AIR-BLOWN VERSUS OXYGEN-BLOWN GASIFICATION

Report No. COAL R089

by

M J Fisher

CLEAN COAL
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GROUP

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SUMMARY

The Air Blown Gasification Cycle (ABGC) uses an air-blown gasifier which the Clean Coal Power Generation Group (CCPGG) claims provides the most economic coal-fired power generation cycle. The economics have been evaluated in previous studies, by Bechtel for example. However, air separation technology and costs are known to have changed as a result of air separation units (ASU) becoming more tailored to oxygen-blown gasification combined cycle (GCC) applications.

An audit carried out by the Electric Power Research Institute (EPRI) in December 1993 found there to be 17 commercial coal gasification plants worldwide, all using oxygen rather than air. However, the use of oxygen in these plants is mainly because the fuel-gas is used as a feedstock for chemical production rather than power generation. For chemical feedstock production, oxygen blowing is clearly desirable because the resulting fuel-gas has a high concentration of hydrogen and carbon monoxide, ie it provides a high 'cold gas' conversion efficiency. For power generation, however, 'cold gas' efficiency is largely irrelevant, since sensible, as well as potential energy, can be converted into electricity via the steam and gas turbine cycles. The concentration of fuel-gas combustibles need only be high enough to sustain combustion in the turbine.

Resolving the question of which is best, air-blown or oxygen-blown gasification for combined cycle power generation, is made complex by the wide variety of gasifier configurations available, with their generic variations in dry and slurry coal feeding methods, plus the unresolved debate regarding whether to clean the fuel-gas using relatively novel, but more efficient, hot (dry) fuel-gas cleaning processes, or to stay with long proven, cold (wet) fuel-gas cleaning methods. Another factor, specific to oxygen-blown gasification, is whether to integrate the ASU with the gas turbine, making an integrated gasification combined cycle (IGCC) process.

With so many factors influencing the final decision, it is possible to compose a set of ground rules which bias the result. This report is based upon previous studies and publicly available literature. Collectively, the public domain studies illustrate a positive trend favouring air-blown over oxygen-blown gasification, but the studies need to be normalised with respect to the more influential design features before the true effect of air-blown versus oxygen-blown gasification can be seen.

By finding sets of data which differ by only one of the factors needing to be isolated it is possible to compose a set of typical indices that can be applied to the various gasifier configurations and bring them all to a common basis. The basis chosen for this report is

dry coal feeding, hot fuel-gas cleaning and, in the case of oxygen-blown gasifiers, a non-integrated ASU.

After applying the indices and achieving a common basis, the cycle efficiencies of all the air-blown gasifiers come into reasonably close agreement at 46.3% (±2 percentage points) net electricity output (lower heating value basis), while the cycle efficiencies of the oxygen-blown gasifiers are also in reasonable agreement at 44.5% (+2.1, -1.7 percentage points). Thus, on the chosen basis, air-blown gasifiers offer typically 1.8 percentage point advantage over oxygen-blown gasifiers.

Having established what efficiency advantage is likely when air-blown and oxygen-blown GCCs are compared on a common basis, the indices developed for data normalisation enable relative cycle efficiencies to be predicted for dissimilar air-blown and oxygen-blown GCC arrangements. Thus when comparing an air-blown, dry coal feed, gasifier using hot fuel-gas clean-up, to an oxygen-blown gasifier receiving a dry coal feed but conventionally equipped with cold fuel-gas clean-up, the advantage to air blowing is found to be some 2.6 percentage points.

Data comparison generally indicates a 1.1 percentage point cycle efficiency improvement to oxygen-blown gasifiers that have an integrated ASU, ie IGCC processes. Even this improvement, however, still leaves air-blown gasification with a clear efficiency advantage of 0.7 percentage points on the common basis or 1.5 percentage points if the oxygen-blown IGCC has cold gas cleaning.

Considering different levels of hot gas clean-up sophistication, the simplest arrangement relies upon in-bed feeding of limestone for desulphurisation with a downstream ceramic filter for particulate clean-up. This simple arrangement saves around 0,5 percentage points of cycle efficiency over more advanced methods. Consequently, the cycle efficiency advantage of an ABGC with simple hot gas clean-up, compared to an oxygen-blown GCC with cold gas cleaning, could be up to 3.1 percentage points.

Environmental emissions, even from advanced hot fuel-gas cleaning, are not as good as can be achieved from commercially proven cold fuel-gas cleaning. Adopting cold gas cleaning with air-blown gasification imposes a greater cycle efficiency deficit than it does with oxygen blowing because of the greater heat losses associated with cooling the larger fuel-gas volume. Air-blown gasification might lose 1.8 percentage points while oxygen blowing might lose only 0.8 percentage points.

Environmental emissions from the ABGC, even if equipped with cold gas cleaning of its fuel-gas, would not match those from oxygen-blown gasifiers. This is because stack emissions contributed by its associated circulating fluidised bed combustion (CFBC) char combustor, although as clean as the best CFBCs, are higher than are achievable by GCC unless the CFBC is equipped with expensive flue gas clean-up. In addition, the sulphide bed ash, resulting from limestone fed to the gasifier, needs to be fully oxidised to calcium sulphate in order to be environmentally acceptable. This has been successfully demonstrated with a range of sorbents, but there remains a preference for unleachable slag, as comes from ash slagging, entrained flow, gasifiers.

Installation cost estimates for the various air-blown and oxygen-blown GCC technologies are subject to large uncertainties. Cost estimates vary widely, though entrained flow oxygen-blown gasifiers have consistently the highest estimates while air-blown fluidised beds have the lowest. Taking all technologies into account, the bias noticeably favours air-blown gasification.

A 1990 EPRI study comparing air-blown and (non-integrated ASU) oxygen-blown fluidised bed gasifiers is arguably the most appropriate indication of the relative cost of the two technologies. This study was carried out in considerable detail, though unintentionally penalises the oxygen-blown case by omitting a carbon burn-up cell, thus reducing its cycle efficiency. If the oxygen-blown case is compensated, the additional net power output lowers the plant's specific cost to US$1,160 per kW_e. This compares to US$1,042 per kW_e (10% reduction) for the air-blown case. Costs are saved in both cases by basing upon an existing electric utility site and facility sharing.

Integrating the ASU is expected to reduce the cost of an oxygen-blown installation by 2-3%.

Electricity generation costs are found to favour air-blown gasification, as would be expected from the higher cycle efficiency and lower plant costs. European and American studies predict electricity costs around 4.6USc per kWh for air-blown gasifiers and 5.8USc per kWh for oxygen-blown gasifiers. An Indian study estimates consistently higher prices, due to the efficiency loss attendant on using high ash coal. Nonetheless, air-blown gasifiers have the lowest electricity costs, with the air-blown fluidised bed being lowest of all, at 5.6USc per kWh.

CONTENTS

AIR-BLOWN VERSUS OXYGEN-BLOWN GASIFICATION

INTRODUCTION

The process of gasification has been applied since the beginning of the iron age, when wood was partially gasified into charcoal which was then used in the manufacture of iron[1]. By the 17th century the UK was suffering a depletion of its forests and prohibitions were placed on the making of charcoal. This led to coke (formed by the partial gasification of coal) being used as a substitute.

Credit for the first commercial demonstration of coal gasification is given to a Scot, William Murdoch, at the end of the 18th Century[2]. His invention was applied in London during the early 19th century, when coal-derived 'town-gas' was first distributed to houses and factories. Since the gasification process was essentially conducted in the absence of air, merely distilling off the coal volatiles, the fixed carbon portion of the original coal, ie coke, was left behind as a by-product. Such partial gasification of coal remained in use within the UK until the mid-1960s, when firstly the gasification of light oil, and subsequently the availability of natural gas (methane) from the North Sea, took the place of coal-derived town-gas.

Air- and steam-blown gasification of the coke that remains as a by-product of town-gas manufacture was carried out in the early gas industry as a means of supplementing gas supplies during peak demand periods. The coke was gasified in cyclically-operated 'water gas' plants, which used a combination of air-blown exothermic heating (a 'producer gas' reaction), followed by endothermic steam blowing to make water gas, mainly H_2 and CO, with some residual N_2 and $C0_2$. The Haber process, making ammonia for fertiliser production, originally relied upon a similar mixture of air- and steam-blown gases to provide its constituent hydrogen and nitrogen.

A similar low calorific value fuel-gas to that from cyclically-operated water gas plants is generated in air-blown coal gasifiers such as are now under development for gasification combined cycle (GCC) electrical power generating plants. The Air Blown Gasification Cycle (ABGC) is one example of an air-blown GCC process[3].

A variety of GCC processes are currently under development, utilising three generic forms of gasifier, ie the entrained flow, the fluidised bed and the moving bed (sometimes termed the fixed bed) design. The three types are shown in Figure 1. These generic forms encompass a large number of proprietary configurations, illustrating that process developers are far from unanimous of what process parameters are optimum. Of the variations that exist, possibly the most fundamental design aspect on which there is no unanimity is whether to make the gasification process air-blown or oxygen-blown.

The use of oxygen with steam, as an alternative gasification medium to air with steam, evolved as a result of the commercial availability of pure oxygen during the 1940s. By excluding the nitrogen diluent it became possible to combine the endothermic gasification and exothermic partial combustion reactions into a concurrent arrangement, turning the previous cyclical producer and water gas processes into a continuous one.

Dr Karl von Linde is usually credited with starting the air separation industry when he formed the German company Linde Eismachinen AG in 1879[4]. George Claude, who founded the French company l'Air Liquide in 1902, refined the Linde process by adding an expansion engine to produce refrigeration. This addition substantially reduced the energy required to liquefy air. By 1907, the Linde Air Products Company, a division of Union Carbide Corporation, acquired the rights to the Linde process.

The oxygen market, and hence the air separation industry, grew to supply the needs of oxy-acetylene gas welding applications. For that use, the supply of oxygen in small, heavy, gas cylinders is still standard practice. However, further market growth demanded a more efficient distribution method and this was achieved in 1932 with the first bulk delivery of oxygen in liquefied form. From that time, the steel, chemical and petrochemical industries boosted the need for oxygen to a degree whereby, in the early 1940s, large liquid oxygen production facilities were in operation, exporting oxygen via vacuum-insulated rail cars. The inevitable progression for large user industries was the installation of dedicated oxygen making facilities, as was occurring by the end of the 1940s.

Conservative industry estimates suggest that the existing world capacity for oxygen production is now some 280,000 tonne per day from 7,000-10,000 plants[5]. Individual plant sizes have increased from 200 tonne per day in the 1960s to currently 3,000 tonne per day, with a single stream 5,000 tonne per day plant having been conceptually studied by 1'Air Liquide[6]. This supplier equipped the Sasol coal gasification process in 1976 with 14 plants producing a combined 30,000 tonne per day of oxygen. Commercial oxygen plants are available from eight major suppliers who generally compete throughout the world[5].

Oxygen or Air?

In considering the relative merits of air-blown versus oxygen-blown gasifiers, it is easy to be misled that the improvement in off-gas product quality which inspired the development of oxygen-blown gasification processes from the 1940s, automatically means that oxygen-blown gasification is a technically better solution. However, before making such a judgement, it is necessary to consider the process improvements inspired by the availability of bulk oxygen and consider whether those benefits apply in the case of GCC.

The early oxygen-blown gasifier developments originated either from the desire for continuous complete gasification of both coal and petroleum feedstocks into town-gas or the production of chemical by-products. The chemicals include liquid hydrocarbons, as required by Sasol in the Fischer Tropsch process for making motor fuels, for methanation reactions producing substitute natural gas, for methanol production as a fuel in its own right which can also be used as a feedstock for making other chemicals including petrol (via the Mobil - MTG process), and finally as a source of hydrogen in fertiliser manufacture[2].

Early gasifiers intended for the complete gasification of coal originated in Germany. During the late 1920s, Winkler and Lurgi were both working on the gasification of lignite, a fuel that could not be gasified by conventional coking methods using externally heated retorts. It was Lurgi who, in 1930, made a major step forward by showing how oxygen injection into their gasifier raised the temperature of combustion and, hence, the rate of gasification. By 1936 Lurgi had built a gasifier that operated under pressure, another factor which helped increase its throughput. Although the Lurgi pressurised gasifier was intended primarily for town-gas manufacture, operation under pressure enabled its easier integration with downstream chemical processing stages. This advantage assisted in the application of Lurgi gasifiers to the production of motor fuels by Sasol in 1955.

The Winkler process, a low pressure process during its early developments, operates on the fluidised bed principle, using coal particles approximately 2mm in size, injected into the gasifier and suspended in a flow of oxygen and steam. The Lurgi process uses a thick moving bed of coal, fed from the top and gasified by oxygen and steam as it moves downwards under gravity. The third generic form of gasifier, using entrained flow of finely ground coal, is also exemplified by an early German process. The Koppers-Totzek, which was further developed by the Koppers Company of the USA around 1950, uses pulverised coal, and is blown by oxygen and steam into a gasifier which operates above the ash slagging temperature. Such a design enables any type of coal to be used and also reduces the formation of tars and phenols in the off-gas.

British involvement with the Lurgi concept during the 1950s extended the Lurgi process application to bituminous coals, thus making it comparable with the capabilities of the Koppers-Totzek process. By the late 1950s these two designs emerged as the basis for large capacity coal gasifiers throughout the world. The Winkler fluid-bed process has since been developed for use in pressurised GCC applications, but remains primarily suited to lignite.

Further development of the Lurgi gasifier has continued in the UK since 1974, with the conversion of a conventional dry ash gasifier to slagging operation at the British Gas site at Westfield, Scotland. This technology is known as the British Gas/Lurgi gasifier (BGL).

