx reduction. In general, these systems are relatively simple to install and operate, and low in both capital and operating costs. Hence, they are the preferred option in many situations requiring moderate NOx reduction. In recent years, effort has been increasingly concentrated on further development of combustion modifications to achieve maximum NOx reduction in the most cost-effective manner.

Low-NOx burners

Low-NOx burners are designed to reduce both temperature and oxygen availability in the flame by various means. Staged supply of air to the burner was found to be an effective method of NOx reduction. The design of burner to achieve this varies widely between manufacturers. Generally, these designs delay combustion by the way in which the air and fuel are introduced. Oxygen is reduced in zones that are critical for NOx formation and the amount of fuel burnt at peak temperature is also decreased. By staging the addition of air, coal is devolatilised under conditions of low stoichiometry. This promotes the conversion of fuel nitrogen to molecular nitrogen.

Present generation low-NOx burners can achieve 30–60% NOx reduction in commercial plants, depending on the coal used and the operating conditions. Low-NOx burners are currently considered to be the most cost- effective method of achieving NOx reduction from existing coal-fired plants (ETSU, 1997).

The costs for low-NOx burner retrofits have been reported in a wide range. In the UK, a low-NOx burner has a capital cost of 8–23 $/kWe and a levelised cost of 110–210 $/t NOx removal (ETSU, 1997). A study by Richardson and Hall (1999) has shown that the retrofits cost typically in the order of 9 $/kWe. The costs in the USA are reported to be 7–36 $/kWe and 0.15–0.54 mills/kWh respectively (FGD & DeNOx, 2000b). An evaluation of a retrofit on a 300 MWe wall-fired boiler has shown that a low-NOx burner reduced NOx emissions by 37% and cost around 14 $/kWe (US DOE, 2001). The US EPA data show the capital cost for a low-NOx burner in the range of 650–8300 $/MBtu (616–7867 $/GJ). The operating cost can vary from 340 to 1500 $/MBtu (322–1422 $/GJ) and the levelised cost can range from 240 to 4300 $/t NOx removal (Hilton, 2000).

Richardson and Hall (1999) estimated the costs of a low-NOx burner retrofit on a 500 MWe front wall-fired boiler. The costs were presented in mills/kWh and $/t NOx removal. The following assumptions were made: 10 years of plant operation; 40% unit load factor; 40% NOx emission reduction; 9 $/kWe capital cost. The capital costs for reduced NOx emissions were estimated to be 0.3 mills/kWh and 200 $/t NOx removal. It was assumed that the low-NOx burner retrofit would not increase the carbon-in-ash (CIA) level and affect the saleability of the ash. Mill modification was not required. The costs for minimising CIA, lost saleable ash and increased ash disposal were therefore all zero. The

low-NOx burner required a higher level of operation and maintenance. It was assumed that the cost for operation and maintenance labour was increased by 1%. This would increase the labour cost by 0.015 mills/kWh and 11 $/t NOx removal. The total costs for low-NOx burner retrofit were therefore 0.315mills/kWh and 211 $/t NOx removal.

As an alternative to new low-NOx burner retrofits, conventional burners can be modified by control of the fuel/air mixing, reducing NOx formation (Melick and others, 2000). The low-NOx burner modification has been applied to a number of wall-fired boilers with both circular and cell burners. The NOx reduction levels have generally been comparable to those achieved by new low-NOx burners.

The costs for low-NOx burner modification are reported to be much lower than those for new low-NOx burners since the hardware changes are simple (Melick and others, 2000, 1999; Melick and Payne, 1999). In most cases, there is no need to change flame scanning and ignition equipment, electric actuator drives or registers. There is no need to modify coal conduits or the windbox. This results in a relatively low capital cost for the modification, typically in the order of only 25 to 50% of the cost for new low-NOx burners. As the modification achieves comparable NOx reduction levels to new low-NOx burners, it can be much more cost effective on a $/t of NOx removal basis.

The suitability of low-NOx burner modification depends on the condition of existing burners. If the burner steel is in good condition with a reasonable remaining life, the burners are suitable for modification. Specific cases where the modification is particularly attractive include:

- older units where the expense of new burners are difficult to justify over the remaining boiler life;
- units with a high firing density or other conditions which limit the effectiveness of new low-NOx burners;
- units where the retrofit of new low-NOx burners will require substantial windbox modifications, replacement of ignition/scanning equipment which are uneconomic;
- units with first generation low-NOx burners, requiring further NOx reduction.

Furnace air-staging

Furnace air-staging involves staging the supply of combustion air, with 70–90% being supplied to the burners (primary air) and the rest to the furnace at a level above the burners (overfire air). The primary air is mixed with the fuel, producing a relatively low-temperature, oxygen-deficient, fuel-rich zone near the burner which helps to reduce the formation of fuel-NOx. The overfire air is injected above the primary combustion zone, through a special windbox with overfire air ports and/or nozzles mounted above the top level of burners, to achieve complete combustion. A relatively low-temperature secondary combustion zone limits the formation of thermal-NOx (DTI, 2000).

Furnace air-staging is a well proven commercially available technology for both wall-fired and tangentially-fired boilers. In a wall-fired boiler, separate overfire air injectors are located above the main burners, whilst in a tangentially-fired boiler, the system can be installed as close-coupled overfire air (CCOFA) or separated overfire air (SOFA). The latter design is typically used in the Alstom low-NOx concentric firing system (LNCFS™). The LNCFS I, II and III provide progressively greater NOx reduction attributable principally to greater separation of the overfire air from the main burner zone and greater quantities of overfire air. The TFS 2000™® is an enhanced low-NOx firing system based on the LNCFS III design. In general, furnace air-staging reduces NOx emissions by 20–40%. However, the LNCFS™ and TFS 2000™® can achieve up to 55% and 60% NOx reduction respectively (World Bank, 2001; Gessner and others, 1999).

The costs for furnace air-staging are similar to those for low-NOx burners. ETSU (1997) reported that the capital cost for furnace air-staging is about 8–23 $/kWe and the levelised cost ranges 110–210 $/t NOx removed. If the system is combined with low-NOx burners, the capital cost will increase to 15–30 $/kWe. The levelised cost will be around 110–240 $/t NOx removal, with a relatively small increase. A study by Richardson and Hall (1999) has shown that the retrofits of furnace air-staging in tangentially-fired boilers are generally more expensive than those in wall-fired boilers. The wall-fired retrofits cost typically 5–11 $/kWe, whilst the tangentially-fired retrofits cost 11–23 $/kWe depending on the options involved.

The cost for the TFS 2000™® system at W A Parish Unit 7 (USA) was reported (Gessner and others, 1999; Hoh and Jennings, 2000). The installation of the system has resulted in a 60% NOx reduction. It costs approximately $5,000,000. This is relatively low compared with the cost for an SCR system which would cost nearly $40,000,000 for the same level of NOx reduction. In addition, there are no additional operating costs associated with the TFS 2000™® system. In contrast an SCR system would require chemicals or catalyst, hence increasing the overall operating costs.

Flue gas recirculation

Flue gas recirculation (FGR) involves recirculating 20–30% of the flue gas, usually at 300–400°C, back into the furnace or to the burners. This decreases the flame temperature and oxygen concentration and thus helps to reduce the formation of thermal NOx. In coal-fired units, the NOx reduction achieved is less than 20%, due to a relatively low contribution of thermal NOx to total NOx formation. FGR can be used in combination with low-NOx burners, overfire air or air/fuel ratio control, achieving substantial NOx reductions (Wu, 2000; Benz, 2000).

Retrofit of FGR on an existing boiler requires a system to extract flue gas from its back end, additional ductwork, a fan, and a fly ash collection device to remove particulates prior to recirculation. FGR systems also need careful design to avoid problems associated with flame stability and furnace heat/gas flow distribution.

The cost of FGR is reported to be considerably lower than that of burner retrofits or flue gas treatments (Benz, 2000). Dharmarajan and others (1999) evaluated the economics of FGR at a number of gas-fired units of Central & South West Services in the USA. The installation of FGR included the following major fieldwork:

- tapping into individual ducts feeding their respective stacks, to attach the recirculation gas ducts;
- installation of support structure for the dampers and the flue gas ducts;
- inter-connection of the recirculation gas ducts to the ducts feeding the stacks and the fan inlets;
- installation of the forced draft fan inlet shrouds.

The installation has resulted in up to 60% reduction in NOx emissions through a defined load range. The installed costs ranged from 1.3 $/kWe to 1.5 $/kWe, depending on the unit size. The operating costs would increase due to the increased fan power consumption.

Furnace fuel-staging (reburn)

Reburn involves staged supply of fuel and combustion air, in the furnace, to create three distinct combustion zones - primary zone, reburn zone and burnout zone. In the primary zone, coal is fired with some excess air, producing NOx from both fuel-bound and combustion air nitrogen. A secondary fuel (reburn fuel), which can be natural gas, coal or oil, is injected above the primary zone to create the reburn zone. Hydrocarbon radicals released from the reburn fuel in this zone react with NOx formed in the primary zone to reduce it to molecular nitrogen. Finally, the remaining combustion air is injected above the reburn zone (the burnout zone) to complete combustion.

Natural gas reburn alone can achieve over 50% NOx reduction, for example, at Longannet Power Station in the UK (DTI, 2000c). A combination of low-NOx burners and a gas reburn system can give over 70% NOx reduction. The technology is now considered sufficiently developed for commercial application (DTI, 2000b). Coal reburn is being developed and demonstrated. The results have shown a 50–60% NOx reduction is possible (ETSU, 1999; Wu, 2000).