Early developments into the oxygen gasification of petroleum feedstocks for fuel-gas and chemical by-products were undertaken by the American Texaco company and by Dutch Shell. Both of these companies opted for entrained flow gasification.

Research at Texaco began in the late 1940s at Texaco's Montebello laboratories, California[7]. It was first commercialised in the early 1950s as a means to convert inexpensive (at that time) natural gas into by-products for the manufacture of intermediate chemicals, ammonia etc[8]. As the relative cost of fossil fuels changed over the years, Texaco examined a wider range of feedstocks. By the mid-1950s, Texaco had experience of gasifying a range of liquid hydrocarbons, including heavy oils, asphalts and tars, and had also carried out initial developments on coal. In the period 1956-58, coal was gasified in a 100 tonne per day Texaco gasifier at the Olin Mathieson Chemical Plant, Morgantown, West Virginia. The coal was prepared as a fine ground, water-based slurry, and gasified using the same entrained flow techniques as had previously been developed for oils.

Although over 100 commercial Texaco plants have been licensed worldwide, gasifying a variety of hydrocarbon feedstocks, only eight of these have been coal-based plants. The latter were inspired by the oil price increase in the early 1970s which saw the installation of pilot plants at the Montebello laboratories (15 tonne per day), at Ruhrchemie and Ruhrkohle at Oberhausen, Germany (165 tonne per day), and at an existing Tennessee Valley Authority (TVA) ammonia plant at Muscle Shoals, Alabama (190 tonne per day). Only the Cool Water plant at Daggett, South California (operated 1984-89) was a GCC plant, gasifying 1,000 tonne per day of coal and generating 120MW_e[9]. The remaining coal-based plants produce ammonia or intermediate chemicals. A second GCC-based (250MW_e) Texaco coal gasification plant (1,900 tonne per day) is presently being commissioned for Tampa Electric in South-western Polk County, Florida[5].

Shell's experience in gasification dates back to the early 1950s, when the first Shell Gasification Process (SGP) was commissioned with oil as the feedstock. To date there are more than 140 SGP units licensed worldwide[6]. They convert a variety of hydrocarbon liquids and gases to carbon monoxide and hydrogen to produce high value chemicals, such as ammonia and methanol[7].

Shell Internationale Petroleum Maatschappij B V began work on coal gasification in 1972 with a 6 tonne per day development plant at the Shell Amsterdam Laboratory, followed by a 150 tonne per day pilot plant at Deutsche Shell Harburg refinery in 1978.

From 1974 until mid-1981, Krupp-Koppers (previously Heinrich Koppers) with their own entrained flow coal gasification technology, co-operated with Shell in the development of coal gasification at the Hamburg plant. The partnership was dissolved in June 1981, with Krupp-Koppers subsequently developing their own pressurised entrained-flow gasification cycle, known as PRENFLO, using a 150 tonne per day pilot plant at Fuerstenhausen, Germany. This technology is currently being applied at the 300MW_e integrated gasification combined cycle (IGCC) demonstration plant at Puertollano, central Spain[10].

Further development of Shell 'know-how' has been progressed by Shell's affiliate in the USA, the Shell Oil Company. Shell built a demonstration unit at its oil refinery and chemical complex in Deer Park, near Houston, Texas. This plant, which started operation in 1987, is known as SCGP-1, and gasifies some 250 tonne per day of bituminous coal[11]. Shell technology is currently being demonstrated as an IGCC at the commercial scale by Demkolec BV at Buggenum, near Roermond, in the Netherlands. This plant is designed to generate 250MW_e from a single Shell gasifier, processing 2,000 tonne per day of coal[12].

The preceding text provides ample evidence that those companies with experience in gasification dating back to the 1940s, when oxygen became commercially available, chose to use an oxygen and steam gasifying medium in order to continuously gasify their chosen feedstock while producing medium calorific value town-gas, substitute natural gas or a mixture of hydrogen and carbon monoxide suited to the production of chemicals and synthetic oil. The fact that such designs continue to be based upon oxygen in the development of coal-based GCC makes commercial sense, in that it makes maximum use of prior experience, but must not be taken to imply that oxygen based GCC is necessarily the best approach. There are arguments and counter-arguments.

Those designers in favour of oxygen-blown gasifiers[13] cite such factors as:

- The gasifier and all equipment intermediate to the gas turbine combustor must be at least double the volume in the case of air-blown gasification, since the gas consists of at least 50% by volume nitrogen. This must increase plant costs and decrease the potential maximum throughput of a single gasifier stream.

- The conversion efficiency of energy within the coal into chemical energy within the fuel-gas is less in the case of air-blown gasification, since more combustion of the coal takes place to compensate for the sensible heat in the nitrogen. This equates to **a** lower 'cold gas' efficiency.

- Since gasifier systems have a typical pressure drop around 3bar, there is a saving in gas compression energy by not having to compress the nitrogen to the inlet pressure of the gasifier. In the case of oxygen-blown gasification, the nitrogen is either vented at near ambient pressure from the outlet of the air separation unit (ASU) or is only compressed to the lower pressure of the gas turbine's combustor.

Designers who favour air-blown gasification[14] point out that:

- the capital cost, operating complexity and parasitic power consumption of an ASU are all substantial liabilities to be set against oxygen-blown GCC designs;

- separating air into oxygen and nitrogen, only to mix them back together again at the gas turbine, as occurs with nitrogen integrated ASU schemes (Section 4.1), is a fundamental thermodynamic inefficiency.

1.2 Cold Gas Efficiency and Heat Bypass Ratio

It is important to differentiate between the conversion efficiency of primary fuel energy into electrical power, as is the goal of a GCC plant, and 'cold gas' efficiency, which is the percentage of primary fuel energy appearing as potential energy within the emergent fuel-gas.

In the case of a GCC plant, fuel conversion efficiency is a measure of the effectiveness with which the potential energy within the primary fuel is converted into electricity, generated by the gas and steam turbines. The result of this conversion defines the cycle efficiency or 'heat rate' of the process. Obtaining a high efficiency requires near 100% conversion of the raw fuel into potential or sensible energy and the subsequent minimisation of heat losses such as occur in the downstream gas cleaning process, in the steam turbine condenser and in the final stack gases. Astute optimisation is necessary to minimise these losses, but the reward is a high conversion of fuel energy into electricity.

In papers extolling the virtues of specific GCC processes, there is frequently mention of the term 'cold gas' efficiency[15]. This is particularly the case where oxygen-blown designs are favoured and is a reminder of the heritage of oxygen-blown gasifiers as producers of chemical by-products rather than electricity.

When a gasifier is used solely for gas production, particularly where the resulting fuel-gas may have to be transmitted a significant distance to the point of use, the efficiency of the process can only be measured by the potential heat content (sometimes termed the chemical energy) of the product gas, relative to the potential energy of the raw feedstock. It is this which is termed the cold gas efficiency.

Comparison of Cold Gas Efficiencies

To a large extent, debates and comparisons of cold gas efficiency are of only academic importance in the context of a GCC plant, since the efficiency of the process is determined by the total sensible as well as potential energy delivered to the power generating equipment. Nonetheless, Table 1 illustrates that the cold gas efficiency of air-blown gasification can compete with oxygen-blown gasification when an appropriate design of gasifier is chosen, ie not a single-stage entrained flow gasifier nor one with a wet coal feed.

The cold gas efficiency performance of various gasifier types, air-blown and oxygen-blown, have been predicted using CRE Group Ltd's ARACHNE (acronym for 'a reactor and cycle handling network evaluator') flowsheet modelling package. The comparison was made using a consistent set of design assumptions, all fed with a UK bituminous coal (Daw Mill). The gasifiers were assumed to be uncooled refractory lined vessels except for one of the dry feed single stage gasifier cases where a water/steam cooled vessel was used, to represent the Shell/Krupp-Koppers PRENFLO design.

It can be seen that air blowing the wet feed single-stage (Texaco type) gasifier gives a very low cold gas efficiency (55% vs 75%), but the efficiency improves when air blowing a wet feed two-stage (Destec type) and a dry feed single-stage (Shell type) gasifier (69-73% vs 81%; 69-73% vs 84%). The cold gas efficiency for a dry feed two-stage air-blown entrained gasifier is higher still, particularly with a high degree of air preheat, and is similar to the best commercially available oxygen-blown gasifiers (83%). This type of air-blown gasifier is similar to the two-stage gasifier being developed in Japan by Mitsubishi Heavy Industries Ltd (MHI).

The highest cold gas efficiency comes from a dry feed, two-stage entrained flow, oxygen-blown gasifier, such as is being developed by Hitachi. However, Hitachi intend their design for synthesis gas production rather than the basis for a GCC process. This is implied by the name, 'HYCOL', meaning Hydrogen from Coal[7].

The cold gas efficiency of the ABGC gasifier is lower than that of any of the entrained gasifiers, with the exception of the air-blown wet feed single-stage type (59% vs 55%). However, if cold gas efficiency is re-calculated, excluding the potential heat of the primary fuel which remains as ungasified char, the cold gas efficiency is similar (78%) to that of the best air-blown entrained gasifier (80%) and only slightly lower than that of the best oxygen-blown gasifier intended for GCC applications (84%).

Heat Bypass Ratio

The air-blown gasifier, because it confers sensible heat to nitrogen, inevitably converts a greater proportion of the original feedstock energy into sensible heat, with less appearing as potential fuel-gas energy. This lowers the cold gas efficiency and increases the ratio of electrical energy coming from the steam turbine relative to the gas turbine. Some texts refer to this as an increase in the heat bypass ratio[16]. A higher heat bypass ratio is a potential disadvantage because of the lower efficiency of the steam turbine cycle (typically 45%) compared to the gas turbine cycle (typically 55%).

In a natural gas fired combined cycle, the gas turbine to steam turbine power ratio is around 2:1[17]. With GCC systems this ratio falls, as heat generated in the raw-gas coolers and during gas clean-up bypasses the gas turbine and enters the steam cycle.

The gas turbine to steam turbine power ratio is different for different GCC systems, the highest ratio being obtained with the moving bed oxygen-blown gasifier, as demonstrated by the oxygen-blown British Gas/Lurgi design[18], where the fuel-gas leaves the gasifier at a relatively low temperature, so minimising heat bypass during fuel-gas cooling. Here, the gas:steam turbine power ratio might be 1.8:1. The oxygen-blown two-stage entrained flow design, where the otherwise high fuel-gas exit temperature is moderated by a second stage injection of coal, exemplified by the Destec (formerly Dow Chemicals Company) design[19,20], has a similar, though slightly lower, gas turbine:steam turbine power ratio, around 1.6:1. The single-stage entrained flow design, such as that by Texaco and Shell, convert a greater proportion of feedstock heat into steam and so have ratios varying between 1.1:1 and 1.3:1, depending upon whether they use wet (Texaco) or dry (Shell) coal feeding and whether gas clean-up is carried out using a cold, wet washing process or is carried out dry, at high temperature, around 500°C.

The ABGC has a lower gas turbine:steam turbine power ratio than any of the above, partly due to the sensible energy in the fuel-gas nitrogen, which increases the amount of energy recovered as steam, and partly because the carbon conversion within the gasifier is 70-80%, requiring a separate circulating fluidised bed combustor (CFBC) to complete the utilisation of the residual carbon char. All the heat released within the CFBC bypasses to the steam cycle. Off-setting these factors are the low (1,000°C) fuel-gas exit temperature from the gasifier and the use of hot gas clean-up. Even so, the gas turbine:steam turbine power ratio is low, around 0.8:1.

Thus, the ABGC would be expected to show a lower power generating efficiency compared to oxygen-blown designs, simply because it generates a greater proportion of total power from the lower efficiency steam turbine cycle. Comparative studies do not, in fact, bear this out, as will be demonstrated in the following section of this report. The ABGC derives advantages which more than compensate for the less advantageous gas:steam turbine power ratio. These advantages are derived from the routes described below.

- By relaxing on the need for the highest possible carbon conversion, less steam injection is required into the gasifier, which confers the advantage of reduced final stack gas losses from latent heat of water vapour in the flue gas.

- Provided only that enough fuel-gas can be produced from the coal to ensure correct operating conditions for the gas turbine, the presence of the CFBC means that responsibility for efficient carbon conversion does not rest with the gasifier alone. Thus, operation of the gasifier can be optimised to the needs of the gas turbine, leaving the CFBC char combustor to ensure high total carbon utilisation. This is not to say that oxygen-blown designs do not achieve the same overall carbon conversion, but the ABGC, via its separate CFBC char combustor which operates at benign low temperature (1,000°C) oxidising conditions, enables the use of a high efficiency steam cycle, rather than the steam cycle being compromised by gasifier conditions.

- The use of partial desulphurisation within the fluidised bed of the air-blown ABGC, together with high temperature dry gas downstream clean-up, minimises efficiency losses from the gas clean-up processes. Also, the sulphur is recovered in a low energy state as calcium sulphate, $CaSO_4$ instead of in a high energy state as elemental sulphur.

- The air-blown gasifier avoids the high parasitic power consumption of the oxygen producing ASU, leaving more of the gross power generated from the GCC to be exported (net power production), hence improving the net cycle efficiency.

The ultimate judgement on whether an air-blown or oxygen-blown GCC is the better design, requires the appreciation of many complex issues, each of which have their advantages and disadvantages. Also, because so many gasifier configurations are available, some more suited to one blowing medium than the other, it is possible to compose a set of ground rules for a study of this kind which, intentionally or otherwise, unfairly prejudice the result. Therefore, a conscious decision was taken during this report's preparation to only cite information already in the public domain, ie no new study was commissioned, and to avoid previous studies which, by virtue of finding directly in favour of the ABGC, might be thought to hold an in-built bias. In consequence, the UK Department of Trade and Industry (DTI) report[20a], which compared the ABGC with the BGL (oxygen-blown) gasifier and found in favour of the ABGC to the extent of a 2.4 percentage point efficiency gain and a 7% reduction in electricity generating costs, has not been used to develop the air-blown versus oxygen-blown comparison. Instead, the current report relies upon truly independent studies covering a range of air-blown and oxygen-blown gasifier configurations. The studies were carried out by the IEA Coal Research Ltd (IEACR) in the UK, Central Research Institute of Electric Power Industry (CRIEPI) in Japan, Electric Power Research Institute (EPRI) in the USA RWE Energie AG in Germany, Council of Scientific and Industrial Research (CSIR) in India and the University of Ulster in the UK.

GCC EFFICIENCY AND COST COMPARISONS

Comparative Studies Conducted by IEA Coal Research

IEACR, in their report IEACR/55, consider a matrix of GCC design variations and compile tables comparing plant efficiency, capital charges and electricity costs[21]. The study endeavours to compare not only different GCC systems but also pulverised fuel (pf), CFBC and pressurised fluidised bed combustion (PFBC). In so doing, it illustrates the difficulties that arise when attempting to make like-with-like comparisons. For example:

- By restricting the environmental performance of all plants to that achievable by flue gas desulphurisation (FGD) with pf-firing, the cold gas washing process, as frequently used in oxygen-blown GCC design studies, imposes an efficiency penalty without any benefit for its potentially much higher sulphur removal efficiency. (Note, however, that in this study cold gas sulphur removal is by the Purisol process, credited with achieving only 90% sulphur removal). Since fluidised bed gasifiers can equal the sulphur control of pf with FGD merely by feeding limestone direct to the gasifier, these designs have inherent advantages of lower capital cost and higher energy efficiency compared to designs equipped with full downstream fuel-gas clean-up equipment. This is especially true when compared to gasifiers equipped with cold (wet) gas clean-up processes, as are conventionally applied to oxygen-blown gasifiers.

- A similar situation occurs with NOx control. Since the IEACR study limits the environmental control to that possible with pf and low-NOx burners, there is no requirement for GCCs designed around hot gas cleaning to incorporate downstream NH_3 and HCN removal in order to match the lower NO_x values possible from cold, wet wash, fuel-gas clean-up.

- Since the range of commercially available gas turbines is limited, the study is sensibly based around commercially available (or near available) designs: the Class I GE Frame 7E, the Class II Siemens V94 and the Class III GE Frame 9F. Manufacturer's parameters for these three turbines were fixed throughout the study (ie the firing and exhaust temperatures of the power turbine, and the air mass flow, inlet and delivery temperatures and pressure ratio of the compressor turbine). Even with these turbine parameters fixed, the performance of such machines is variable; they can accommodate variations in fuel/flue gas flows through the power turbine with consequent variations in generated power. When such variations are compounded with variations in steam cycle power which occur as a result of the differing amounts of sensible heat recovery required by the various GCC designs (Section 1.2), the total electrical power output from the combined cycle plants vary considerably between design concepts that were intended to be directly comparable. IEACR attempt to resolve this issue by devising and applying a cost scaling factor to bring all plants to a 300 MWe output, but concede this can only be theoretical since it implies (non-existent) intermediate sized gas turbines.

Bearing the above factors in mind, the IEACR report ranked the various GCC schemes as shown in Table 2.

Cases 12, 13 and 14 which are the closest representation of the ABGC (especially Cases 13 and 14 with hot gas cleaning) are referred to in the IEACR report as hybrid cases, since the design utilises a separate CFBC char combustor and does not aim to maximise gasifier carbon conversion. It seeks, instead, for a carbon conversion which enables the steam and gas turbine cycles to be mutually optimised. Since this design is still under development, the evaluations were credited with the efficiency benefit of an advanced GE Frame 9F gas turbine, since such machines were considered likely to be commercially proven and available by the time this GCC design is at the demonstration phase.

The use of advanced steam conditions in Case 14 yielded the highest cycle efficiency values of all the cases studied, being 48.8% on the coal's net (lower) heating value (LHV). IEACR admit, however, that such an arrangement is relatively immature and represents the greatest technical risk. The supercritical steam conditions of 240bar, 565/565°C (from 125bar, 538/538°C in Cases 12 and 13), are possible only because of the superheating/reheating potential conferred by the CFBC char combustor. Such high steam temperature supercritical conditions are not anticipated for other GCC designs, where the turbine exhaust gas temperature limits the cycle temperature to 540/540°C.

Cases 15 and 16 are fluidised bed gasifier systems which seek to maximise carbon conversion in the gasifier. The study anticipates that complete conversion is difficult to achieve in a single stage fluidised bed and anticipates that 5% of the carbon will combust within a carbon burn-up cell, the duty of which is to also oxidise calcium sulphide (CaS) from the gasifier bed into environmentally acceptable $CaSO_4$. These cases are representative of the fluid-bed designs by:

- Kellogg-Rust-Westinghouse (KRW), as demonstrated at the 24 tonne per day Waltz Mill test facility, Pennsylvania and now to be incorporated in the Pinon Pine plant at Tracy Power Station, Reno, Nevada[22].

- the U-Gas process, licensed by Tampella from the Institute of Gas Technology, IGT, Chicago (intended, at one time, for demonstration at Tom's Creek, Virginia[23]);

- the High Temperature Winkler (HTW) process, a pressurised version of the original ambient pressure design, developed by Rheinbraun and Uhde since the mid-1970s (the KoBra plant was to be the first full scale use of the HTW in a GCC[24]).

The KRW and HTW designs have been developed to use both air and oxygen blowing, while the U-Gas is intended to be air-blown. Pinon Pine and (if built) Toms Creek and KoBra are all intended to be air-blown installations because of a small, but significant, cycle efficiency advantage, as confirmed by IEACR in the comparison of Cases 15 and 16.

Case 17 is a generic evaluation of a moving bed, oxygen-blown, slagging gasifier, such as developed by BGL[18], though the first large scale GCC plant at Lünen, Germany in 1971 was air-blown and retained the earlier non-slagging Lurgi arrangement. All such gasifiers were primarily developed for production of pipeline gas, with emphasis on cold gas efficiency (see Section 1.2, above). Much of the generated steam is used within the gasification process and does not become available for power production via the steam cycle. Since this GCC design relies heavily on the gas turbine cycle, application of more advanced turbines would benefit this design more than other GCCs.

Cases 18 to 24 are all entrained flow, ash slagging, oxygen-blown gasifier based, combined cycles. Such designs are less suited to air blowing (though air blowing can be used, especially in dry coal feed, two-stage designs). The IEACR report presumes that all entrained gasifiers would be oxygen-blown.

Since entrained flow gasifiers have generally developed from systems conceived to process oil it is not surprising that both the Texaco and Dow gasifiers are designed to receive coal as a slurry (referred to as a wet coal feed in Table 2). The consequent steam provides a gasifying medium, but its quantity reduces operational flexibility compared to controlled steam injection with dry feeding, ie as in the Shell and Krupp-Koppers PRENFLO gasifiers. Thus, dry coal feeding in Case 19 (with hot gas cleaning) shows an advantage over wet coal feeding in Case 20; similarly, Case 18 shows an advantage over Case 22.

Comparing Case 19, an oxygen-blown, dry coal feed, hot gas cleaning, entrained gasifier, with Case 16, which uses similar operating parameters but is based upon the fluidised bed principle, Case 19 has the slightly lower efficiency (0.8 percentage points) because it requires more oxygen to maintain the higher entrained flow gasification temperature, ie it requires more combustion. Consequently, Case 19 produces a lower calorific value gas with increased sensible energy that has to be taken up by the less efficient steam cycle.

Case 25 is an entrained flow gasifier working in conjunction with an ASU which is integrated with the gas turbine, forming an IGCC plant. In this report, IGCC will be used to reference oxygen-blown GCCs where all, or part, of the air for the ASU is taken from the turbine's compressor.

Comparing Case 25 to the similar, but non-integrated ASU case, Case 23, cycle efficiency improves by 1 percentage point and auxiliary power consumption reduces by 28% due to there no longer being an ASU compressor. The IEACR study conjectures that, if a similar cycle efficiency improvement occurred with the fluidised bed Cases 15 and 16, the small advantage in favour of air blowing might become a similar advantage in favour of an oxygen-blown design.

IEACR Comparative Plant Capital Costs

IEACR claim to have developed their costs (mid-1991 US dollars; US$) by making use of those demonstration GCC plants either already built or being built, together with component equipment pricing taken from other user industries. In the case of the hybrid GCC systems, such as the ABGC, their relative novelty and the fact that no complete coal-based system has been built means that cost figures for such systems are less reliable.

A common basis is claimed for the cost figures but IEACR emphasise that absolute cost levels have an accuracy of only ±30%. However, IEACR point out that absolute values are less important than the cost relationship between the various systems, which are subject to less uncertainty because of the common basis on which costs are estimated.

Table 3 reproduces the IEACR costing information in an abbreviated form. The air-blown ABGC cases, Cases 12 and 13, show significant cost advantages compared to all other schemes, though they are assisted in this comparison by being based upon the advanced GE9F gas turbine.

IEACR term Cases 12 and 13 (which represent the ABGC) hybrid cycles, and suggest they are sufficiently novel as to make their costings less certain. Comparing Case 23 with Case 22 suggests that the use of the Class III GE9F gas turbine may be giving these cases a specific cost advantage of 8.8%. Case 14, the supercritical steam cycle, has not been costed as it was felt that this added further technical uncertainty to an already undemonstrated concept.

Cases 15 and 16 allow comparison of fluidised bed systems, such as that by KRW, when air-blown and oxygen-blown. Elimination of the ASU reduces costs for the air-blown case but savings are substantially offset by increased costs in other plant components due to an increase in size of equipment to handle the inflated fuel-gas flow. The specific cost of the oxygen-blown fluidised bed, Case 16, is significantly less than any of the entrained flow (oxygen-blown) cases.

Case 17 was disadvantaged relative to the entrained gasifier designs in areas of acid gas treatment and waste heat recovery. However, substantial savings in other system components resulted in a final specific cost similar to the entrained gasifier base case, Case 22.

Cases 18. 19, 20 and 22 allow comparison of costs for wet and dry coal feeding and hot and cold fuel-gas clean-up. Specific costs are similar, though with a slight increase when using dry coal feeding.

Cases 21. 22 and 23 show how the effect of plant scale (power output) can have a significant effect on specific costs. This is addressed in the subsequent comparison table, Table 4, where plant electrical outputs for all cases are brought to a common value and costs are adjusted appropriately using an IEACR developed scaling factor. Case 21 is omitted in Table 4 since its electrical output is so much below $300MW_e$ that scaling to $300MW_e$ is more sensibly achieved using Cases 22 and 23.

Case 24 shows a 5.4% specific cost increase over the similar Case 23, due to inclusion of a supercritical pressure (conventional 540°C/540°C temperature) steam cycle.

Case 25. the integrated ASU IGCC, has no cost data presented as IEACR were unable to obtain costing data for such an integrated ASU from equipment manufacturers.

IEACR noted, when carrying out the cost comparison shown in Table 3, that the higher electrical outputs of certain equipment combinations, specifically when moving from Class I through to Class III gas turbines and incorporating higher efficiency steam cycles, caused specific plant costs (US$ per kW_e) to reduce, though absolute plant costs increased. Thus, in order to eliminate the effect of changes in plant electrical output, IEACR developed a scaling factor to enable all costs to be brought to values appropriate to a common electrical output, namely $300MW_e$. It was recognised that this could only be a theoretical exercise, as practice would require intermediate sizes of gas turbine. However, it enabled all GCCs to be compared strictly on their technology merits. Results are summarised in Table 4.

When costs are normalised to a $300MW_e$ output, the specific costs for all four fluidised bed cases, Cases 12, 13, 15 and 16 are similar, averaging US$1,391(±39) per kW_e, with the lowest cost being Case 15 (an air-blown KRW type gasifier) and the highest cost being Case 16 (an oxygen-blown KRW type gasifier). The ABGC fluidised bed with hot gas cleaning, Case 13, is second lowest in specific cost, US$1,380 per kW_e, and has the highest cycle efficiency, 47.6% (LHV).

The remaining GCC technologies (excluding Case 24 with the supercritical steam conditions), all based upon oxygen blowing, have an average specific cost of US$1,827(±32)per kW_e.

IEACR Comparative Costs of Electricity

IEACR developed estimates for the cost of electricity for five of their GCC cases, Case 13 (as the ABGC with hot gas clean-up), Case 15 (a KRW type fluidised bed air-blown gasifier with hot gas cleaning), Case 17 (a Lurgi type moving bed, oxygen-blown, gasifier with cold gas cleaning), Case 18 (a Shell type entrained flow, oxygen-blown gasifier, dry coal feed, cold gas cleaning) and Case 22 (a Texaco type entrained flow, oxygen-blown gasifier, slurry coal feed and cold gas cleaning).

Fuel costs are based upon a typical coal price of US$40 per tonne (US$1.61 per GJ). For a plant of η% net cycle efficiency and a fuel of US$1.61 per GJ higher heating value (HHV), the contribution to cost of electricity is $58/\eta$USc per kWh.

The capital charges contribution to the cost of electricity is calculated from the basic capital cost plus additional allowances, ie pre-paid royalties, spare parts inventory, organisation and start-up costs, working capital and interest during construction. This gives the lifetime capital investment. An annual recovery rate (US$ million) is then derived from the expected plant life (presumed 25 years) and a suitable interest rate for discounted cash flow (DCF-the standard rate used in IEACR member countries is 7.5%). The annual recovery rate is then divided by the rated plant output, assuming an annual average load factor of 65%, to obtain the capital charge contribution in USC per kWh.