The retrofit of reburn to existing plants is a relatively complex issue, with many site specific conditions needing to be taken into account. The capital cost depends on the size of the unit, ease of the retrofit, the control system upgrade requirements and, for gas reburn, availability of natural gas at the plant (Farzan and others, 1999). The capital cost for gas reburn is significantly reduced when natural gas is already available on site (no gas pipeline required). The retrofits in the UK cost typically about 15–20 $/kWe for gas, coal or oil reburn, excluding the cost of any gas pipeline (Richardson and Hall, 1999). The cost of gas reburn retrofit on a 600 MWe unit at Longannet Power Station was around 19.5 million US dollars but it is likely that this will reduce to 15 million US dollars for further unit retrofits (Fernando, 2000). In the USA, the nominal turnkey cost for a gas reburn retrofit to a 300 MWe unit is approximately 10–15 $/kWe (Engelhardt and others, 2000a,b). However, a typical cost in the 15–20 $/kWe range (including engineering, equipment, labour) was also reported (Schimmoller, 1998).

The operating cost for reburn retrofit is near zero except for the differential cost of the reburn fuel over the main fuel (Engelhardt and others, 2000a; 1999). For coal reburn, the differential cost is zero. Reburn fuels such as natural gas or oil are generally more expensive than the main fuel. However, part of the differential cost may be offset by reductions in SO_2 emissions, and reductions in ash and pulveriser power. The operating cost for gas reburn is determined by the amount of the gas usage and the differential gas to coal price. This can be potentially reduced should the price of gas drop (Richardson and Hall, 1999).

Farzan and others (1999) reported the retrofits of gas reburn technology to three cyclone-fired boilers at Eastman Kodak Company in the USA. The retrofits were more capital intensive due to the need for new boiler controls. The total capital cost for the Unit 43 retrofit was approximately 5.9 millions US dollars. The retrofits at Unit 41 and 42 cost around 3 millions US dollars. The capital cost per unit capacity would typically fall into the range of 10–20 $/kWe. The operating cost was estimated from the differential price of gas and coal. For Unit 43, the differential gas/coal price was 1.74 $/MBtu (1.65 $/GJ). The respective levelised cost was therefore 1335 $/t of NOx removal. The economics for Unit 41 and 42 were more favourable since less gas was used in the units. While a full economic study was not performed, the levelised cost was estimated to be about 50% of that for Unit 43. It should be noted that the above analysis did not include the additional advantage of the gas reburn in reducing SO_2 and CO_2 emissions.

The US DOE (2001) evaluated a gas reburn/low-NOx burners retrofit on a 300 MWe wall-fired boiler. The retrofit is expected to achieve at least 64% NOx reduction with a gas heat input of 12.5%. The capital cost for the installation is estimated to be 26.01 $/kWe (1996 dollars) plus gas pipeline costs, if required. This cost includes both equipment and installation costs and a 15% contingency. The capital costs for gas reburn and low-NOx burners can be easily separated from one another because they are independent systems. The capital cost for the gas reburn system only is estimated at 12.14 $/kWe. The low-NOx burners system capital cost is 13.87 $/kWe.

The operating costs almost entirely depend on the differential cost of natural gas and coal. A fuel differential cost of 1.00 $/MBtu (0.95 $/GJ) was used because gas costs more than coal on a heating value basis. The boiler efficiency was estimated to decline by 0.80%. The cost of this decline was calculated using a composite fuel cost of 1.67 $/MBtu (1.58 $/GJ). Overfire air booster and cooling fan auxiliary loads will be partially offset by lower loads on the pulverizers. No additional operating labour is required, but there is an increase in maintenance costs. Allowances were also made for overhead, taxes, and insurance. It should be noted that the operating costs are reduced by the value of the SO_2 emission credits received due to absence of sulfur in the gas. An SO_2 credit allowance of 95 $/t (1996 dollars) was therefore assumed. Based on these assumptions, the net operating cost for the retrofit is estimated to be 2.14 million dollars per year. The levelised cost is around 786 $/t of NOx removal.

4.2 Flue gas treatments

Flue gas treatments for NOx control include selective

catalytic reduction (SCR) and selective non-catalytic reduction (SNCR). They have been developed and applied on a commercial basis in a number of countries. SCR and SNCR are capable of achieving up to 90% and 50% NOx reduction respectively. Compared to combustion modifications, the technologies are generally more expensive to install and operate. SCR systems for a single plant can cost more than 100 million US dollars (Mcllvaine, 1999). They are therefore recommended for the most stringent NOx control. A combination of SCR with combustion modification and/or SNCR provides the potential of overall high NOx reduction in a cost-effective way.

Selective catalytic reduction

In an SCR system, vaporised ammonia (NH_3) is injected into the flue gas stream at about 300–400°C which is passed over a catalyst. The NOx is then reduced by the NH_3 to molecular nitrogen. The catalyst is usually based on oxides of titanium, vanadium and tungsten. The arrangement of catalyst can be of flat plate or honeycomb configuration. SCR systems generally achieve a 80–90% NOx reduction. They have been used commercially on coal-fired power plants burning mainly low-sulphur coals, and in some cases medium-sulphur coals. There is little experience with high-sulphur coals (Stallings, 1999).

SCR systems are most commonly installed in a high-dust configuration (before particulate removal), which lowers the installation costs and eliminates the need for expensive flue gas reheating. For economic reasons, anhydrous ammonia is used as reagent, accounting for over 90% of world SCR applications. Catalyst selection is fairly evenly split between plate and monolith designs, while emphasising the importance of matching catalyst to the fuel used (Schimmoller, 2000).

The capital costs for an SCR system are associated with (Kitto and others, 1999):

- catalyst and reactor system;
- flow control skid and valving system;
- ammonia injection grid;
- ammonia storage;
- all piping;
- flues, expansion joints and dampers;
- fan upgrades/booster fans (as required);
- air heater changes (if required);
- foundations, structural steel and electricals;
- installation.

It should be noted, however, that the complexity of retrofit can significantly affect the capital costs. The operating costs may include the following major items:

- ammonia usage;
- pressure drop changes;
- excess air change;
- unburned carbon change;
- ash disposal (costs or credits);
- catalyst replacement;
- vaporisation/injection energy requirements;
- other auxiliary power usage.

The capital costs for an SCR system are reported to range

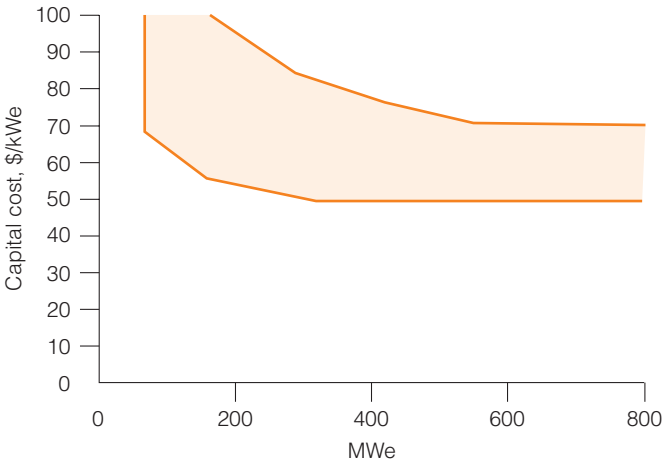


Figure 10 Capital cost range for most coal-fired SCR retrofits (Andover Technology Partners, 2001)

from 50 to 150 $/ kWe, depending on the level of NOx removal and other site-specific conditions (Breeze, 1998; Schimmoller, 2000). Figure 10 shows the capital costs can vary from 50 to 100 $/kWe for unit size of 50–800 MWe and NOx reduction of 60–90%. In the UK, SCR has a capital cost of 45–90 $/kWe and a levelised cost of 800–1000 $/t NOx removal (ETSU, 1997). A study by Richardson and Hall (1999) has shown that the retrofits cost typically in the order of 100 $/kWe. The costs in the USA are reported to be 40–90 $/kWe and 1.30–2.40 mills/kWh respectively (FGD & DeNOx, 2000b). However, an annual cost of only 0.35–0.45 mills/kWh was reported for the 330 MWe Indiantown Plant, where SCR is combined with low-NOx burners (Bullock and Taylor, 2000). The US EPA data show the capital cost for an SCR system is in the range of 2,400–20,000 $/MBtu (2275–18,957 $/GJ). The operating cost can vary from 1500 to 5800 $/MBtu (1422–5498 $/GJ) and the levelised cost can range from 1,800 to 10,900 $/t NOx removal (Hilton, 2000).

The retrofits of SCR on 500 MWe units where low-NOx burners are already in use were evaluated (Gaikwad and Boward, 1999). The evaluations were based on the following assumptions: ammonia cost at 260 $/t; catalyst cost at 8000 $/m³; catalyst replacement at every 12,000 hours for high-sulphur coal applications. The capital costs for SCR retrofits were estimated to be 85–90 $/kWe. The respective fixed operating costs varied from 0.43 million dollars per year to 0.45 million dollars per year. The variable operating costs were in the range of 2.38–2.81 million dollars per year.

Gaikwad and others (2000) estimated the variable operating costs of an SCR system for a 600 MWe coal-fired unit. The SCR system is assumed to operate on a seasonal basis over a period of 20 years. The ammonia cost is based on 150 $/t anhydrous ammonia delivered to the site. The catalyst cost is assumed to be 8000 $/m³ of replacement catalyst. The power cost includes gas side pressure drop across the SCR system (fan power), ammonia vaporisation and other small parasitic loads at 2.5 ¢/kWh. The total variable operating costs were therefore found to be 795,000–923,000 $/y, depending on the reactor design.

Table 14 Costs of SCR as a function of inlet NOx concentration and unit size (Schmidt and others, 1999)

Unit capacity, MWe	Cost type	NOx concentration at SCR inlet, lb/MBtu (mg/m³)			
		>0.60 (738)	0.40–0.60 (492–738)	0.25–0.40 (308–492)	<0.25 (308)
>400	Capital cost, $/kWe	53	51	50	50
	Levelised operating cost, $/t	260	345	444	560
200–400	Capital cost, $/kWe	65	61	60	60
	Levelised operating cost, $/t	260	345	444	560
50–200	Capital cost, $/kWe	91	86	84	84
	Levelised operating cost, $/t	260	345	444	560
<50	Capital cost, $/kWe	120	115	115	115
	Levelised operating cost, $/t	260	345	444	560

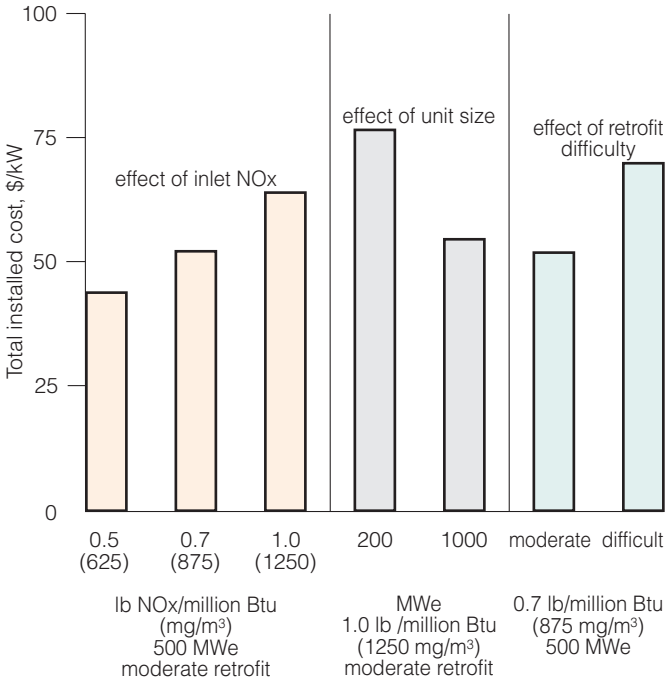


Figure 11 Effects of inlet NOx unit size and retrofit difficulty on the cost of SCR (Kitto and others, 1999)

Table 14 shows the capital and levelised operating costs as a function of inlet NOx concentration and unit size. The SCR system is based on a 80% NOx reduction, with a retrofit factor (retrofit difficulty) of 1.5. It is assumed that the SCR system operates only during the five-month summertime ozone season. During the remainder of the year, there would be no consumption of ammonia. The price of anhydrous ammonia is assumed to be 300 $/t. The capital cost is shown to range from 50 $/kWe to 120 $/kWe, mainly depending on unit size. There is only a small decrease in capital cost with decreasing inlet NOx concentration. The levelised operating cost falls in the range of 260–560 $/t of NOx removed. In contrast to the capital cost, the levelised operating cost is not affected by unit size but increases dramatically as the inlet NOx concentration decreases.

Kitto and others (1999) have also shown how the retrofit costs for SCR systems are affected by inlet NOx concentration, unit size and retrofit difficulty (Figure 11). The capital costs are estimated for a target NOx emission level of 0.15 lb/MBtu (185 mg/m³). The results show that the capital costs vary widely with all the factors examined as follows:

- capital costs can increase by over 30% as the inlet NOx concentration increases from 0.5 lb/MBtu (615 mg/m³) to 1.0 lb/MBtu (1230 mg/m³);
- capital costs can decline by approximately 30% if unit size increases from 200 MWe to 1000 MWe;
- capital costs can increase by 20–35% as the degree of retrofit difficulty moves from moderate to difficult (including significant fan upgrades plus major flue, structural steel and foundation changes).

The levelised costs ($/t of NOx removal) follow similar trends for unit size and degree of retrofit difficulty. However, the effect of inlet NOx concentration on a $/t basis is not pronounced. This is because the ammonia and related energy costs change directly with the NOx reduction.

Selective non-catalytic reduction

In the SNCR process, an amine-based chemical reagent, most commonly NH_3 or urea, is injected in aqueous or gaseous form into the flue gas at about 900–1100°C. The NOx is then reduced to molecular nitrogen. A 30–50% NOx reduction is usually achieved, depending on the reagent type and operating conditions. In general, at temperatures above 1000°C, the NOx reduction rate tends to decrease due to thermal decomposition of NH_3. On the other hand, the NOx removal rate decreases considerably below 1000°C and the amount of unreacted NH_3 in the flue gas goes up. The technology has found limited commercial application, mainly in small-size (<150 MWe) coal-fired units. It was reported, however, that the urea-based SNCR technology has recently been demonstrated at a 600 MWe unit in the USA (US DOE, 2000).

The capital costs for an SNCR system are reported to be around 5–10 $/kWe for a new plant and 10–20 $/kWe for retrofit to an existing plant (Breeze, 1998). A study by Richardson and Hall (1999) has shown that the retrofits in the UK cost typically in the order of 13 $/kWe. In the USA, the capital costs for SNCR can vary from 2–18 $/kWe,

Table 15 Costs for SNCR retrofits at coal-fired units in the USA (Andover Technology Partners, 2001)

NOx reduction at 40% from lb/MBtu (mg/m³)	to	Capital cost	Chemical utilisation	Annual control		Seasonal control	
		$/kWe	%	$/t	$/MWh	$/t	$/MWh
0.45 (554)	0.27 (332)	15	35–60	860–1160	0.78–1.05	1370–1670	0.51–0.63
1.00 (1230)	0.60 (738)	15	35–60	620–920	1.24–1.84	845–1145	0.71–0.95
1.50 (1845)	0.90 (1107)	15	35–60	550–850	1.66–2.55	705–1005	0.88–1.25

depending on the system complexity and unit size (Southern Illinois University, 2001). The US EPA data show that the capital costs for an SNCR system are in the range of 1600–3300 $/MBtu (1517–3128 $/GJ). The operating costs can vary from 680 to 1200 $/MBtu (645–1137 $/GJ) and the levelised costs can range from 700 to 1300 $/t NOx removal (Hilton, 2000).

Schmidt and others (1999) estimated the costs for SNCR retrofits at fossil-fired units in the USA. The capital cost for SNCR was based on 15 $/kWe. For the 218 units studied, the average levelised cost was estimated to be around 1704 $/t NOx removal at a 25% NOx reduction. If SNCR had a 40% NOx reduction, the average levelised cost would drop to about 1196 $/t NOx removal.

Another cost analysis for SNCR retrofits at coal-fired units in the USA was prepared by Andover Technology Partners (2001). Table 15 summaries the results of the analysis. A 40% NOx reduction was assumed. The capital cost for SNCR was again based at 15 $/kWe. Depending on the NOx levels and chemical utilisation, the levelised cost falls in the range of 550–1160 $/t NOx removal on a basis of annual control. The respective annual cost varies from 0.78 $/MWh to 2.55 $/MWh. For seasonal control, the levelised cost increases to 705–1670 $/t NOx removal. However, the annual cost drops down to 0.51–1.25 $/MWh.

Gaikwad and Boward (1999) evaluated the retrofits of SNCR on a number of 150 MWe units where low-NOx burners are already in use. The evaluations were based on the following assumptions: ammonia cost at 260 $/t; urea cost at 0.91 $/gallon (0.24 $/L). The capital costs for the retrofits were estimated to be around 15 $/kWe. The fixed and variable operating costs were about 0.11 million dollars per year and 0.56 million dollars per year respectively.

In the UK, Richardson and Hall (1999) estimated the costs of SNCR retrofit on a 500 MWe front wall-fired boiler. The cost estimates were based on the following assumptions: 10 years of plant operating period; 40% of unit load factor; 40% of NOx emission reduction; 13 $/kWe of capital cost. The capital costs for reduced NOx emissions, in mills/kWh and $/t NOx removal, were estimated to be 0.47 mills/kWh and 526 $/t NOx removal respectively. The operating costs included the following items:

- cost for increased ash disposal due to ammonia slip: 1.20 mills/kWh, 1351 $/t NOx removal;
- cost for lost salable ash due to ammonia slip: 0.14 mills/kWh, 156 $/t NOx removal;
- cost of increased power due to ammonia injection system: 0.03 mills/kWh, 31 $/t NOx removal;
- cost of reagent consumption (at the anhydrous ammonia cost of 225 $/t): 0.21 mills/kWh, 241 $/t NOx removal;
- cost of forced outages for maintenance (assuming the additional outages at 5% of lost operating time): 5.45 mills/kWh, 6128 $/t NOx removal;
- cost for operating and maintenance fixed labour (at 5% of the capital cost): 0.11 mills/kWh, 119 $/t NOx removal.

The total costs for SNCR retrofit were therefore estimated to be around 7.61 mills/kWh and 8553 $/t NOx removal.

4.3 Hybrid NOx control

In many cases, a combination of technologies may be the most cost-effective method to reduce NOx emissions. The combinations may include SCR or SNCR with combustion modifications and SCR with SNCR which are discussed below.

SCR/combustion modifications

The costs of SCR can be reduced when it is used in combination with combustion modifications. This is because combustion modifications lower the inlet NOx concentration, which reduces the catalyst volume, support systems and installation cost of SCR. Cichanowicz and Broske (1999) reported that a reduction in NOx levels from 450 ppm to 275 ppm resulted in a 3 year payback in terms of reduced SCR cost. It should be noted, however, that combustion measures may not be incorporated on units where furnace corrosion or other operating problems are anticipated.

Figure 12 shows the effects of combustion modifications on SCR costs (Kitto and others, 1999). The baseline NOx level is assumed to be 1.00 lb/MBtu (1230 mg/m³). In Case 1, low-NOx burners and overfire air are added, reducing the NOx level to 0.34 lb/MBtu (418 mg/m³) at the inlet of SCR. In Case 2, SCR is used alone. Both cases are designed to achieve the emission level of 0.15 lb/MBtu (185 mg/m³). It can be seen that the addition of low-NOx burners and overfire air reduces the capital costs for catalyst, support systems and installation of SCR. This can actually reduce the overall capital cost of the system as long as the combustion retrofit costs are moderate. The levelised costs are also considerably reduced by the addition of combustion measures, because of the minimal operating costs of the upgraded combustion system.