Non-fuel operating costs include fixed costs for labour to operate and maintain the plant, for overheads (30% of labour costs) and maintenance materials (2% of basic capital investment per year). Variable costs include limestone (as required) at US$19.6 per tonne delivered, sulphuric acid (as required) at US$134 per tonne delivered, ash disposal (presumed as a negligible cost, but recognised as potentially significant if certain wastes, ie fluidised bed ash (sulphided limestone and coal ash, termed LASH), were to come under hazardous waste regulations, water, ie for steam cycle make-up, at

US$0.72 m^{-3}, catalysts (as required) taken as 0.2 to 0.35% of the annual capital charge, liquid waste disposal at US$21 per tonne, by-products (as manufactured by the different GCCs) credited at US$100 per tonne for sulphur, US$145 per tonne for ammonia and US$214 per tonne for phenols. Gypsum is credited at a negligible value, since market saturation is anticipated.

A tabulated comparison of electricity costs is given in Table 5. The ABGC type fluidised bed shows the electricity cost advantage that would be expected for a plant that previous tables have shown to have one of the lowest specific costs and the highest cycle efficiency.

IEACR comment that the hybrid type GCC (as the ABGC) has an electrical cost advantage which should prove encouraging for development of a suitable demonstration plant. However, it also acknowledges the relative immaturity of the technology involved, making it a higher risk option than alternative GCCs.

Japanese Study of Gasification Combined Cycle Optimisation

A Japanese study of GCC system options has been carried for the Japanese government departments of Ministry of International Trade and Industry (MITI) and New Energy and Industrial Development Organisation (NEDO). Results have been reported by the Central Research Institute of Electric Power Industry (CRIEPI) and Mitsubishi Heavy Industries Ltd (MHI)[25].

Initial GCC development work in Japan commenced in 1974 by the Japanese Coal Mining Research Centre and the Electric Power Development Company, with tests being carried out using a 5 tonne per day facility at Yuubari City, Hokkaido. Subsequently, from 1980-1990, a 40 tonne per day two-stage fluidised bed gasifier equipped with hot gas clean-up was operated at the same site. Meanwhile, CRIEPI and MHI commenced test work during 1982 using a 2 tonne per day entrained bed gasifier. This latter development was sufficiently successful that a decision was taken to adopt the entrained bed gasifier for a 200 tonne per day pilot plant. This latter plant was completed in February 1991 at Iwaki City, Fukushima Prefecture.

The results from the early testwork were used to enable a comparative study of the options:

- dry or wet coal feeding;
- air-blown or oxygen-blown gasification;
- dry- (hot-) or wet- (cold-) gas clean-up.

These six options were considered relative to the operation of both single-stage and two-stage entrained flow gasifier operation.

The design combination which offered the highest net thermal efficiency was the air-blown, two-stage, arrangement using hot, dry (460°C)[26] gas cleaning. Based upon a gas turbine inlet temperature of 1,300°C, the values in Table 6 are predicted for a $200MW_e$ plant. These results have some direct relevance to the ABGC design.

A two-stage gasifier operates by producing high temperature ash slagging conditions within its primary chamber (minimum 1,600°C in the case of the Japanese design) but then attemperates the resulting fuel-gases by injecting further coal into them within the second stage gasifier. Endothermic gasification takes place, resulting in a final temperature around 1,000°C, ie not dissimilar to the fuel-gas temperature from a (inherently single-stage) fluidised bed gasifier.

Where the Japanese design differs from a single-stage fluidised bed is that the high temperature, short residence time, gasification reactions are not conducive to sulphur capture by injection of limestone. Therefore, all sulphur capture must take place downstream, a disadvantage compared to the fluidised bed.

The two-stage Japanese arrangement does, however, have two advantages over an ABGC:

i. ungasified char, escaping the gasifier but captured downstream, is recycled back to the high temperature slagging zone where it is totally consumed - there is no requirement for a carbon burn-up cell (as in the KRW fluid-bed gasifier) or separate CFBC (as in the ABGC design);

ii. all the coal ash is produced as a benign glassy slag, instead of the LASH material emanating from a limestone fed fluid-bed.

Putting aside the relative differences between the two-stage entrained flow gasifier and the single-stage fluidised bed, the Japanese find a cycle efficiency advantage from using air, rather than oxygen, as the blowing medium. They also favour a dry coal feed and hot gas clean-up for the highest efficiency, all factors with which champions of the ABGC would concur.

The Japanese authors also describe a future $379MW_e$ gross, $349.2MW_e$ net electrical output GCC based upon the same preferred combination of features as deduced from Table 6. That plant is expected to show a net thermal efficiency of 46% (LHV). They also postulate that for a (European) lower ash fusion temperature coal (1,450°C instead of 1,600°C) and a lower steam cycle condenser pressure (some $0.03bar_a$ instead of $0.05bar_a$, the latter as in the calculations leading to Table 6 and as used by IEACR in the calculations leading to Table 2), the future plant would attain a net efficiency of 48% (LHV). Further performance improvement is anticipated from the use of gas turbines with firing temperatures of 1,400 - 1,500°C and advanced hot gas clean-up, combining simultaneous ammonia and sulphur removal.

It is of interest from Table 6 that the Japanese did not consider an oxygen-blown two-stage gasifier or a single-stage air-blown gasifier. The omission of these arrangements is probably because the Japanese recognised the relative incongruities within such schemes. The use of oxygen blowing with a dry coal feed generates considerable heat, making for difficult control of 1,000°C at the exit of the second stage of a two-stage gasifier (or directly at the exit of a single-stage low temperature gasifier design, such as a fluidised bed). Similarly, single-stage air blowing a slurry coal feed requires considerable air preheat to enable sufficient combustion for attainment of the ash slagging temperature, while still retaining sufficient cold gas efficiency to enable operation of the gas turbine cycle.

Design ingenuity makes any combination possible if so desired, ie the KRW and HTW fluidised bed gasifiers, operating around 1,000°C with a dry coal feed, have both been developed for use with oxygen as well as air, as has the two-stage entrained flow, dry coal feed, HYCOL process. Similarly, the slurry fed, entrained flow, single-stage Texaco gasifier, operating above 1,400°C, has been considered in air-blown studies[27] as well as in pilot plant trials[28]. Nonetheless, it is clear that certain combinations of oxidant, temperature and coal feed conditions are thermodynamically preferable.

CRIEPI Study - Plant and Electricity Production Costs

This aspect appears not to have been considered in detail, the emphasis being more upon optimising cycle efficiency, thus reducing fuel usage (almost all energy resources in Japan are imported from abroad) and reducing the environmental impact of power generation.

The intention of the Japanese GCC development is to produce electricity at an equivalent cost to that from a pf coal-fired power plant, albeit at an equipment installed cost of perhaps +10%.

2.3 Comparison of Air-blown and Oxygen-blown Versions of the Krupp-Koppers PRENFLO Gasifier

The Electric Power Research Institute (EPRI) sponsored a study to compare air-blown and oxygen-blown versions of the Krupp-Koppers PRENFLO gasifier in near $500MW_e$ gross output arrangements[29]. The Krupp-Koppers PRENFLO gasifier is currently being applied (oxygen-blown) in the $300MW_e$ fully integrated IGCC facility at Puertollano, Spain[10].

Both the air-blown and oxygen-blown concepts within this study are IGCCs, ie they feature full air integration with the gas turbine; all of the oxidant for the gasifier, whether directly via a booster compressor (air-blown design) or indirectly via the ASU and subsequent boosters, originates at the gas turbine compressor discharge.

Since the Krupp-Koppers PRENFLO design is a single-stage, dry coal feed, entrained flow gasifier, operating above the ash slagging temperature, it requires considerable air preheat to be air-blown. Thus, the study requires the air, leaving the gas turbine at 440°C, to be pressure boosted using a special high temperature (though commercially available) compressor, sufficiently to overcome the resistance of the gasifier and downstream gas clean-up. During compression, the air acquires further preheat to 550°C.

With the oxygen-blown arrangement, pressurised air from the gas turbine is first cooled and then passed through the ASU. The resulting oxygen is further pressurised to the requirement of the gasifier, while the nitrogen is pressurised to the requirement of the turbine combustion chamber.

Since the oxygen and nitrogen from the ASU are relatively cool, the study found that the parasitic compressor power requirement for the high temperature air compressor in the air-blown case (23.2MW$_e$ out of 525.2MW$_e$ gross, ie 4.42%) exceeded the parasitic power consumption of the ASU (20.7MW$_e$ out of 495.3MW$_e$ gross, ie 4.18%). This contradicts what is usually anticipated for air-blown versus oxygen-blown comparisons. The cause, in this instance, is the need for inefficient high temperature air compression in order to accommodate air blowing within an inappropriate single-stage slagging gasifier technology. The oxygen-blown version benefits from power savings commensurate with full ASU integration with the gas turbine. Thus, this comparison of parasitic power consumption is extreme in terms of favouring oxygen blowing and penalising air blowing.

The rest of the fuel-gas treatment process in the study is identical for both blowing mediums, consisting of initial gas cooling by clean gas recycle, heat recovery boilers, a cyclone and ceramic filter for dedusting (dust is recycled back to the gasifier) and multi-wet wash processes for halide, nitrogen compounds and sulphur removal.

The steam cycle operates at 133bar and 537°C/537°C for both concepts.

Overall, the air-blown gasifier produces the greater net power output, 479.7MW$_e$ against 457.3MW$_e$. The extra power derives mainly from the steam turbine due to the greater amount of sensible heat to be removed from the fuel-gas.

Despite the fact that a single-stage entrained flow gasifier is inappropriate for air blowing, the difference in cycle efficiency was surprisingly small. The net plant cycle efficiency was found to be 46.7% (LHV) when air-blown and 46.9% (LHV) when oxygen-blown. The study acknowledges that these values are commendably close but criticises the air-blown concept as unlikely to be suited to high ash, high moisture fuels such as lignites and biomass. This is because air preheat would need to be even higher than 550°C in order to achieve a satisfactory thermal balance.

Such factors merely reinforce the clear fact that air blowing is not the preferred solution for a slagging, single-stage, gasifier. Credit therefore, that, when feasible (ie, with a low ash slagging temperature, high quality coal) the air-blown cycle efficiency compares closely with the oxygen-blown result.

Krupp-Koppers PRENFLO Gasifier Study - Plant and Electricity Production Costs

The study concludes that the cycle efficiency predictions of air-blown and oxygen-blown gasifiers is so close that a final assessment is only possible after a thorough economic evaluation. This would establish whether the marginal efficiency advantage that oxygen blowing appears to confer in the Krupp-Koppers PRENFLO design is offset by any differential in capital costs between incorporating an ASU for oxygen blowing and the larger vessel sizes required when air blowing. No subsequent paper providing costing comparison data was found.

Near economic equivalence was found in a comparable 1980 EPRI study of oxygen and air-blown Texaco entrained flow gasifiers[27]. Although the cost of the oxidant feed system was reduced by a factor of five by omitting the ASU in the air-blown case, this decrease was offset by the increased costs for the gasification, raw-gas cooling and acid gas removal systems, mostly due to the higher raw-gas volume when air blowing.

2.4 Comparison of Air-blown and Oxygen-blown High Temperature Winkler Gasifiers

The High Temperature Winkler (HTW) fluidised bed gasifier, adopted by Rheinbraun in close co-operation with Uhde, their engineering partner, is a development of the early atmospheric pressure, lignite gasifier developments started in the 1920s.

Between 1955 and 1964 Rheinbraun converted lignite (brown coal) into synthesis gas using atmospheric pressure Winkler generators, each with a 10 tonne per hour capacity. Development into a high pressure process began in the mid-1970s. In 1978 a pilot plant was commissioned at Rheinbraun's Wachtberg factory, operating at l0bar and with a 34 tonne per day coal throughput. Since that time, two industrial scale plants have been installed generating synthesis gas for methanol and ammonia production[24].

In October 1989, Rheinbraun commissioned a 160 tonne per day advanced pilot plant, operating at 25bar, which had facility for using either oxygen/steam or air/steam as the gasifying agent. This plant was specifically intended to enable the HTW to be used in conjunction with a gas turbine. On oxygen, the HTW is claimed to have achieved gasification rates equivalent to $100MW_{th}m^{-3}$, and $50MW_{th}m^{-3}$ on air. All development work has concentrated on continued use of high moisture lignite (50-60% moisture). For GCC concepts, the lignite is first dried using a steam fluidised bed drier, to bring the moisture content down to 12%.

From the early 1990s, Rheinbraun AG, Uhde GmbH and Lurgi AG have been co-operating towards commercialising the HTW GCC concept. With this in mind, a $300MW_e$ demonstration plant concept was studied for RWE Energie AG. The plant has the German acronym KoBra. Four concepts were originally considered; one or two gasifier streams providing the total gasification duty, both either oxygen- blown or air-blown[30].

The fluidised bed gasifier operates at 900-950°C, well below ash slagging temperature but high enough to maximise carbon conversion. With the high volatile content lignite, Rheinbraun claim 80-85% conversion to fuel-gas potential energy (cold gas efficiency). Some unconverted char remains to be separately burned. In the case of the KoBra concept this was intended to be carried out in existing, adjacent, CFBC boilers.

In many respects the KoBra plant has much in common with an ABGC, except that the intention of the designers, assisted by the reactive fuel, is to maximise carbon conversion. With the ABGC there is significant residual carbon, but this is advantageous in optimising the steam cycle using a purpose designed CFBC. The KoBra steam cycle is based upon 110bar, 520°C, with reheat at 29bar, 520°C. Condenser pressure is $0.05bar_a$.