SNCR/combustion modification

A combination of urea-based SNCR with low-NOx burners has recently been demonstrated at the 600 MWe Cardinal

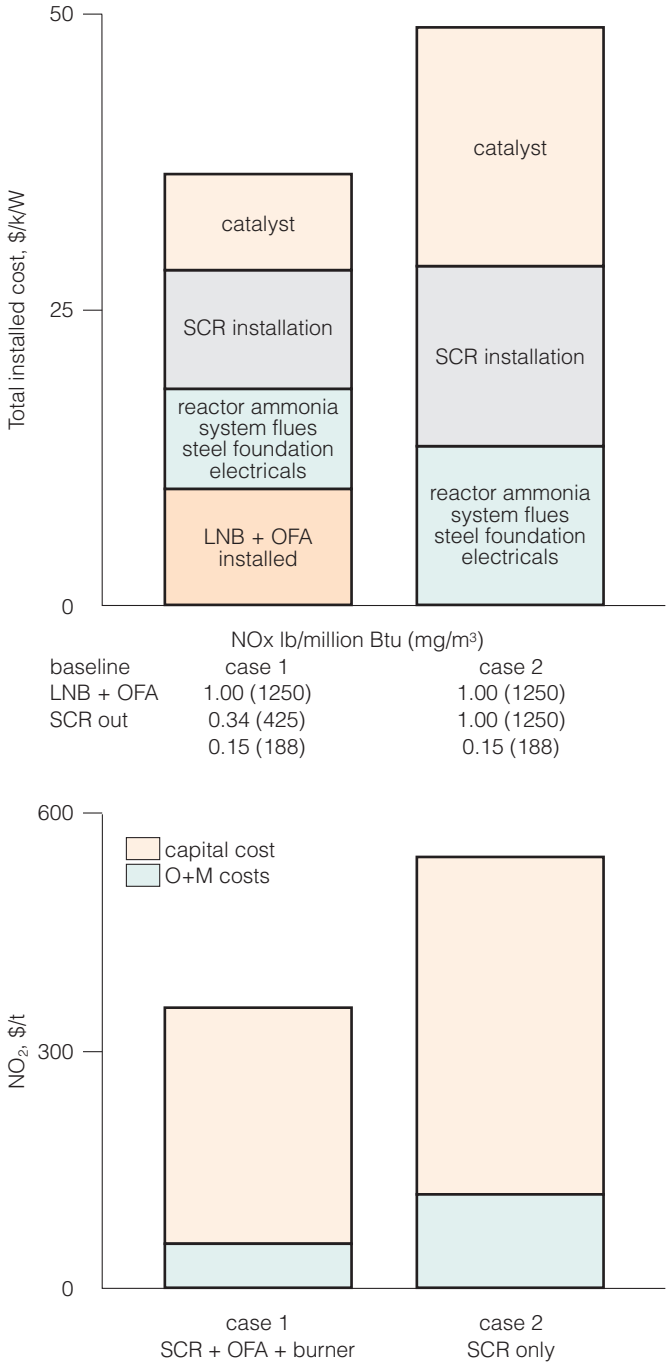


Figure 12 Effects of combustion modifications on SCR costs (Kitto and others, 1999)

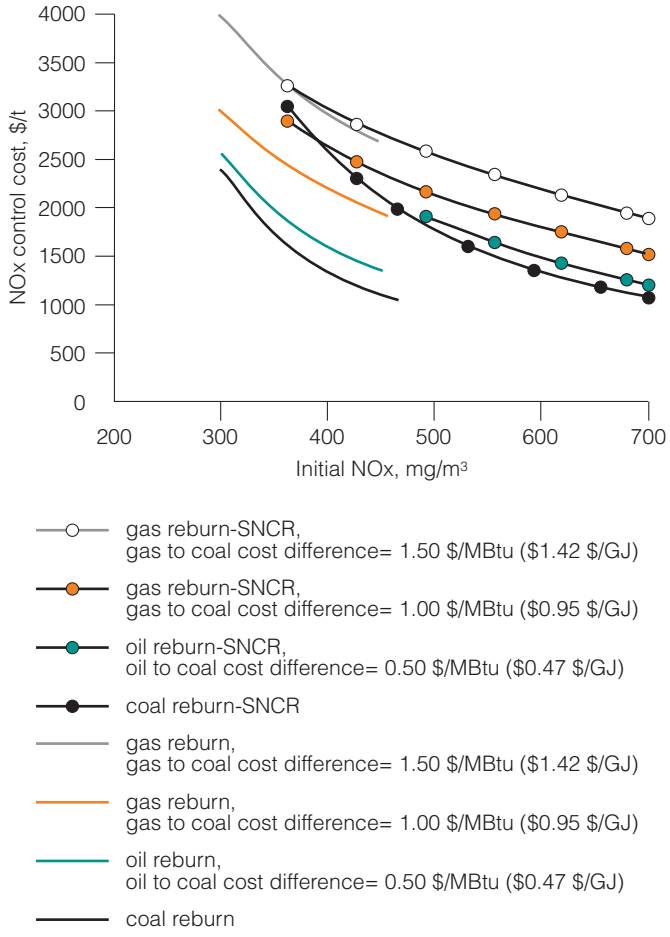


Figure 13 Costs of reburn and reburn-SCR for a 300 MWe unit (Folsom and Tyson, 2000)

Plant Unit 1 in the USA (Malone and others, 1999; US DOE, 2000). The hybrid technology has achieved a 65% NOx reduction. The capital cost for the retrofit is relatively low, approximately 12 $/kWe.

A combination of reburn with SNCR is reported to achieve a higher reduction in NOx emissions at a relatively low cost. The technology reduces NOx emissions by up to 73% (Folsom and Tyson, 2000). In the case of gas reburn-SNCR, a reduction of 85% has been claimed. The capital cost for gas reburn-SNCR is reportedly 50–75% less than that for SCR (GRI, 2001).

Folsom and Tyson (1999, 2000) compared the costs of reburn, reburn-SNCR, and overfire air-SCR. A combination of overfire air with SCR is used here because it has a lower cost compared to SCR alone (due to a smaller catalyst volume). Figure 13 shows the costs of reburn and reburn-SNCR for a 300 MWe unit with NOx reduced from various initial levels to 0.15 lb/MBtu (185 mg/m^3). The reburn fuels include gas, oil and coal with variable reburn fuel to coal cost differential. The maximum NOx reductions for reburn and reburn-SNCR are set at 60% and 73%, respectively. In order to achieve the 0.15 lb/MBtu (185 mg/m^3) level, the respective maximum initial levels are therefore 0.38 lb/MBtu (467 mg/m^3) and 0.56 lb/MBtu (689 mg/m^3). It is shown that the cost in all the cases decreases as the initial NOx concentration increases. The cost of technology is also influenced by the reburn fuel to coal cost differential. For the initial NOx levels where reburn can be applied, reburn has a lower cost than advanced reburn except at the highest reburn fuel to coal cost differential. For the higher initial NOx levels, the cost for reburn-SNCR decreases down to the same cost range of reburn.

Figure 14 compares the full cost range of reburn and reburn-SNCR from Figure 13 (shown as a closed region) with the costs of overfire air-SCR. It is shown that the costs for all the SCR cases are much higher than those for reburn/reburn-SNCR at the same initial NOx levels. For example, at the initial NOx of 0.40 lb/MBtu (492 mg/m^3), the highest cost for reburn-SNCR is 68% of the cost for the SCR = 60/kWe case. As with reburn/reburn-SNCR, the cost for SCR

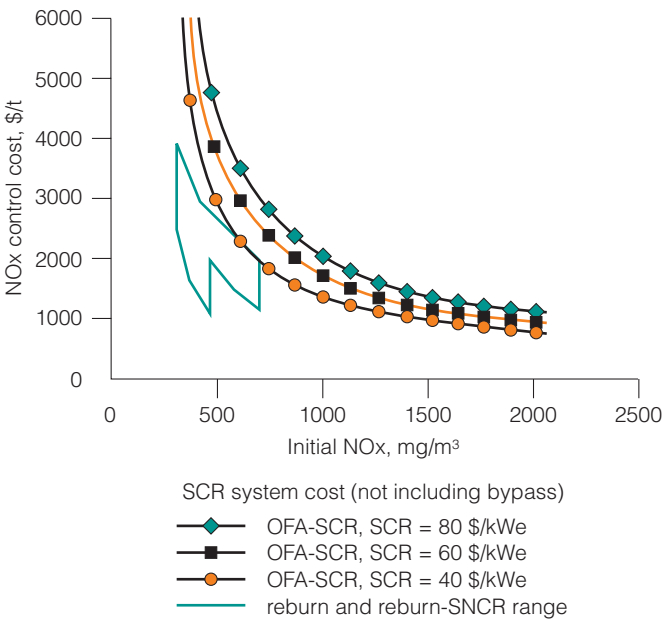


Figure 14 Comparative costs of reburn/reburn-SNCR and overfire air-SCR (Folsom and Tyson, 2000)

decreases as the initial NOx concentration increases. For higher initial NOx levels, the cost for SCR is seen to drop into the $/t range of reburn/reburn-SNCR at lower initial NOx levels.

Similar analyses were conducted for a range of unit capacities from 50 MWe to 1300 MWe. The capacity effect is shown to be greater for overfire air-SCR. At high capacity, the costs for overfire air-SCR are close to those for reburn/reburn-SNCR. In summary, for control to 0.15 lb/MBtu (185 mg/m^3), the lowest cost technology depends on the initial NOx levels. Reburn has the lowest cost for the initial NOx less than about 0.38 lb/MBtu (467 mg/m^3). For the initial NOx range of 0.38–0.56 lb/MBtu (467–689 mg/m^3), reburn-SNCR provides the lowest cost option. Overfire air-SCR is the lowest cost technology for the initial NOx levels greater than 0.56 lb/MBtu (689 mg/m^3), where reburn/reburn-SNCR cannot achieve the required NOx reduction.

Table 16 shows the costs for urea SNCR, gas reburn, and urea SNCR/gas reburn hybrid. When urea SNCR or gas reburn is applied individually, the NOx emissions are reduced from 1.0 lb/MBtu (1230 mg/m^3) to 0.4 lb/MBtu (492 mg/m^3). The hybrid technology lowers the emissions from 1.0 lb/MBtu (1230 mg/m^3) to 0.36 lb/MBtu (443 mg/m^3). It is shown that urea SNCR and gas reburn have comparable economics at the same NOx reduction. However, the combination of two technologies results in considerably lower costs while even achieving a slightly higher NOx reduction. The total annual and levelised costs for the hybrid technology are 2.66 million dollars and 729 $/t NOx removal respectively, about two thirds of the costs for the individual technologies.