Gas dedusting is by dry ceramic filter elements but removal of trace elements and desulphurisation is intended to be via a wet process. The lack of limestone feeding to the fluid-bed is at variance with the ABGC design. However, some in-bed sulphur capture is achieved, since the Rheinish lignite contains a high calcium content in its ash.

The gas turbine chosen was a Siemens V94.3 with a 1,120°C inlet temperature.

Rheinbraun found that the relative merits of air-blown gasification and oxygen-blown gasification required close study, but eventually (1992) opted for air blowing. Relative net cycle efficiencies found in the early comparative study were 45.5% (LHV) with air blowing and 45.0% (LHV) with oxygen blowing.

HTW KoBra Gasifier Study - Plant and Electricity Production Costs

Neither reference to the HTW/KoBra technology[24,30] makes any specific comment upon plant capital and electricity production costs. Emission pollution reduction seems to have been the main driving force, though it was acknowledged[24] that 'the next generation of brown coal-based power plants, which will ultimately replace the present installations, will have to retain their profitability in an open European market'.

<u>Comparison of Air-blown and Oxygen-blown KRW Fluidised Bed Gasifiers</u>

The US Department of Energy (DOE) sponsored a study jointly with Southern Company Services Inc (SCS), which compared various air-blown KRW fluidised bed gasifier configurations. A final report was produced in December 1990[31]. This study was carried out by the Morgantown Energy Technology Centre (METC), M W Kellogg, The Tennessee Technology University Centre for Electric Power and the General Electric Company. One of the configurations examined, a $400MW_e$ air-blown GCC with hot gas clean-up, was compared to the results from an earlier EPRI/SCS study of a similar scheme involving a non-integrated ASU, oxygen blowing and cold gas clean-up[32]. This comparison of air and oxygen blowing has been separately reported[33].

Both the US DOE and EPRI studies were site specific to the Plant Wamsley site of Georgia Power Company, which is part of SCS, one of the USA's largest investor-owned electric utility systems.

Both studies were highly detailed and considered plant design, operability, reliability, capital costs, performance, operation and maintenance (O&M) costs, cost of electricity, environmental characteristics and technology risks. Amongst various objectives, they aimed to determine if air-blown gasification with hot gas clean-up is more cost effective than oxygen-blown GCC with cold gas clean-up.

The General Electric GE MS7001F gas turbine was selected for the studies. The design fuel was Illinois No.6 bituminous coal and the nominal plant output was $400MW_e$.

The major differences between the air-blown and oxygen-blown cases are in the oxidant feed, the gas clean-up and the by-products.

In the air-blown case, air for the gasifier and for fuel-gas combustion is entirely taken from the gas turbine compressor, though the air passing to the gasifier is boosted in pressure by a separate electric motor driven compressor. In the oxygen-blown case, only the air for combusting the fuel-gas is taken from the gas turbine; auxiliary compressors are used to supply air to the ASU making oxygen for the gasifier. Nitrogen from the ASU is substantially vented to atmosphere.

The hot gas clean-up for the air-blown case consists of limestone feeding to the gasifier for bulk sulphur removal followed downstream by a ceramic filter for particulates capture plus a sulphur polishing stage, consisting of a chloride guard bed upstream of regenerative zinc ferrite. Overall sulphur capture is credited as 99.4%, but there are no useful by-products.

The partly sulphided lime and coal ash from the gasifier bed (LASH), plus fines from the ceramic filter, are drained to a separately fluidised sulphator bed where the calcium sulphide is converted to sulphate and ungasified carbon is combusted to an overall efficiency of 99.9%.

The cold gas clean-up system within the oxygen-blown case consists of fuel-gas cooling, metal filters for particulate removal, acid gas washing (Selexol), sour water stripping, sulphur recovery from the acid gas (Claus) and treatment of the sulphur recovery plant tail gas (SCOT). Overall sulphur capture is credited as 96.4% from the cold gas system. This is a lower retention than from the air-blown version, but the plant produces 78 tonne per day of saleable sulphur.

The oxygen-blown gasifier does not require a sulphator for LASH gasifier bed material. Since the sulphator also acts as a carbon burn-up cell, it follows that the oxygen-blown gasifier is disadvantaged by having a lower carbon utilisation than the air-blown arrangement, only 95.8% compared to 99.4%.

Due to the use of cold gas cleaning, the fuel-gas temperature entering the turbine combustor is nearly 280°C colder than fuel-gas from the air-blown hot gas clean-up system.

Table 7 gives comparison data from the study and shows that the air-blown gasifier is predicted to achieve a 2.3 percentage point advantage compared to the oxygen-blown version.

A thermodynamic analysis was carried out as part of the study. This showed that the oxygen-blown gasifier is of higher thermodynamic efficiency than the air-blown design, with the air-blown deficit being attributable to heating up nitrogen to gasifier temperature and the fact that more combustion takes place. However, the advantage to the oxygen-blown gasifier is partially offset in the gas turbine system where a greater air flow has to be compressed and, more importantly, heated in the combustion chamber.

As regards the relative thermodynamic efficiency of hot gas (air-blown) and cold gas (oxygen-blown) clean-up systems, the hot gas case was estimated to have an efficiency 0.4 to 0.8 percentage points higher than the cold gas case. However, the efficiency advantage would have been greater were it not for the inclusion of the external (to the gasifier) zinc ferrite sulphur polishing stage. By reference to another case within the same study (Case 6[31]) the authors estimate that the advantage to an air-blown gasifier using only in-bed limestone additions for desulphurisation (as in the case of the simplest ABGC concept), compared to cold gas washing an oxygen-blown fuel-gas, could be 2.6 to 3.0 percentage points.

In another comparison, but this time of the relative thermodynamic efficiencies of oxygen blowing versus air blowing when both plants are equipped with hot gas cleaning, the advantage was thought likely to favour oxygen blowing, by an estimated 0.2 to 0.8 percentage points.

Since the thermodynamic advantage of hot gas cleaning (including the external separate zinc ferrite stage) essentially cancels out the similar advantage of oxygen blowing, the study examined what major loss accounted for the 2.3 percentage points (5.8% of thermal input) cycle efficiency advantage to air blowing. Not surprisingly, this was found to be attributable to the higher unconverted carbon losses from the oxygen-blown gasifier when operating without the air-blown gasifier's carbon burn-up sulphator cell. A secondary factor was a more efficient heat recovery steam generator (HRSG) at the turbine exit on the air-blown cycle, due to better matched temperature profiles.

This study concludes that a carbon burn-up cell must be included, whatever blowing medium is chosen. It also concludes that, since there is an apparent thermodynamic advantage favouring both oxygen-blown gasification and hot gas clean-up, this combination, when optimised with respect to its HRSG, should be more efficient than either of the cases examined.

The ultimate conclusion favouring oxygen blowing and hot gas clean-up (in conjunction with a carbon burn-up cell) pre-supposes that the thermal advantage associated with hot gas clean-up and an air-blown gasifier would be eliminated by changing from cold to hot gas clean-up in the oxygen-blown mode. The IEACR study would question this speculative assumption. IEACR Cases 15 and 16 (Table 2) show a residual efficiency advantage to an air-blown KRW gasifier (44.3% vs 43.7%) when air-blown and oxygen-blown gasifiers are compared, both with hot gas clean-up.

KRW Gasifier Study - Comparative Plant Costs

Detailed cost break-downs are available for both the KRW air-blown gasifier and the oxygen-blown version, site specific to the SCS Wansley site. Table 8 compares the total capital requirements of the air-blown and oxygen-blown KRW gasifiers on the basis of mid-1990US$.

The oxygen-blown study was carried out prior to the air-blown study (Dec 1987 versus mid-1990) and so, when the later US DOE funded study was carried out[31] the oxygen-blown design was updated in various ways in order to make the studies directly comparable.

These included:

- Modifications to the gasifier process arrangement due to a significant revision of the GE MS7001F gas turbine. This resulted in an increase in the coal consumption (up from 3,459 tonne per day to 3,845 tonne per day, +11.1%), with concomitant changes to the gas turbine power (up from 254.9MW_e to 298.8MW_e) and to steam cycle power (up from 153.2MW_e to 159.4MW_e).

- Allowance for 10.3% inflation, directly applicable to the combined-cycle power generation equipment (up from US$134.15 million to US$147.97 million).

The change to the coal feed rate was converted into a cost inflation index by ratioing the old to the new coal feed rate and allowing a scaling exponent of 0.65 (equivalent to +7.1%). The 10.3% inflation cost index and the 7.1% coal increase cost index were combined (1.103x1.071=18.15%) and used to increase the costs of the ASU, the gasifier, acid gas removal, sulphur recovery tail gas treatment, sour water stripping and waste water treatment.

Further cost amendments were incorporated, including:

- Substantially revising the costs of the coal reception, handling and preparation (up from US$5.90 million to US$16.88 million) and the fines and ash handling system (up from US$1.87 million to US$3.19 million). In both instances these changes are made in recognition of more definitive costings carried out in the later study.

- Making minor changes to the general plant facilities, some entailing cost increases and some cost savings.

The specific capital cost values (US$ per kW_e) in Table 8 are composed of the same cost elements as the IEACR study. Both sets of costs exclude interest charged on capital during construction (the US DOE study refers to this as the AFUDC - 'allowance for funds used during construction') and as such are 'overnight construction' costs. Thus, the specific cost figures in Table 4 (Case 15) and Table 8 (Air-blown) should be comparable, apart from the IEACR values being based upon mid-1991 rather than mid-1990 US$, being normalised to 300MW_e rather than 493.8MW_e, and not including the cost of downstream fuel-gas H_2S polishing.

If the IEACR scaling factor is applied to bring Case 15 up to a 493.8MW_e output, and a -3% factor allowed to compensate for cost differences between mid-1991 and mid-1990, the IEACR Case 15 specific cost becomes US$1,142 per kW_e, compared to the US$1,042 per kW_e contained in Table 8.

Other workers have noted discrepancies when comparing costing studies[34] and have concluded an average specific cost of US$1,401 per kW_e for various oxygen-blown gasifiers (not fluidised bed), when normalised to 500MW_e output and based upon end-1994 US$. Taking an IEACR average for such gasifiers from Table 4, ie US$1,827 per kW_e at mid-1991 US$ for a 300MW_e plant size, and normalising at 500MW_e using

IEACR's scaling factor, together with a 10.9% inflation allowance (3% per annum for 3.5 years), gives the result US$1,731 per kW_e.

IEACR, in their study, concede the likelihood of a ±30% error on their capital cost estimates, which would apply when comparing costs from one study to another. The US$1,401 per kW_e figure for oxygen-blown gasifiers is only 20% below the IEACR scaled result of US$1,731 per kW_e, and so the comparison lies within IEACR's ±30% accuracy limit.

KRW Gasifier Study - Comparative Costs of Electricity

The study team developed estimates for the cost of electricity from the two types of KRW gasifier using a similar technique to that employed by IEACR.

Fuel costs are based upon a typical price of US$38.6 per tonne (US$1.42 per GJ). For a plant of η% net generation cycle efficiency and a fuel of US$1.42 per GJ (HHV), the fuel cost contribution to cost of electricity is $51/\eta$ USc per kWh.

Capital charges contribution to the cost of electricity uses the total capital cost (see Table 8) plus additional allowances; ie pre-paid royalties, spare parts inventory, organisation and start-up costs, working capital and land (referred to as 'owner's costs', plus interest during construction, referred to as AFUDC. This gives the lifetime capital investment, referred to as the 'total capital requirement'. An annual recovery rate (US$ million) is then derived from the expected plant life (presumed to be 30 years) and a suitable interest rate for discounted cash flow (DCF - the study uses both 11.8% and 7.0%; the 7% factor is used in the ensuing Table 9 of this report to give easier comparison with the IEACR results). The annual recovery rate is then divided by the rated plant output, assuming an annual average load factor of 65%, in order to obtain the capital charge contribution in USc per kWh.

Non-fuel operating costs include the O&M fixed costs for labour to operate and maintain the plant, for overheads (20% of labour costs) and maintenance materials. They also include the O&M variable costs which include (as required) limestone, nahcolite, zinc ferrite, fuel oil for start-up, natural gas for operation of the emergency flare, cooling water treatment chemicals, service water corrosion inhibitor, demineralisation costs, nitrogen for inert purging, hydrogen for cooling, and ash disposal (charged at US$2.94 per tonne, typical of the current cost to SCS). A credit is allowed against the oxygen-blown gasifier for the by-product sulphur.

A tabulated comparison of electricity costs is given in Table 9. The air-blown fluidised bed shows the electricity cost advantage that would be expected for a plant configuration that Table 8 showed to have the lower specific costs and the higher cycle efficiency.

The oxygen-blown gasifier results in an electricity cost 8.8% greater than the air-blown gasifier. A large part of this discrepancy arises from the greater capital charge emanating from a lifetime capital investment figure some 15% higher when the gasifier is oxygen-blown. Costs for the ASU, Selexol, Claus and SCOT plant components more than offset the larger vessel sizes required for the air-blown, hot gas clean-up, case. The lower variable cost O&M figure for oxygen blowing is a consequence of requiring no limestone or zinc ferrite, while enjoying substantial by-product credit for sulphur.

The IEACR study (Table 5) estimates 4.75USc per kWh for a KRW air-blown gasifier operating with hot gas clean-up. This is commendably close to the 4.91USc per kWh figure given in Table 9, though some degree of chance enters into this as the financial calculation is not entirely on a comparable basis.

Comparison of Air-blown and Oxygen-blown Gasifiers for High Ash Indian Coals

This study examines the relative merits of air-blown (KRW and moving bed) and oxygen-blown (Shell and Texaco) gasifiers in a $600MW_e$ GCC scheme gasifying bituminous coals of up to 35% ash content[35].