SNCR/SCR

A combined SNCR/SCR process is being investigated with a view to reducing the size of SCR reactor and minimising the level of ammonia slip. In the hybrid system, NOx levels are significantly lowered by the upstream SNCR. The size of the downstream SCR is therefore reduced. The smaller catalyst reactor and elimination of reheat requirement would result in considerable cost savings. The hybrid technology can also reduce the potential for oxidation of SO_2 to SO_3 and hence the formation of catalyst-plugging ammonium salts. The ammonia slip would also be minimised because the unconverted ammonia from the SNCR is utilised as the feed in the SCR reactor.

At present, the SNCR/SCR hybrid is only being demonstrated at pilot scales. Its operating conditions need to be refined before it becomes commercialised. In the UK, the capital cost for the hybrid technology is estimated to be around 45 $/kWe. The levelised cost is approximately 5791 $/t of NOx removed for a 500 MWe boiler, based on 10 years operating period, 40% load factor and 50% NOx reduction (Richardson and Hall, 1999).

Table 16 Costs for urea SNCR, gas reburn, and urea SNCR/gas reburn hybrid (Andover Technology Partners, 2001)

Cost type	Urea SNCR NOx reduction: 1.0–0.4 lb/MBtu (1230–492 mg/m^3)	Gas reburn NOx reduction: 1.0–04 lb/MBtu (1230–492 mg/m^3)	Urea SNCR/gas reburn NOx reduction: 1.0–0.36 lb/MBtu (1230–443 mg/m^3)
Reburn annual costs, M$		3.84	1.51
Annual NOx removed, t		3416	2280
$/MWh		3.37	1.33
SNCR annual costs, M$	3.90		1.15
Annual NOx removed, t	3416		1367
$/MWh	3.42		1.01
Total annual costs, M$	3.90	3.84	2.66
Annual NOx removed, t	3416	3416	3647
$/MWh	3.42	3.37	2.34
$/t NOx removed	1142	1124	729

4.4 Comparative costs

Figure 15 compares the levelised costs for several NOx reduction methods in a wall-fired unit (Schimmoller, 1998). The baseline NOx emissions are assumed to be 0.50 lb/MBtu (615 mg/m^3). The costs for reburn are based on a gas to coal price differential of 0.50–1.50 $/MBtu (0.47–1.42 $/GJ) and a zero price differential between orimulsion (a liquid fossil fuel composed of 70% bitumen and 30% water) and coal. The SNCR technology uses aqueous ammonia.

It is shown that overfire air and SNCR have relatively low costs, but achieve less (25–40%) NOx reduction. Reburn technologies, using coal, gas or orimulsion, have comparable economics to overfire air and SNCR, while achieving 50–60% NOx reductions. In certain instances, orimulsion may be the best fuel option for reburn because it is less expensive than natural gas (while achieving comparable NOx reductions) and more effective than coal (at a similar fuel price). SCR offers the greatest NOx reductions but at considerably higher costs. The annual and levelised costs for SCR are around 12–13 million dollars per year and 2000 $/t NOx removed respectively.

The costs for various NOx reduction technologies for 20 power stations in the UK were reported by Richardson and Hall (1999). The costs, in mills/kWh and $/t NOx removed, were estimated for unit load factors of 10%, 40% and 75% and operating periods of 5, 10 and 15 years. The results have shown a number of trends:

- combustion modifications appear to be more economically favourable. Their costs, either in mills/kWh or $/t NOx removed, remain fairly constant through variation in the unit load factor and operating time. This is because the combustion technologies have a relatively low capital outlay and their operating costs should be relatively constant;
- in contrast to combustion modifications, flue gas treatment technologies (except for SNCR) generally have a significant capital outlay. There is a noticeable drop in costs as operating time passes, particularly for low unit load factors. SNCR reverses the trend somehow. This is because SNCR has much less initial capital outlay compared to SCR but like SCR there is a significant operating cost associated with the consumption of ammonia. Both the factors result in a slight increase in costs at a constant load factor, as operating time passes, probably due to inflation;
- for costs correlated against the unit load factor, SNCR represents a similar trend to combustion measures. The SNCR costs in mills/kWh and $/t NOx removed, remain reasonably constant across the range of load factors, although they are higher.

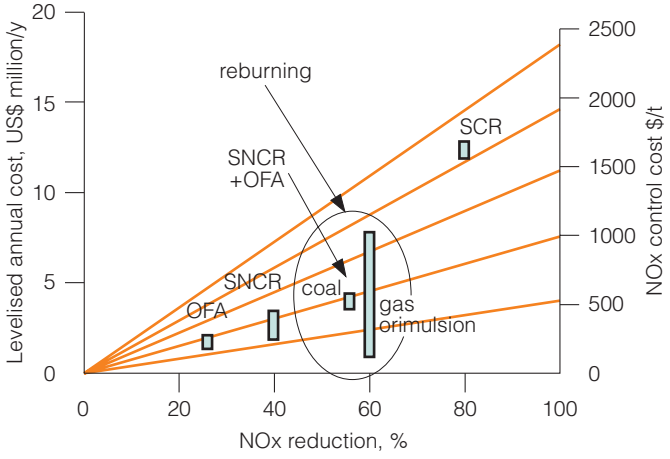


Figure 15 Cost estimates for various NOx reduction methods (Schimmoller, 1998)

The World Bank (2001) presented an economic assessment of NOx reduction technologies for coal-fired power plants in China. Table 17 shows the capital cost estimates for retrofit of a 300 MWe boiler firing bituminous coals. Combustion modifications for tangentially-fired boilers use the LNCFS series of burner arrangements. For wall-fired boilers, combustion measures include overfire air and low-NOx burners. Gas reburn, fuel-lean gas reburn, and post-combustion technologies (SNCR, SCR) are equally applicable to both boiler types. The costs are estimated with the average NOx reduction anticipated for each technology. In general, the capital costs depend on equipment requirements and, especially, the need for imported equipment.

It is seen that overfire air and fuel-lean gas reburn provide the lowest capital costs, less than 3 $/kWe. However, the NOx reductions are relatively low, generally below 40%. Technologies with NOx reductions around 50% (except for gas reburn) will generally require a capital cost of about 13–21 $/kWe. The capital costs for SNCR and SCR are estimated to be around 9 $/kWe and 51 $/kWe respectively.

For new boiler installations, the costs are generally less than those of a retrofit. The costs for gas reburn, fuel-lean gas reburn, SNCR and SCR are likely to vary between 60% and 80% of the cost of a retrofit, inversely dependent on the complexity of the system – for example, 80% for reburn and 60% for SCR. Low-NOx burners and LNCFS I cost essentially zero if new boilers are supplied with these burners. Incremental costs for overfire air in addition to low-NOx burners, or for LNCFS III instead of LNCFS I, are likely to be 50–60% of the corresponding differential costs for retrofits.

Table 18 gives operating cost estimates for these control technologies. For energy and consumables, the estimates are based on changes in plant heat rate and major uses of materials. Combustion modifications are likely to have an operating cost due to some losses in combustion efficiencies. These losses result principally from an increase in unburned carbon, and in applications with deeper NOx reductions, also from small changes in steam temperatures and excess air levels. An increase in the unburned carbon level in the fly ash may prevent the sale of the ash to the cement industry. The potential charges for changing a revenue stream to a waste disposal cost are not included in the table. The operating costs for reburn are due to the differential cost between reburn and main fuels. For SNCR and SCR, major expendables include ammonia reagent and, in the case of SCR, replacement catalyst. The operating labour impacts are likely to be minor for most combustion modifications.

Table 17 Capital cost estimates for retrofit of a 300 MWe boiler in China (World Bank, 2001)

Technology	NOx reduction, %	Capital costs, 1000 $			$/kWe
		Equipment & installation	Instrumentation & control	Total	
Overfire air	20	459	343	802	2.7
Low-NOx burners	50	2874	1176	4049	13.5
LNCFS I	35	1028	1215	2243	7.5
LNCFS II	40	3595	1215	4810	16
LNCFS III	50	5048	1215	6263	20.9
Gas reburn	55	1383	526	1909	6.4
Fuel-lean gas reburn	40	500	289	789	2.6
SNCR	35	2470	235	2704	9
SCR	75	14,994	320	15,314	51

Table 18 Operating cost estimates for retrofit of a 300 MWe boiler in China (World Bank, 2001)

Technology	Plant heat rate change, %	Major expendables	Other, $/y
Overfire air	-0.92	–	–
Low-NOx burners	-0.41	–	–
LNCFS I	0	–	–
LNCFS II	-0.92	–	–
LNCFS III	-0.92	–	–
Gas reburn	+0.17	557,000 MJ/h of natural gas	–
Fuel-lean gas reburn	+0.10	247,000 MJ/h of natural gas	–
SNCR	-0.19	697 kg/h of 29% NH_3 solution	29,000
SCR	-0.53	1229,000 $/y for replacement catalyst; 684 kg/h of 29% NH_3 solution	42,000

Table 19 Costs of retrofits for NOx reductions for cyclone boilers (O'Connor and others, 1999)

Technology	Capital cost, $/kWe	Operating cost, 1000 $/y	Levelised cost, $/t NOx removed
Water injection	0.4–1.6	340–770	80–180
Overfire air	3–8	minimal	not available
Fuel-lean gas reburn	6	fuel differential cost	not available
Gas reburn	15–35	fuel differential cost	1283
Coal reburn	40–65	not available	1072
SNCR	10–25	1050–2900	701–1075
SCR	50–120	not available	537; 1400–1700

However, maintenance labour costs will increase for post-combustion technologies due to the complexity of the system and the potential impacts on downstream equipment, especially air heaters.

The NOx reductions for cyclone boilers can be achieved by several technologies including combustion modifications and flue gas treatments (O'Connor and others, 1999). Table 19 shows the capital and operating costs of retrofits for these technologies. It should be noted that the cost estimates are based on only a few units for each technology and vary widely with site-specific factors. Nevertheless, water injection and overfire air show considerable cost savings when compared with alternative technologies. In general, reburn and SNCR have moderate costs. SCR is shown to have the highest capital cost (50–120 $/kWe) and levelised cost (up to 1700 $/t NOx removed).

4.5 Summary

The costs for NOx control technologies are again site and performance specific. They depend on degree of retrofit difficulty, unit size, uncontrolled NOx levels, required NOx reduction and a number of other factors.

Combustion modifications achieve 20 to 60% NOx reduction when applied individually. A combination of the techniques provides the potential for higher NOx reduction. In general, these systems are relatively simple to install and operate, and low in both capital and operating costs. Hence, they are the preferred option in many situations requiring moderate NOx reduction. Their costs, either in mills/kWh or $/t NOx removed, remain fairly constant through variation in the unit load factor and operating time.

Flue gas treatments, SCR and SNCR, are capable of achieving up to 90% and 50% NOx reduction respectively. Compared to combustion modifications, the technologies are generally more expensive to install and operate. They are therefore recommended for the most stringent NOx control. The costs for SCR drop significantly as operating time passes, particularly for low unit load factors. SNCR reverses the trend somehow. There is a slight increase in costs at a constant load factor, as operating time passes, probably due to inflation.

The comparative costs for various technologies are generally as follows:

- low-NOx burners can achieve 30–60% NOx reduction at relatively low costs. They are currently considered to be the most cost-effective method of achieving NOx reduction from existing coal-fired plants;
- overfire air has relatively low costs, but achieves less (20–40%) NOx reduction. It is best considered in conjunction with other technologies such as low-NOx burners and reburn;
- FGR has considerably lower costs but offers less NOx reduction (less than 20%). It is therefore used in combination with low-NOx burners, overfire air or air/fuel ratio control;
- reburn can achieve a 40–70% NOx reduction. It has moderate costs, generally higher than those for low-NOx burners and overfire air. The operating costs for reburn are mainly due to the differential cost between reburn and main fuels;
- SCR offers the greatest NOx reductions but at considerably higher capital costs than other technologies. Its operating costs are also considerably higher, mainly due to reagent consumption and catalyst replacement. The process becomes more economically competitive at longer operating periods;
- SNCR provides a 30–50% NOx reduction. Its capital costs are comparable to combustion modifications such as reburn. However, the operating costs of the process are considerably higher, mainly attributable to reagent consumption.

In many cases, hybrid control technologies provide the potential of overall high NOx reduction in a cost-effective way. In the SCR/combustion modification hybrid, the NOx concentration at the inlet of SCR is lowered. This reduces the capital requirement for catalyst, support systems and installation of SCR. This can in turn reduce the overall capital cost of the system as long as the combustion retrofit costs are moderate. The operating costs of the system are also considerably reduced, because of the minimal operating costs of the upgraded combustion system. Similarly, the SNCR/combustion modification hybrid has lower costs than the individual technologies at the same NOx reductions. In the SNCR/SCR hybrid system, the NOx levels are significantly lowered by the upstream SNCR, reducing the size of the downstream SCR. A smaller SCR reactor and elimination of reheat requirement result in considerable cost savings.

5 Combined SO_2/NOx processes

Conventional combined SO_2/NOx removal processes use a wet limestone/lime FGD system for SO_2 removal and a SCR system for NOx control. The technologies can achieve more than 90% SO_2 and 80% NOx removal from the flue gas. With the technologies, FGD and SCR work separately according to different processes. This allows each desired reduction to be set independently of the inlet SO_2/NOx concentration ratio.

In recent years, simultaneous SO_2/NOx removal processes have been developed with the view of being more cost effective than the conventional FGD/SCR processes. Most of these processes are under development or demonstration. The costs for some of the processes are discussed below.

SO_2/NOx adsorption process

The SO_2/NOx adsorption process (SNAP) is a dry, regenerable combined SO_2/NOx process developed by FLS miljø a/s (Denmark). It recovers SO_2 to elemental sulphur and converts NOx to nitrogen. The SNAP process has attracted the Danish utilities. In 1994, SK Power carried out an economic analysis, which compares the SNAP process with a combination of limestone-based FGD and SCR. The comparison was based on retrofitting a 2x200 MWe coal-fired power plant for peak load conditions, burning 1.8% sulphur coals. Table 20 summaries the results. It shows that the SNAP process is 21% lower in capital cost and 17% lower in operating cost than the combination of FGD and SCR.

In 1997, the Danish utility Vestkraft also conducted a similar economic comparison with a view to retrofitting a 90 MWe coal-fired unit at Herning, Denmark (Felsvang and others, 1999). The comparison was based on a 15-years depreciation of the equipment. The capital cost for the SNAP process was estimated to be 18.6 million US dollars which is higher than that for combination of FGD/SCR (17.9 million US dollars). However, the process has a considerably lower operating cost (98 $/h) compared with the combined system (267 $/h).

Electron beam with ammonia process

The electron beam with ammonia (EBA) process, developed by EBARA Corporation (Japan), uses ammonia injection and electron beam irradiation. It produces by-products of ammonium sulphate, ammonium nitrate and ammonium chloride. Hirano and others (1999) reported the economics of the process, for an installation at Chengdu Power Plant (China) firing a high-sulphur bituminous coal. The process was designed to treat 300,000 m^3/h of the flue gas, equivalent to 90 MWe in capacity. The capital cost for the retrofit was 72 million yuan (8 millions US dollars, 1996 value), excluding the cost for site clearance. This represented a unit capacity cost of 89 $/kWe. The total cost, including capital charges and operating cost, was 0.013 yuan/kWh (1.4 mills/kWh), 4% of the electricity sale price. The cost for a unit weight of SO_2 removed was $111 $/t SO_2 removal.

Table 20 Comparison of SNAP with combination of FGD and SCR (Boscak and others, 1998)

	Capital cost, million US$	First year cost, million US$/y	First year cost, mills/kWh
FGD	68.3	9.7	5.4
SCR	18.7	3	1.7
FGD & SCR	87	12.7	7.1
SNAP	68.9	10.6	5.9
Cost reduction,%	21	17	17

The EBA process was also evaluated for a reference application on a 300 MWe new unit in the USA firing a 2.6% sulphur bituminous coal. The process was compared with the combinations of EBA/coal reburn and wet limestone FGD/SCR. The comparison was based on the following assumptions:

- credit for NOx removal: 2000 $/t of NO_2 equivalent;
- electricity cost: 0.06 $/kWh;
- steam cost: 6.00 $/t;
- water cost: 0.25 $/t;
- ammonia cost: 150 $/t;
- EBA process by-product sale price: 100 $/t;
- gypsum sale price: 2.00 $/t;
- operating labour cost for EBA process (including coal reburn: $40,000 annually per employee (12 employees);
- annual capital charges for EBA process: 18% of capital investment;
- annual maintenance cost for EBA process: 1% of capital investment.

The EBA process is shown to have an economic advantage over the combination of wet limestone FGD/SCR. Its capital cost is estimated to be 60 million US dollars which is lower than that for the combined system (63 million US dollars). The estimated levelised cost (215 $/t SO_2 removed) is also low in comparison with 327 $/t SO_2 removed for the combined system. The economics of the EBA process is further improved when the process is combined with coal reburn. This is because the coal reburn reduces the inlet NOx concentration of the EBA system and therefore, the electron beam dosage/capacity and power consumption.

Another EBA process has been developed in Poland (Chmielewski and others, 1999). Figure 16 compares the costs of the process and other processes for a 300 MWe unit firing a 2.6% sulphur coal. The capital cost for the EBA process ranges 150–200 $/kWe, competitive to the other processes. A comparison of the SO_2 removal costs for FGD plants in Poland shows that the EBA process has an economic advantage over many others. The cost for the EBA process at Pomorzany power plant is 635 $/t SO_2 and NOx removed. This is lower than the cost for many other processes which only remove SO_2.

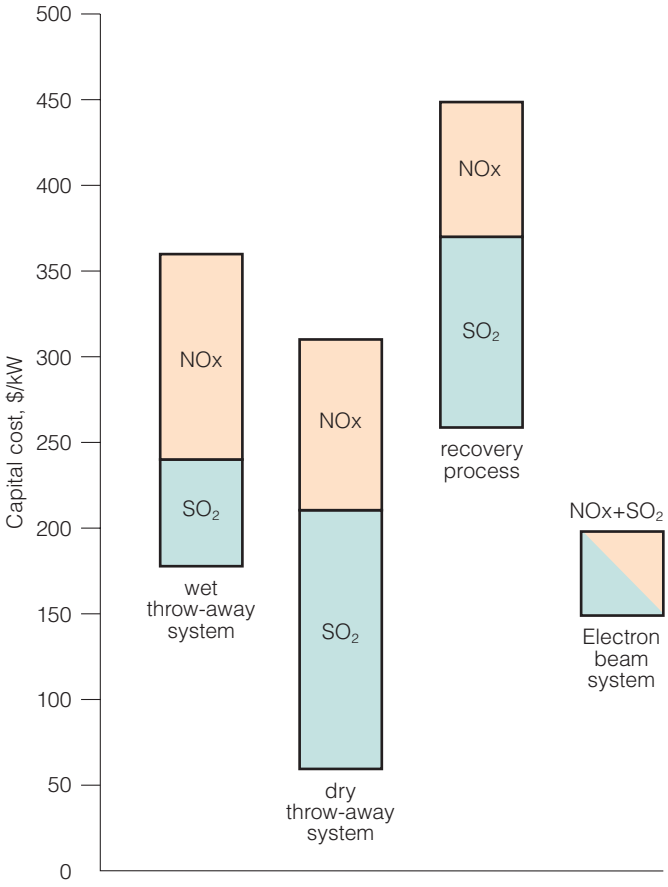


Figure 16 Cost comparison for the EBA and other processes (Chmielewski and others, 1999)

ThioClear/TecoLytic process

The wet ThioClear/TecoLytic process combines the ThioClear process for SO_2 removal with the TecoLytic process for NOx removal. The cost for the TecoLytic section of the system was reported by Breault and others (1998). The capital cost is estimated to be 30 $/kWe for a 2.5% power requirement. It is expected, however, that this cost can be reduced to 22 $/kWe for a 2% power requirement with technology improvement. The operating cost is nearly proportional to the cost of power. Figure 17 is a plot of the removal cost as a function of power cost in the range 10–50 mills/kWh. With the present technology, the operating cost is estimated to be 1100 $/t NOx removed for a power cost of 30 mills/kWh. A reduction to 800 $/t NOx removed may be possible with future technology development. This is low compared with the cost for SCR which is about 2000 $/t NOx removed.