At the time of this study there were three gasification test facilities in India. A 150 tonne per day air-blown gasifier at Trichy working on the moving bed principle developed by Bharat Heavy Electricals Ltd (BHEL). The same company operate an 18 tonne per day fluidised bed gasifier at Hyderabad. In a parallel development to that of BHEL, the CSIR has operated a 24 tonne per day moving bed facility at Hyderabad since 1980.

Technical input to the study was provided by CSIR using their 24 tonne per day facility. They were assisted throughout by Bechtel. The study was funded by the US Agency of International Development (USAID) and was based upon a conceptual design for a mine-mouth GCC plant located at the Northern Karanpura coal field of Bihar State, in NE India. Fuel-gas production is sized to fully load two GE9F gas turbines. The design coal is a non-caking, unwashed run-of-mine from the Dakra seam. CSIR found that the ash content of the coal did not have a significant impact on gas composition.

A minimum of 70% sulphur retention was required in the conceptual study from the chosen 0.7% sulphur (dry, ash free basis) coal. A 75vppm NOx emission at 15% oxygen (155mg Sm^{-3}) was also specified.

Essential features of the four gasifier schemes are given below, with a tabulated performance comparison in Table 10.

KRW Fluidised Bed Gasifier

The coal is dried to 5% moisture and crushed to <6mm. Air from the gas turbine is boost compressed. Six gasifiers are required, each rated at 1,500 tonne per day of coal. Limestone injection is adequate for sulphur emissions control. The raw-gas is indirectly cooled, producing 538°C saturated steam. The gas is then subjected to hot gas clean-up for particulates only; there is no further sulphur polishing required. Ash and sulphided lime from the gasifier is oxidised in a sulphation unit and carbon burn-up cell. NOx emissions from the gas turbine are inherently controlled via the high nitrogen content of the fuel-gas.

Moving Bed Gasifier

Graded 6-50mm coal is assumed to be available, albeit at a premium price. Air from the gas turbine is boost compressed and fed to the gasifier. There are 14 gasifiers, each rated at 770 tonne per day. Raw fuel-gas is cooled and scrubbed to remove tar and oil which are used as a boiler fuel to generate superheated steam. The fuel-gas contains much CO_2 and hence uses a selective (wet) sulphur removal process. The $C0_2$ content avoids the need for additional NOx emission control.

Shell Gasifier

The coal is dried and milled to <88μm. Limestone, similarly milled to <88μm, is added as a flux because of the high ash fusion temperature of the Indian coal. Four gasifiers are required, each of capacity 2,500 tonne per day. Four oxygen plants, each of 1,320 tonne per day of 98% purity are also required. There are two combined cycle streams, each using an advanced GE9F gas turbine. The steam cycle uses reheat, operating at 101bar, 538/538°C.

Texaco Gasifier

In this design the coal is fed as a <420μm, 68% solid, slurry. No limestone fluxing agent is required as the Texaco gasifier operates at a higher temperature than the Shell design. In consequence, there are more combustion products in the fuel-gas (CO2) which requires a selective wet sulphur removal process (by Dow) to minimise $C0_2$ capture. Due to limitations in the size of the slag lock-hopper valve, the $600MW_e$ nominal output GCC required seven gasifiers, each receiving 1,800 tonne per day of coal. Texaco offered direct quench, rather than indirect heat transfer cooling, in order to minimise costs from the impact of needing seven gasifiers. Four oxygen plants each provide 1,860 tonne per day of 98% purity oxygen. No steam or nitrogen injection for NOx control is required due to the low calorific value fuel-gas emanating from quench gas cooling.

Process conclusions from the study are that the Texaco case is worst of all. The high ash content of the Indian coal impacts on the lock hopper size and so requires seven gasifiers. To ameliorate the situation, Texaco chose quench cooling to try to reduce cost, but this causes a significant thermal penalty. To be competitive the Texaco design would need to use washed, ie low ash coal, so reducing the ash lock-hopper bottle neck.

The KRW fluidised bed air-blown gasifier has advantages over both the Shell and the CSIR moving bed designs, as a consequence of using in-bed sulphur removal and hot gas cleaning. Cycle efficiencies and other data are compared in Table 10.

The study concludes that, based upon a cycle efficiency comparison, the air-blown KRW fluidised bed gasifier, or similar, is the most attractive GCC technology for high ash Indian coals. However, at the time of the study (1989) the technologies of hot particle filtration and the ability to convert sulphided LASH bed material into benign sulphate were still largely unproven. Thus, the moving bed gasifer, by using less novel design parameters, was thought to be attractive.

Indian Coal Study - Plant and Electricity Generation Costs

Table 11 presents plant capital costs and cost of electrical power generation for the four gasifier types, based upon end-1989 US$.

The capital costs also include plant costs, engineering, fees and owners costs. It is unlikely that the costs in Tables 11, 4 and 8 can be directly compared, since part (at least) of these additional costs have (in Tables 4 and 8) appeared as capital charges within the breakdown price of electricity generation. Also, the original USA costs have been adjusted to reflect Indian labour rates etc.

Similarly, the cost of electricity was calculated using an Indian estimating procedure and financing structure. Thus, these costs cannot be directly compared to those in Tables 5 and 9.

Although some comparisons may not be directly like-with-like, the trend of the benefits of air-blown gasification over competing systems is shown across all the studies examined.

HOT VERSUS COLD GAS CLEANING

The matrix of cases studied by IEACR allowed the authors to draw some conclusions regarding the cycle efficiency benefits of hot versus cold gas cleaning. The IEACR comparison is given in Table 12.

The ABGC process gains nearly 2 percentage points by the use of hot gas cleaning. Oxygen-blown gasifiers are less enhanced, due to the lesser quantity of fuel-gas affected by the clean-up process.

The CRIEPI (Japanese) study enables further comparisons to be made between hot and cold fuel-gas clean-up for single-stage and two-stage entrained flow gasifiers, see Table 6.

The New University of Ulster has carried out a study of which GCC configuration is best suited to a particular type of coal[36]. The fuel selectivity aspect is more fully commented upon in Section 10 (Fuel Flexibility). In carrying out their study, an ECLIPSE process simulation of various GCC processes was set up, with both hot and cold fuel-gas cleaning. The study reports upon the ABGC, the oxygen-blown Shell, the oxygen-blown Texaco and the oxygen-blown BGL gasifiers. Because the BGL produces its product fuel-gas at a low 500-600°C, accompanied by tars and heavy hydrocarbons which need to be condensed by water scrubbing for recycle to the gasifier, the University of Ulster suggest this gasifier is unlikely to be suited to other than cold gas cleaning. Thus, their work enables comparisons to be drawn between hot and cold gas cleaning for the ABGC and the dry and wet coal feed, entrained flow, oxygen-blown, Shell and Texaco gasifiers.

The advanced hot gas cleaning system modelled for the entrained flow gasifiers consists of cooling to 870°C by clean, recycled, fuel-gas before limestone powder is injected which absorbs some of the H_2S and COS, forming CaS. After cyclone particle separation the gas is further cooled to 600-650°C in a steam raising heat exchanger. The fuel-gas is then reacted with sodium carbonate in a fluidised bed absorber for removal of HC1, which protects the gas turbine from possible condensation of alkali-metal chlorides. The fuel-gas next enters the H_2S removal polishing stage, where it contacts zinc oxide (as zinc titanate) in a fluidised bed absorber which operates in a continuous solids recirculation loop with an adjacent fluidised bed regenerator. The clean gas then passes to the gas turbine, ready for any NOx suppression conditioning that may be necessary (see Section 6.3).

The ABGC gains advantage from its capability for >90% in-bed desulphurisation merely by limestone dosing of the gasifier bed. After downstream particulate capture at 650°C in a ceramic filter, the ABGC produces a fuel-gas which can be fed directly to a gas turbine, albeit giving higher levels of environmentally polluting emissions than with advanced hot gas clean-up. This report will refer to this concept as a 'simple' hot gas cleaning system.

A 'cool' gas variant of this was considered by the University of Ulster, differing from 'simple' hot gas cleaning only by the fuel-gas being cooled to 250°C before passing through the ceramic filter.

It should be noted that although the University of Ulster adopted 650°C as the upper temperature limit for hot gas clean-up in their study, the UK's Clean Coal Power Generation Group (CCPGG) - the developers of the ABGC - currently believe 600°C is a more sustainable value.

In order to complete the ABGC hot gas cleaning comparisons, the University of Ulster also modelled an 'advanced' form of hot gas cleaning system, based upon the addition of a fluidised bed zinc oxide (zinc titanate) absorber/regenerator system within the fuel-gas stream, polishing the removal efficiency of H_2S to around 99%. To prevent the zinc oxide being attacked by fuel-gas HC1, which could form volatile salts needing to be removed before the gas is safe to admit to the turbine, a pre-desulphurisation HC1 removal stage was also added, requiring the addition of $Ca(OH)_2$ to capture chloride as $CaCl_2$. This is subsequently removed in the ceramic filter.

'Advanced' hot gas cleaning is also being studied by British Coal Corporation's Coal Technology Development Division (CTDD), to ascertain what cycle efficiency deficit will occur if the ABGC has to be equipped with such an arrangement in place of 'simple' hot gas clean-up. The CTDD system encompasses not only HC1 removal and H_2S polishing but also has a facility for partial NH_3 removal, for NO_x emission reduction. It includes two stages of particulate removal. Such an arrangement will be necessary if hot gas cleaning is to rival the environmental control standards endemic to cold gas cleaning and which are now the basis of UK emission legislation[37].

In the cold, wet, gas cleaning system modelled by the University of Ulster, the fuel-gas is firstly cooled to 125°C and contacted with water in a scrubber, where NH_3, HC1 and alkali metals are removed. There follows catalytic conversion at 180°C of COS and HCN to H_2S and NH_3, respectively. The exit gas is then cooled, to condense water vapour, before entering the solvent scrubber (M-Sulphinol) for removal of H_2S and NH_3. The solvent is separately treated to recover sulphur in a Claus unit. This system is considered in conjunction with both the Shell and Texaco entrained flow gasifiers and also the ABGC[38].

The results of the University of Ulster's modelling of the various gasifiers, each with hot and cold gas cleaning variations, is given in Table 13.

In Table 13, the negative values refer to percentage point reductions when moving from hot to cold gas cleaning options. It is interesting to note how different coal compositions (the study compared ten coals, ranging from anthracites to lignites) affect the relative efficiencies of the different cleaning methods. Thus, no single set of values is correct for all coals.

The ABGC, although losing only 1.8 percentage points when comparing cold gas cleaning with 'advanced' hot gas cleaning, loses 2.4 percentage points when comparing cold gas cleaning with 'simple' hot gas cleaning. In other words, the University of Ulster calculate the deficit to the ABGC of 'advanced' hot gas cleaning compared to 'simple' hot gas cleaning to be 0.6 percentage points (46.6% becoming 46.0% or 47.3% becoming 46.7%). The current CTDD study suggests somewhat less than this, around 0.4 percentage points.

Comparing the Shell and Texaco processes operating with cold gas cleaning to the 'simple' form of hot gas cleaning used by the generic ABGC process, the ABGC gains an advantage of typically 1.8 percentage points of cycle efficiency (Shell) and 6 percentage points (Texaco). The Texaco system is particularly penalised by cold gas cleaning, due to all the water vapour present in the fuel-gas, both from the water content of the slurry coal feed and also the additional combustion reactions necessary to evaporate the feed slurry to dryness. When the product fuel-gas is subsequently subjected to cold gas cleaning, this water vapour is condensed, moving considerable thermal energy into the less efficient steam cycle.

With a dry coal feed, entrained gasifier, such as the Shell design, IEACR anticipated that cycle efficiency would be scarcely affected. However, a Shell study[61] disputes this conclusion, claiming a 1.3 percentage point cycle efficiency advantage for the incorporation of a General Electric (GE) moving bed hot gas clean-up system[39] into an entrained flow GCC Shell gasifier design study (excluding any on-plant use of derived sulphuric acid). Another Dutch study, looking at Shell gasifiers, found 2.2 to 2.4 percentage points cycle efficiency advantage to hot gas cleaning[40].

The University of Ulster results suggest that a Shell gasifier would benefit up to 1.6 percentage points, dependent upon the coal type. The cycle efficiency of the Shell gasifier was affected by various fuel parameters, in particular the coal's ash content, fixed carbon content, sulphur content, carbon:hydrogen ratio (especially with cold gas cleaning) and higher (gross) heating value (especially with hot gas cleaning). Some of these factors cancel each other out when comparing hot and cold gas cleaning, with the result that the Shell gasifier is difficult to predict regarding what advantage will be obtained by moving from cold to hot gas cleaning. There may, with some coals, be no improvement. A zero improvement was anticipated by the IEACR in their study.

The IEACR suggest that in the case of a (Texaco) wet coal feeding entrained flow gasifier, they would expect a small increase, 0.4 percentage points, in cycle efficiency when moving from cold to hot gas clean-up. This appears an underestimate in comparison to Japanese claims of 0.9 percentage points and figures from the Ulster study of 1.8 to 2.8 percentage points. The latter attribute the significant cycle efficiency penalty of cold gas cleaning to the movement of heat from the gas turbine cycle to the steam turbine cycle when the fuel-gas is cooled and water vapour condensed, prior to cold gas cleaning.

AIR SEPARATION UNITS - THEIR COST AND POWER CONSUMPTION

The two major factors which encourage the use of air-blown gasification rather than oxygen-blown are the cost and parasitic power consumption of the ASU. These factors are influenced by the design of the ASU, with its associated compressors, and the degree of integration of the ASU with the gasifier. A typical ASU flowsheet with a process description is given in Figure 2. Such factors have received consideration in numerous studies because of their strong influence upon the design, cost and net cycle efficiency of a GCC plant.