CAT-RA process

The CAT-RA process (developed by Catra Engineering Oy in Finland) uses potassium sulphite solution as a reagent (Poutanen, 2001). It produces by-products of potassium sulphate and magnesium chloride. The capital cost for a 650 MWe unit application is estimated to be 31.2 million US dollars (1990 value). The annual operating cost is about 16.7 million US dollars (12.5 million US dollars for variable cost, 2 million US dollars for fixed cost, 2.2 million US dollars for capital charge). If the income from by-product sale is 13.1 million US dollars, the net annual operating cost will be 3.6 million US dollars. Table 21 compares the process with other competitive processes for installations with a 1.5% sulphur coal. The CAT-RA process appears to have cost advantages over the wet and semi-dry processes.

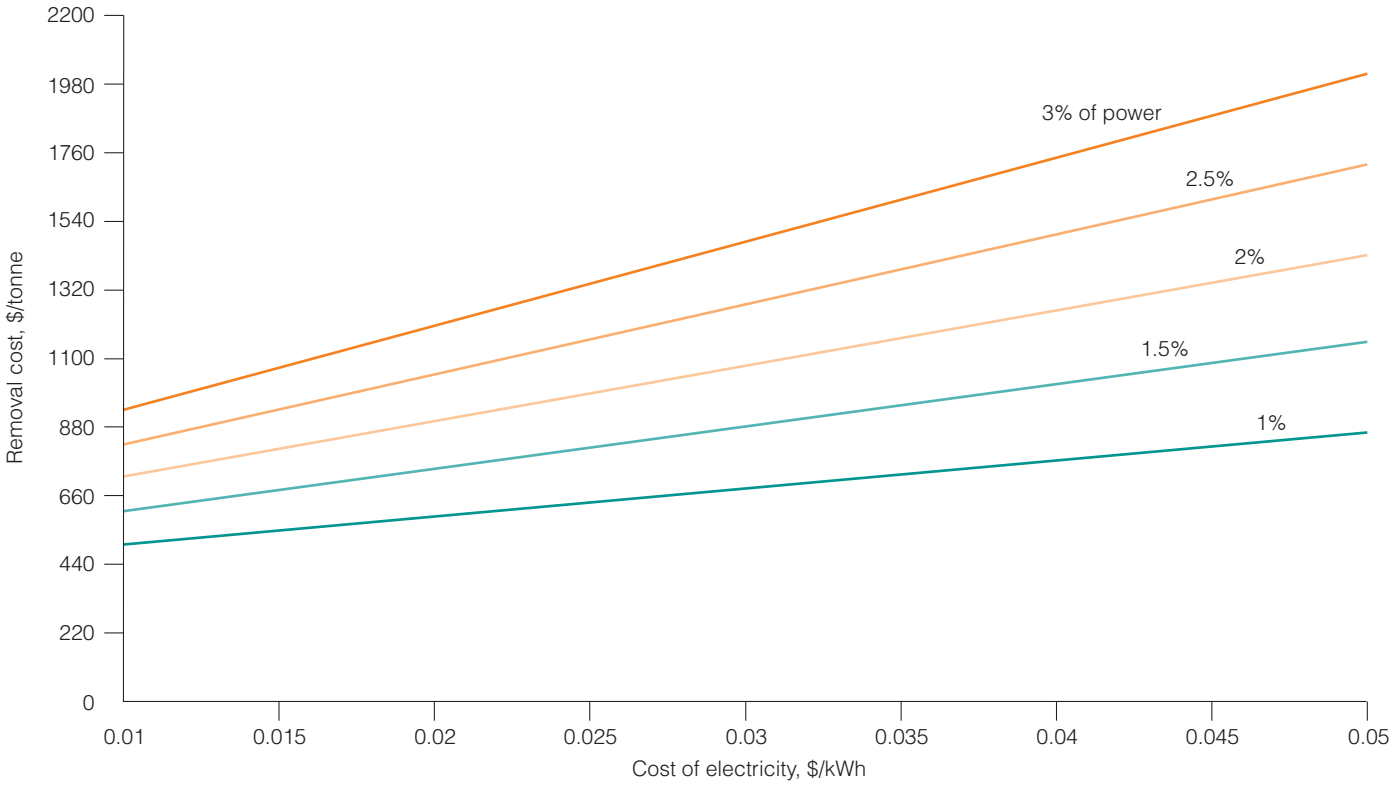


Figure 17 Removal cost as a function of power cost (Breault and others, 1998)

Table 21 Comparison of Cat-Ra with competitive processes (Poutanen, 2001)

Process	Wet	Semi-dry	Cat-Ra
Capital cost, million US$	35.1	26.2	30.3
Operating cost, million US$/y	7.6	6.1	16
By-product sale, million US$/y	–	–	13.1
Net operating cost, million US$/y	7.6	6.1	2.9
Cost of SO_2 removal, US$/MWh			
6000 h/y of operation	5	4.05	1.95
8000 h/y of operation	5	3.58	1.43

Comments

Combined SO_2/NOx removal processes are generally considered costly. However, they may be less expensive than the conventional combination of SO_2 removal by FGD and NOx removal by SCR. As most of the processes are in the development stage, their costs are expected to decrease with technological improvements.

6 Effects of air pollution control on the cost of electricity

The generating cost of electricity greatly depends on the operating cost of power generation facilities which include

- boiler, turbine and generator;
- particulate control system;
- FGD system;
- post-combustion NOx control system;
- waste water treatment system.

Figure 18 illustrates the shares of the operating cost of each system in the total power generating cost in the USA. It shows that ESP has a relatively small share (2%). FGD and SCR account for 10% and 6% respectively. The total operating cost for air pollution control systems amounts to 18% of the total generating cost. The percentages have slightly decreased compared to those found in the 1995 report by the Clean Coal Centre (Takeshita, 1995). The earlier report shows that particulate control adds to the cost of electricity by 3–4% whilst FGD and NOx control increase the cost of electricity by about 15–20%, depending on technical and economic conditions. If the generating cost is around 60 mill/kWh, Figure 18 would project an operating cost of 1.2 mills/kWh for ESP, 6 mills/kWh for FGD and 3.6 mills/kWh for SCR.

Particulate control

The cost for ESP upgrades at a number of coal-fired units in the USA was estimated by Gaikwad and Boward (1999). The upgrades included the installations of

- one additional field on two 500 MWe units to reduce the emissions from 0.03 lb/MBtu (37 mg/m^3) to 0.015 lb/MBtu (18 mg/m^3);
- one additional field on one 500 MWe unit to reduce the emissions from 0.05 lb/MBtu (62 mg/m^3) to 0.03 lb/MBtu (37 mg/m^3);

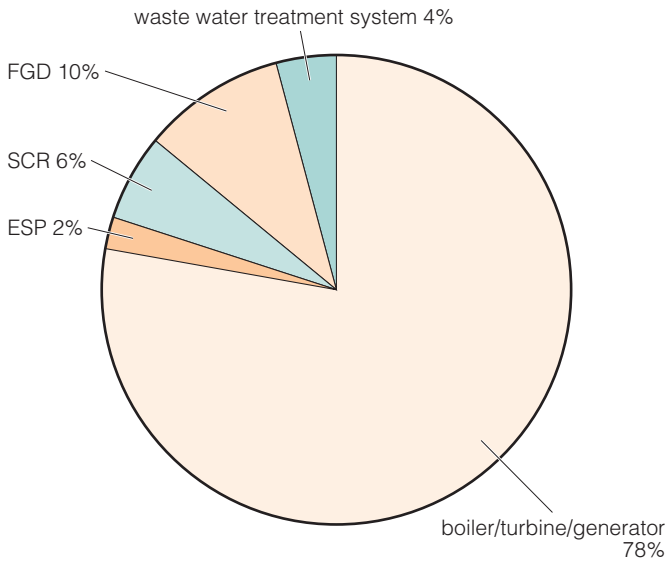


Figure 18 Shares of the operating cost of each system in the total power generating cost (Ogi and others, 1999)

- one additional field and FGC system on one 500 MWe unit and two 300 MWe units to reduce the emissions from 0.07 lb/MBtu (86 mg/m^3) to 0.03 lb/MBtu (37 mg/m^3);
- COHPAC on one 300 MWe unit and three 150 MWe units to reduce the emissions from 0.10 lb/MBtu (123 mg/m^3) to 0.03 lb/MBtu (37 mg/m^3).

The total capital cost for the upgrades is approximately 83 million dollars (equivalent to 25 $/kWe). The levelised cost is about 3980 $/t of particulate removed. The increase in the cost of electricity is estimated to be around 0.7 mills/kWh.

SO_2 control

The operating costs for wet FGD systems, on a basis of per unit electricity generated, were recently surveyed in the USA (Blythe and others, 1999). The survey was conducted for different sulphur levels of coals. The cost estimates included the following items:

- feed materials and chemicals;
- labour and supervision;
- waste disposal;
- maintenance, materials and other.

The results showed that the cost for limestone forced oxidation systems ranged from 0.695 mills/kWh to 1.436 mills/kWh, depending on the coal sulphur level. The cost for limestone natural oxidation systems varied between 0.610–1.830 mills/kWh. The magnesium-lime systems had an operating cost of 2.473 mills/kWh for high-sulphur coal applications, whilst the cost for wet lime systems with low-sulphur coals was found to be about 0.919 mills/kWh.

Gaikwad and Boward (1999) evaluated the economics of SO_2 reductions at a number of coal-fired units in the USA. There are three 500 MWe units already fitted with wet FGD systems, one 500 MWe unit firing medium sulphur coal, and two 300 MWe units firing low sulphur coal. The existing SO_2 emission levels range from 0.6 lb/MBtu (738 mg/m^3) to 2.4 lb/MBtu (2952 mg/m^3). A 50% reduction from the base levels would require the installations of dibasic acid system on three existing FGD units, wet FGD on one 500 MWe unit and one 300 MWe unit, and dry FGD on one 300 MWe unit. The total capital cost for the reductions is estimated to be around 156 million dollars (47 $/kWe). The levelised cost is about 609 $/t of SO_2 removed. The cost of electricity is increased by approximately 2.6 mills/kWh.

Table 22 shows the FGD installations at coal-fired power plants in Poland and how they affect the generating cost of electricity. The levelised costs vary between 473–1554 $/t of SO_2 removed, depending on the FGD process and plant specific conditions. The increases in the cost of electricity range from 2.9 mills/kWh to 11.5 mills/kWh (Chmielewski and others, 1999).

Table 22 Effect of FGD installations on the cost of electricity in Poland (Chmielewski and others, 1999)

Power plant, MWe	Removal cost, $/t of SO_2 removed	Increase in electricity cost, mills/kWh
Belchatów	686	4.7
Jaworzno III	670	4.5
Rybnik I (4×200)473	2.9	
Turów	564	7.3
Polaniec I (4×22)880	6.4	
Rybnik II (4×200)	488	3.0
Polaniec II (4×200)	880	6.4
Laziska II (4×200)	901	6.8
Kozienice (2×500)	925	7.7
Kozienice (5×200)	1181	8.4
Ostroleka	1103	8.4
Laziska I (2×120)	1554	11.5
Pomorzany	635	4.4

NOx control

Richardson and Hall (1999) estimated the costs of various NOx reduction technologies for per unit electricity generated. The cost estimates were conducted for a retrofit of a 500 MWe front wall-fired boiler in the UK. They were based on 10 years of plant operating period, 40% unit load factor, and required NOx emission reduction. The estimates considered both the capital costs for reduced NOx emissions and operating costs which included the following items:

- cost for lost salable ash due to either increased levels of unburned carbon (with combustion modifications) or contamination with ammonia (with flue gas treatments);
- cost for increased ash disposal to landfill;
- cost for minimising the levels of unburned carbon in fly ash;
- cost for increased auxiliary power due to additional fans (for FGR, overfire air) or upgraded fans (for SCR) or ammonia injection system (for SCR, SNCR);
- cost of anhydrous ammonia consumption with SCR and SNCR (225 $/t);
- cost of replacement catalyst with SCR (7500 $/m^3);
- cost for operating and maintenance fixed labour;
- cost of forced outages for maintenance (estimated as a percentage of lost operating time: 1.25% for SCR and 5% for SNCR).

Table 23 summarises the cost estimates for each technology. It shows that flue gas treatment-based technologies are much more expensive than combustion modifications on a basis of per unit electricity generated. In the case of SNCR retrofit, the generating cost of electricity is increased by approximately 7.61 mills/kWh. The estimates deviate considerably from those reported in the USA that are 0.15–0.54 mills/kWh for low-NOx burners, 1.30–2.40 mills/kWh for SCR, and 0.78–2.55 mills/kWh for SNCR (FGD & DeNOx, 2000b; Andover Technology Partners, 2001).

Gaikwad and Boward (1999) evaluated the economics of NOx reductions at four 500 MWe units, three 300 MWe units and three 150 MWe units. All of the units are already fitted with low-NOx burners, with NOx emission levels of 0.30–0.45 lb/MBtu (369–554 mg/m^3). A further reduction down to 0.15 lb/MBtu (185 mg/m^3) would require the installations of SCR on all 500 MWe units, combustion modifications on all 300 MWe units, and SNCR on all 150 MWe units. The total capital cost for the reductions is estimated to be around 198 million dollars (equivalent to 59 $/kWe). The levelised cost is about 1510 $/t of NOx removed. The high costs are attributed to the use of SCR and SNCR. The cost of electricity is increased by approximately 2.1 mills/kWh. If the emission level of 0.10 lb/MBtu (123 mg/m^3) is achieved, the solutions would be to install SCR on all 500 MWe units and two 300 MWe units, combustion modifications on one 300 MWe unit, and SNCR on all 150 MWe units. The capital and levelised costs for the compliance are estimated to be 246 million dollars (73 $/kWe) and 1568 $/t of NOx removed respectively. The increase in the cost of electricity is estimated to be about 2.6 mills/kWh.

Comments

The cost estimates presented above vary considerably. They might have been based on different assumptions, for example, the cost components included. The costs also depend on the technical and financial factors such as operating hours, capacity factor, labour cost and interest rate, which vary from country to country and from plant to plant. It is difficult to interpret cost data from different plants and to express them in common terms. The cost for a specific plant should be evaluated under site conditions.

Table 23 Costs of NOx reduction technologies for per unit electricity generated

Technology	NOx reduction, %	Capital cost, $/kWe	Cost of generation, mills/kWh
Low-NOx burners	40	9	0.32
Overfire air	20	8	0.66
Coal reburn	50	15	1.23
SNCR	40	13	7.61
SCR	80	100	6.99
SNCR/SCR hybrid	50	45	6.44

7 Conclusion

The costs for air pollution control systems are highly site and performance specific. They depend on the technical and financial factors such as degree of retrofit difficulty, unit size, capacity factor, removal efficiency required, labour cost and interest rate, which vary considerably from country to country and from plant to plant. The cost for a specific power plant should be evaluated under site conditions. As a general guide, particulate control can add to the cost of electricity by 2% whilst FGD and SCR may increase the cost of electricity by about 10% and 6% respectively.

There have been major reductions in the capital costs of air pollution control systems. This is especially the case with FGD where the cost in the USA is reported to have decreased by 75% since the first installation in the early 1970s. The operating and maintenance costs of air pollution control equipment have also generally decreased. The decline in costs is due mainly to continued advances in technology and increased competition in the market.

Bag filters are generally more expensive than ESPs for collection efficiencies up to 99.5%. However, the units become more cost effective for higher collection efficiencies. The choice of technology depends on the coal type and emission level. For medium to high sulphur coals with low fly ash resistivity, ESP upgrades may provide the lowest cost to meet the emission levels of 12–123 mg/m^3. For low sulphur coals with high fly ash resistivity, COHPAC and bag filter options offer alternatives to ESP upgrades for the emission level of 12 mg/m^3. ESP upgrades with SO_3 conditioning provide significant cost advantages for low sulphur coals.

The capital costs of different FGD processes vary according to the degree of complexity of the process, the amount of engineering required and other factors. Generally, processes in which the sorbent is processed to produce a saleable by-product or the sorbent is regenerated have higher capital costs than other processes. However, processes that produce a saleable by-product have lower net operating costs than those that incur by-product disposal costs. The comparative costs for various processes are generally as follows:

- wet limestone process has a relatively high capital cost, for example, this can be twice the costs for the magnesium and sodium processes. However, it has a low operating cost mainly because limestone is relatively cheap;
- in comparison with the wet limestone process, the ammonia process has high capital costs but low operating costs. Its economics improve with increasing coal sulphur level. The capital and operating costs for the seawater process are lower than the costs for the wet limestone process;
- compared with wet processes, spray dry, sorbent injection and CFB processes represent a lower capital cost but a higher variable operating cost due to the higher and more expensive sorbent use;
- regenerable processes generally have a high capital cost and power consumption. The net variable operating cost is moderate because the processes produce saleable by-products. However, the fixed operating cost is rather high.

Combustion modifications achieve 20 to 60% NOx reduction when applied individually. A combination of the techniques provides the potential for higher NOx reduction. In general, these systems are relatively simple to install and operate, and low in both capital and operating costs. Hence, they are the preferred option in many situations requiring moderate NOx reduction. Flue gas treatments, SCR and SNCR, are capable of achieving up to 90% and 50% NOx reduction respectively. Compared to combustion modifications, the technologies are generally more expensive to install and operate. They are therefore recommended for the most stringent NOx control. The comparative costs for various technologies are generally as follows:

- low-NOx burners can achieve 30–60% NOx reduction at relatively low costs. They are currently considered to be the most cost-effective method of achieving NOx reduction from existing coal-fired plants;
- overfire air has relatively low costs, but achieves less (20–40%) NOx reduction. It is best considered in conjunction with other technologies such as low-NOx burners and reburn;
- FGR has considerably lower costs but offers less NOx reduction (less than 20%). Hence it is used in combination with low-NOx burners, overfire air or air/fuel ratio control;
- reburn can achieve a 40–70% NOx reduction. It has moderate costs, generally higher than those for low-NOx burners and overfire air. The operating costs for reburn are mainly due to the differential cost between reburn and main fuels;
- SCR offers the greatest NOx reductions but at considerably higher capital costs than other technologies. Its operating costs are also considerably higher, mainly due to reagent consumption and catalyst replacement. The process becomes more economically competitive at longer operating periods;
- SNCR provides a 30–50% NOx reduction. Its capital costs are comparable to combustion modifications such as reburn. However, the operating costs of the process are considerably higher, mainly attributable to reagent consumption;
- in many cases, hybrid control technologies (SCR or SNCR/combustion modification, SNCR/SCR) provide the potential of overall high NOx reduction. They are generally more cost-effective than stand-alone technology.

Combined SO_2/NOx removal processes are generally considered costly. However, they may be less expensive than the conventional combination of SO_2 removal by FGD and NOx removal by SCR. As most the processes are in the development stage, their costs are epected to decrease with technological improvements.

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