It should be noted that, throughout this section, data from the USA is quoted extensively. Although not clearly stated in the parent text, it is anticipated that data involving mass of oxygen are in US tons, equivalent to 0.907 SI tonne.

EPRI sponsored an early (1980) study[41] into the technical and economic factors associated with the design and operation of commercial oxygen plants. This study concluded that the ASU might represent 18% of the total installed cost of a GCC power plant and would be the largest internal power consumer. The report quotes the power consumption of an ASU as 397kWh per US ton of oxygen for 99.5% purity and 367kWh per tonne for 95% purity. The cost of gaseous oxygen storage (GOX) for 15mins from a 2,200 US ton per day plant is estimated as US$500,000-US$600,000 (at 1980), while liquid oxygen storage (LOX) for 24h is estimated at US$1.8-US$2.2 million (at 1980).

A similarly dated EPRI report[41] looking at the economics of oxygen-blown Texaco gasifiers found that a 98% oxygen purity ASU contributed 21-22% additional cost to the various schemes studied, based upon a 1,093°C turbine firing temperature (18.3% for a 1,315°C turbine firing temperature). Cold gas cleaning was used in these schemes. Five oxygen plants, each producing 1,673 US ton per day of oxygen (100%) purity basis) were required in each scheme. Capital cost for the oxygen (1978 US$), is calculable as $21,010 per US ton of oxygen per day. Total plant parasitic power is reported as 14.2%o of the plant gross production, with the ASU consuming near 11%. The ASU power consumption is equivalent to 398.5kWh per tonne of oxygen.

Oxygen plant suppliers have recognised the significant market potential for their equipment should GCC oxygen-blown gasifiers become the norm for coal-fired power generation plants. Thus, much effort has been expended to reduce their capital cost and power consumption. The 1980 study was reviewed and updated by representatives of the same company (Praxair Inc, formerly Union Carbide Corporation, Linde Division) in a paper presented in 1994[42]. Over the intervening years, ASU designs for GCC power plants have been honed to the bare needs of GCC technology, cutting away unnecessary peripheral equipment normally required for the co-production of argon, krypton, xenon and neon. Also, designers have concentrated their efforts on large ASUs (2,200 US ton per day of oxygen production) which are rarely needed for merchant gas production. Thus, the ASU enjoys an economy of scale which further helps with cost reduction. As a result, ASU facilities for GCC applications are considerably less complex and less costly than a typical industrial gas production facility.

The ASU design proposed in 1980 produced 98% purity oxygen at 60bar_g at a capital equipment cost, when converted to 1994 US$, of $28,200 per US ton per day. The 1994 design incorporates efficiency improvements which reduce energy consumption by some 10%, while product purity has reduced to 90-95%.

Costs have been reduced not only by simplifying the plant design to the needs of GCC, but also by improvements in control plus instrumentation systems and fabrication plus constructional trends. The resulting cost of a 1994 ASU delivering oxygen only, ie no nitrogen compression or backup system, is claimed by Praxair to be $14,000 per US ton per day, a real cost improvement of 40-50% over the 1980 design. This, the authors believe, reduces the capital cost of an ASU from the 15-20% of total GCC capital reported in 1980 to 7-15% in 1994. Parasitic electrical power consumption of the revised ASU is expected to be 9-13% of the GCC gross system output.

Less optimistic figures are given in an Air Products and Chemicals Inc paper[43], which shows a base cost, in 1987 US dollars, of $18,800 per US ton of oxygen per day for a 2,000 US ton per day single stream plant. They also give cost adders; for a dual stream plant (+15%), for 20 minutes gaseous oxygen storage (+2.6%) and for 12 and 24 hours liquid oxygen storage (+4.8 and +5.4%).

As is usual with process equipment, however, capital cost is not the only criteria on which to make a judgement. O&M costs need to be included in order to make a thorough assessment. Hence, an EPRI study[321 carried out in 1987 compared oxygen plants from BOC Cryoplants, Air Products and Union Carbide (Linde), all producing 1,985 US ton per day of oxygen at 95% purity.

The capital cost and utility requirements of the three were quite different and required a levelised cost analysis in order to select the most economic. Union Carbide were ruled out since their capital and O&M costs were both higher than the BOC Cryoplants' values. The remaining two plants gave very similar total levelised costs, though individual elements varied, see Table 14.

In the above case, the more power efficient Air Products ASU was selected for inclusion in the overall study. Its parasitic power consumption was found to be 6.1% (of 409.5MW_e gross power generation) out of a total parasitic loss of 9.5%. The capital cost of the ASU was 13.5% of the plant construction costs.

Further ways to effect economies within an oxygen-blown GCC are suggested by Air Products, who recommend GCC designers keep an open mind regarding oxygen plant specifications.

Although Air Products envisage a 3% reduction in the cost of producing electricity by using an 'improved' 95% purity ASU, a higher specification plant could also make argon available for sale, which might reduce the capital pay-back time. However, the limited market for argon and its variable price, which is sensitive to geographical location, appear to make this option relatively unattractive.

Other technologies claimed to have been developed, or to be under development, by Air Products, as aids to GCC economics, are the use of oxygen in place of air in a Claus sulphur recovery plant, using COPE℠ technology, the manufacture of methanol from syngas to meet peak energy demand or as a saleable by-product, and an alternative air separation process, MOLTOX™, which absorbs oxygen from the air using a mixture of molten salts which are subsequently regenerated by heating and depressurising.

Japanese workers are endeavouring to develop another alternative to the standard cryogenic method of oxygen separation from air, using Pressure Swing Adsorption (PSA)[44]. They justify their novel approach by the high power consumption of cryogenics, some 0.4kWh Nm^{-3} of oxygen production (280kWh per tonne), compared to the thermodynamic ideal of 0.074kWh Nm^{-3}. They also argue that substantial further reductions in equipment cost for the cryogenic method are unlikely given the current state of the technology. They anticipate that a power consumption of 0.24kWh Nm^{-3} should eventually be attainable for a large scale PSA plant.

ASU plant vendors have carried out their own comparison studies of air-blown versus oxygen-blown GCC designs and inevitably find strong arguments in favour of incorporating ASUs into GCC processes. However, when such studies are based upon high temperature entrained flow gasifiers[15], the conclusions are unfairly biased towards oxygen blowing, since such systems require high air preheat and/or a disproportionate degree of oxidation reaction in order to attain the necessary high temperature slagging operation. Power consumption is adversely affected, due to the presumption of using high temperature air compressors[45] and the resulting fuel-gas volume is maximised, which impacts upon vessel sizes and costs. Limitations on the maximum size of gasifier vessels that can be fabricated and transported then introduce the possible need for multiple air-blown gasifiers for the same duty as a single gasifier of oxygen-blown design.

While the comments within such studies are no doubt factual for the specific circumstances under consideration, the comparisons of plant costs and cycle efficiencies reported in Section 2 show that, under more evenly balanced circumstances, the advantage generally lies with air blowing. The marginal exception is with high temperature entrained flow gasifiers, where plant economics seem roughly equivalent.

One oxygen-blown GCC arrangement which may generally improve on the costs and cycle efficiency of air-blown gasification is where the ASU is integrated with the gas turbine. In such an arrangement part (partial IGCC), or all (full IGCC), of the ASU's air requirement is taken from the outlet of the gas turbine's compressor, rather than from a separate electrically driven compressor.

The air and oxygen compressors account for more than half the capital cost of an oxygen plant. With a fully integrated ASU, the oxygen plant does not require an air compressor, which represents over a third of the cost. Also, the oxygen compressor can be smaller because the oxygen is produced at a higher pressure, around 4.3bar_a. The nitrogen leaving the higher pressure ASU is also around 4bar_a and can be effectively used, via additional compression, by feeding it to the gas turbine's combustion chamber, which provides additional power as well as NOx control[7].

At Puertollano, the integrated ASU is reported to have contributed only 5.2% to the total investment costs (6.5% of plant costs) out of a total investment of US$622.5 million (1991 US$)[46]. The low percentage value is, however, somewhat misleading, since it is based upon a total cost which is relatively high, (reported as US$1,858 per kW_e on a gross electrical output basis, but equivalent to US$2,075 per kW_e on the more usual net electrical output basis).

The Puertollano ASU provides 85% purity oxygen for the gasifier and nitrogen to the gas turbine. Taking oxygen production as 2,245US ton per day (equivalent 100% purity, calculable from the literature[10]) results in an oxygen plant cost of US$14,410 per US ton per day. This cost appears to also cover product compression and also back-up oxygen and nitrogen storage systems.

Studies have indicated that a significant heat rate benefit (2-3% improvement in cycle efficiency) can be obtained with a small (2-3%) reduction in capital cost per kW_e by the expedient of integrating the ASU with the gas turbine[7,34].

1 Extent of Air Separation Unit Integration

There are, essentially, three design approaches for oxygen-blown GCC facilities:

i. a stand-alone (non-integrated) ASU;

ii. a nitrogen integrated ASU;

iii. an air plus nitrogen (fully) integrated ASU[47].

The different approaches are shown in Figures 3a to 3d, which also illustrate a fourth arrangement, derived from the third, namely partial integration. In partial integration a portion of the air for the ASU is taken from the gas turbine and the remainder is taken from a separate compressor.

Stand-alone describes an ASU that is isolated from other units in the GCC facility. Ambient air is separately compressed and sent to the ASU where it is separated into oxygen and nitrogen streams. The nitrogen is vented to atmosphere and the oxygen is further boosted to the pressure needs of the gasifier. Air supply pressure to the ASU is selected so that the vented nitrogen is just above atmospheric pressure, to minimise energy loss.

The nitrogen integrated ASU receives air from a separate compressor, but the ASU may operate at elevated pressure (EP-ASU) compared to the fully stand-alone arrangement. Product oxygen and nitrogen compressors then deliver oxygen to the gasifier and nitrogen to the gas turbine's combustor. Although integrated with the gas turbine through the nitrogen supply, the separate main air compressor allows easier operation and start-up of the ASU, since it is independent of other units in the GCC facility. The pressurised nitrogen, when mixed with the fuel-gas at the turbine's combustor, reduces NO_x formation by reducing the adiabatic flame temperature (see Section 6.3). It also increases the power output of the gas turbine by increasing the mass flow through the expansion stage.

Depending upon the gas turbine design, the nitrogen may either be mixed with the fuel-gas prior to the gas entering the combustor or introduced separately, within the combustion chamber. NOx reduction effects are similar and so introducing the nitrogen directly into the combustion chamber is preferable, since it has the lowest pressure requirement.

Another variable is how hot the nitrogen should be delivered. A multi-stage, intercooled compressor will supply nitrogen either near cooling water temperature (if there is an after-cooler) or up to 120°C. A non-intercooled compressor would deliver nitrogen at 200-260°C, depending upon the pressure ratio. The trade-off is between reduced compressor power versus recovery of the nitrogen's heat energy in the turbine.

Nitrogen integration adds very little to the operating complexity of a GCC, while the elevated pressure of the ASU reduces the size, hence cost, of some components.

The third type of ASU design also operates at elevated pressure, but obtains all or part of its air supply from the gas turbine's compressor. Thus, this plant arrangement is an IGCC configuration. The operating pressure of the ASU is set by the air delivery pressure from the gas turbine, requiring the ASU to operate over a wide pressure range in response to changes in air supply pressure as the gas turbine modulates over its output range. A good ASU control system is key to preserving product purity and flow during turbine load changes and other variable conditions.

IGCC cycle optimisation involves determining the appropriate amount of air to extract from the gas turbine, recognising the amount of oxygen required for gasification and the quantity of nitrogen made available for feeding to the gas turbine's combustion chamber. A further consideration is the recovery of heat contained in the hot air extracted from the turbine. One possibility is to heat exchange it with the returning nitrogen.

With oxygen-blown gasification, the fully integrated ASU (100% of the ASU air comes from the turbine) always has the minimum parasitic power requirement; intuitively one might think that this minimum would provide the optimum IGCC configuration. However, this is not always the case, since the optimum degree of ASU integration is a strong function of the gas turbine's characteristics[15].

In the case of the Buggenum installation, the optimum match with the Siemens V94.2 turbine is where the full air requirement of the ASU is taken from the turbine. Total parasitic power is then 10.9%. In contrast, a 1991 study found that the optimum match with a GE Frame 7F turbine occurred with partial integration scheme, with only 56% of the ASU air taken from the turbine's compressor[48]. Similar results were obtained in a study based upon a Westinghouse 501F gas turbine. The partial integration scheme requires a supplementary air compressor for the ASU, which substantially increases the ASU power requirements (estimated as a +62.5% increase[15]). However, the combined cycle output increases even more substantially, such that the overall IGCC facility improves in terms of net power output, cycle efficiency and specific investment costs. Partial integration appears more complex than other forms of IGCC, but has greater design and operational flexibility. The supplementary air compressor can be designed to provide variable flow to the ASU in compensation of the decreasing gas turbine air flow that inherently occurs on a hot day. It therefore allows the turbine to maintain full power over a range of ambient conditions. It also has the advantage that the ASU can be operated independently of the gas turbine during ASU start-up. The greater the degree of integration of the ASU with the gas turbine the greater is the dynamic control complexity.

It may be justifiable to make a small step backwards from the optimum plant configuration in exchange for a more workable plant configuration. In this context the 1991 GE Frame 7F study[48] also found merit in a nitrogen integrated ASU arrangement, even though this more than doubled the parasitic power compared to the optimum, partially integrated, arrangement. With nitrogen (only) integration, the cycle efficiency is not as good as with partial integration, but the gas turbine has a boost in power output due to air from the stand-alone compressor appearing as additional gas flow at the power turbine. Thus, the specific investment and cost of electricity are competitive with the optimum partial integration case and the scheme has the advantage that the ASU is decoupled from the variable air pressure supply of the gas turbine. Nitrogen integration is being deployed in the Tampa Electric, Polk County project which uses Texaco gasification[18].

Table 15 compares the types of integration included within certain oxygen-blown IGCC demonstration plants.

4.2 Relative Performance and Costs of Oxygen-blown Gasification Combined Cycle versus Integrated Gasification Combined Cycle

TAG™ Assessments

Table 16 presents data taken from the 1993 EPRI TAG™ Technical Assessment Guide[49], showing relative costs of entrained flow GCC plants with non-integrated, partial and fully integrated (IGCC) ASUs. The data is all relevant to a 500MW$_e$ plant, costed in December 1992 US$.

In the TAG™ comparison, the ASU cost is not affected except with the fully integrated case. This may reflect the ASU not operating at elevated pressure in the partial integrated case. In some examples of partial integration the ASU continues to operate at low pressure (LP-ASU), and the turbine extraction air is appropriately expanded downwards in pressure[50]. Under these circumstances, the ASU does not benefit from any cost reduction due to smaller vessel sizes at elevated pressure. Cycle efficiency improves by 2.7% as a result of IGCC full integration, while total plant cost reduces by 14.65%. ASU cost is 9-10% of the total plant cost for all three schemes.

Union Carbide Corporation and General Electric Study of Texaco Gasifiers

This 1987 EPRI study[51] updates and extends an earlier EPRI report[52] by the same authors, completed in 1984. The 1984 non-integrated ASU base case is updated in terms of equipment specification and compared directly with a fully integrated ASU. Both have the same 95% purity oxygen from their respective ASUs (2 per gasifier), **a** production of 4,300 tonne per day. Gasifier operating conditions are kept constant, including the coal feed rate and the oxygen:coal ratio. The oxygen enters the gasifiers at 50.6bar_a. Fuel-gas composition and temperature to the gas turbine are the same in both cases. Both schemes use cold gas clean-up. Where the schemes differ is:

- there are no stand-alone air compressors with the integrated ASU;
- the integrated ASU is re-designed to operate at the pressure of the feed air from the gas turbine (an EP-ASU) rather then from a stand-alone compressor (14.2bar_a vs 5.9bar_a);
- the oxygen compressors are of reduced size with the integrated ASU;
- nitrogen from the integrated ASU is fed to the gas turbine whereas it is vented to atmosphere from the non-integrated case;
- the integrated scheme has a air/nitrogen heat exchanger to exchange heat between the gas turbine air flow (at 380°C) with nitrogen from the ASU. Clean cold air from the turbine eventually enters the integrated ASU at 13.7bar_a and 4°C;
- the integrated ASU scheme incorporates a 24 hour supply of LOX storage from one of the two oxygen plants. GOX storage is provided for 20mins of supply from one oxygen plant.

The power consumption of all plant items apart from the ASUs is 22.1MW_e. The non-integrated ASU adds a further 66.0MW_e while the integrated ASU adds 37.8MW_e. However, the integrated ASU requires the gas turbine to compress 4.8% additional air to compensate for that proportion of the compressor air which goes to the ASU and is converted into fuel-gas (equivalent to the oxygen part of the integrated ASU air supply). This creates an air flow deficit at the turbine combustor compared to the non-integrated case. Producing the additional air flow absorbs 18.5MW_e more gas turbine power. Thus, the overall power consumed by the ASU in the integrated case is 56.3MW_e.

The gross power output for both schemes is 679.8MW_e.

In the non-integrated ASU case, the oxygen plant absorbs 9.7% of the gross power and the remaining auxiliaries consume 3.25%. Thus, the total parasitic plant power consumption is near 13%, leaving 591.7MW_e net electricity production at 38.06% cycle efficiency (% of coal HHV)

For the integrated EP-ASU, the oxygen plant (including the turbines additional work of 4.8% further air compression) absorbs 8.3% (5.7% if the additional turbine compression is ignored) and the remaining auxiliaries consume 3.25%. Thus, the total parasitic plant power consumption is 11.5%, leaving 601.4MW_e net electricity production at 38.7% cycle efficiency. This is a 1.7% heat rate improvement over the non-integrated case.

Plant costs reduce by US$20 million (1986 dollars) with the integrated ASU scheme. The ASU is reduced in cost from US$95.7 million (US$22,256 per tonne of oxygen per day) to US$74.2 million (US$17,256 per tonne per day), saving US$21.5 million, but an estimated US$1.5 million would be required to modify the turbines suitable for the air extraction required for operating in an integrated mode. Cost savings for the integrated ASU are largely due to the elimination of the air compressors and the use of smaller columns and heat exchangers due to the increased operating pressure. The integrated ASU cost savings are equivalent to a reduction of 2-3% in capital costs. Operating costs are anticipated to decrease only slightly.

The same team carried out a further study which was reported in 1991[53], which yields similar results. They report that integrating the ASU results in a power output improvement of 2.8% and a cycle efficiency improvement of 1.7%. Since capital costs associated with integration are approximately offset by capital cost savings associated with the EP-ASU, the unit capital cost in US$ per kW_e is expected to fall by the same percentage as the net power output improvement, ie 2.8%.

Krupp-Koppers PRENFLO Gasifiers with Integrated and Non-Integrated ASUs

The results of a study comparing oxygen-blown Krupp-Koppers PRENFLO gasifiers, operating with their ASUs in non-integrated and integrated modes, was reported in 1991[54].

The non-integrated arrangement incorporates a conventionally designed ASU with a separate air compressor. The ASU products, ie oxygen at 95% purity and 99% purity nitrogen, are delivered at near atmospheric pressure. The nitrogen is vented and control of turbine exhaust NOx is by adding water vapour to the fuel-gas.

The fully integrated concept, where all the air for separation is taken from the gas turbine compressor, has an unconventionally designed ASU, with the air separation column operating at an elevated pressure producing 85% purity oxygen and 97% purity nitrogen at 5bar to 6bar. The nitrogen is then delivered to the turbine's combustor for NO_x control by fuel-gas dilution.

Both schemes are based upon Linde air separation and Siemens V84.4 gas turbines. The nominal plant rating is 400MW_e. The gasifiers operate at about 20bar. Raw-gas exits at 1,000°C and is cooled in syngas coolers to generate steam. It is then dedusted in ceramic filters. Dry dedusting facilitates fly ash recycling and increases plant efficiency by some 0.5 percentage points over wet dedusting, due to a lower pressure drop and less heat loss. After dedusting the fuel-gas is cooled further to 120°C in wet scrubbers and near 100% of the hydrogen sulphide is scrubbed out and converted to marketable sulphur. The high sulphur removal efficiency permits greater sensible heat recovery from the stack gases by virtue of a lower sulphuric acid dew point. Increasing sulphur capture from 95% to 99% potentially raises the power generation efficiency by 0.3 percentage points.

After sulphur removal the fuel-gas passes to the gas turbine combustor. In the non-integrated scheme the fuel-gas is first moisturised to 43.5% volume for NO_x suppression. In the integrated case, the availability of nitrogen diluent reduces the need for moisturising to only 9%. This latter is an advantage, since high moisture saturation can lead to excessive CO emissions.

Table 17 summarises the power generation and economic performance of the two concepts.

A **Fluor Daniel and Destec Energy Study of Integrated and Non-Integrated ASUs**

In this study[55] a techno-economic comparison is presented for two design options of a Destec (two-stage, slurry fed entrained flow gasifier) based GCC power plant. One consists of an IGCC arrangement, where an EP-ASU receives part of its air supply from the gas turbine, while the other is a more conventional approach involving a LP-ASU that is not integrated with the gas turbine.

An ISO ambient temperature of 15°C was utilised for developing performance and cost data, with a site elevation of 212m above sea level. The coal used is a sub-bituminous with 23% moisture content; gross (higher) heating value (HHV) 25,440kJ kg^{-1} (dry basis) and net (LHV) of 24,420kJ kg^{-1} (dry basis).

Plant designs were developed to fully load two General Electric MS7001FA gas turbines. Under these conditions, the LP-ASU design consumes 5,085 tonne per day of coal and the EP-ASU design consumes 5,044 tonne per day of coal (moisture free basis). Costs for the ASUs were provided by Air Products.

The integrated EP-ASU design differs from the LP-ASU design in the following ways:

ASU Design

Both schemes have two ASUs operating in parallel, each with air compressors and oxygen product compressors. The EP-ASUs also have product nitrogen compressors and excess nitrogen expanders.

Each of the LP-ASUs have a capacity of 1,945 tonne per day (100% 0_2 basis) of 95% purity oxygen. Atmospheric air is compressed by a stand-alone compressor to 4.8bar_a, cooled and then passed through the cryogenic ASU. The emergent oxygen is then compressed and fed to the gasifiers. A small amount of nitrogen, required by the gasification system, is also compressed. The majority of the nitrogen is vented. The compressors consume 50.8MW_e of electrical power at full capacity.

The two EP-ASUs each have a capacity of 1,930 tonne per day of 95% purity oxygen. 39% of the ASU air is extracted from the gas turbine while the remainder is supplied by separate air compressors, which compress atmospheric air to the same pressure as the air extracted from the turbine. The product oxygen stream, leaving the ASUs in the pressure range 4bar_a to 5bar_a, is compressed and fed to the gasifiers. The nitrogen stream, also at 4bar_a to 5bar_a, is compressed to 20bar_a before being preheated and fed to the gas turbines' combustors. The compressors consume 68.8MW_e of power at full capacity. Note that this is a higher power consumption than the non-integrated case, presumably reflecting a relatively modest amount of turbine air extraction (only 39% of ASU requirements) which still leaves 61% of the air being delivered by compressor, but to a higher delivery pressure than in the non-integrated case. With the EP-ASU there is further power consumed in boosting the product nitrogen to a pressure sufficient for it to enter the turbines' combustors.

Fuel-gas Moisturisation

In both schemes the clean fuel-gas from the sulphur removal unit (SulFerox) is essentially free of moisture and enters the bottom of a moisturiser vessel.

In the case of the LP-ASU the fuel-gas is saturated within a two-stage column to 41.2% by volume moisture, to control NOx to 25mg Sm^{-3} (15% O_2, dry basis). The fuel-gas is then preheated and supplied to the turbines' combustors.

In the case of the EP-ASU, a single stage moisturiser column is all that is necessary to supply 10.5%) moisture to the fuel-gas. The remaining fuel-gas dilution for NOx control is via injection of ASU nitrogen into the turbines's combustors.

Gas Turbine Combustors

The LP-ASU scheme has reheated and moisturised fuel-gas, at 270°C and 5.2MJ Sm^{-3} (LHV), entering the turbines' combustors. The fuel-gas moisture not only reduces the formation of NOx, it also increases the mass flow through the expanders. Due to mechanical limitations of the gas turbines, their air inlet guide vanes are set to reduce air input, such that net power output from the gas turbines remains at 192MW_e (each turbine).

In the EP-ASU scheme, 39% of the ASU air requirement originates from the gas turbines. Moisturised and preheated fuel-gas at 270°C and 7.9MJ Sm^{-3} (LHV), together with compressed nitrogen at 333°C, are fed to the turbines' combustors. The moisture in the fuel-gas and the injected nitrogen serve to control NOx formation and provide additional motive fluid to the turbines. The amount of nitrogen injected is constrained by flame stability criteria, which results in using 86% of the available ASU nitrogen. The remaining nitrogen is expanded through a power recovery expander. The amount of extracted air from the turbines leaves them fully loaded, each producing 192MW_e.

Table 18 summarises the performance and economics of the two concepts.

The integrated EP-ASU shows a 2.7% advantage in cycle efficiency compared to the LP-ASU and the cost of electricity is 1 to 1.5% lower, depending upon the cost of coal. Specific capital costs, on a US$ per kW_e basis, are almost identical.

The integrated EP-ASU, by virtue of its higher cycle efficiency, confers an environmental advantage by generating less CO_2 per MW_e of power produced. Also, the use of nitrogen as a NO_x suppressant results in 14% lower water usage per MW_e of power generated, which could be an important factor for a GCC installed in an arid climate.

In some respects this study is less encouraging than others using an EP-ASU. In particular it predicts the absorbed power for the EP-ASU being higher than for the LP-ASU. This seems to be a consequence of only 39% of the EP-ASU air coming from the gas turbine. The authors state that GE are making efforts to reduce the required supply pressure for the nitrogen entering the turbines' combustors, which would assist by reducing the EP-ASU power consumption.

ASU costs remain at 13.5% of the total plant costs in both schemes. There is no reduction in ASU cost in the IGCC arrangement, despite the EP-ASU presumably benefiting from the use of smaller vessels (higher operating pressure). The report makes no comment upon this.

The authors concede that the EP-ASU and gas turbine systems have major interdependences which may result in more difficult start-up and shutdown sequences. On the other hand, they foresee occasions when the facility to control the turbine air extraction rate could offer a more efficient means of operating the gas turbines, ie under off-design conditions.

AIR SUPPLY INTEGRATION VARIATIONS WITH AIR-BLOWN GASIFICATION

The previous section has shown how, depending upon the model of gas turbine being considered, capital costs, parasitic power consumption and cycle efficiency can all be improved when the ASU is fully or partially integrated. However, it is not just oxygen-blown gasification systems which offer the design flexibility, from stand-alone to fully integrated ASU, of how the gasifier oxidant supply is linked with the gas turbine. Air-blown gasification also has this option, though air-blown gasification is conventionally presumed to have a fully integrated air supply, where the air compressor delivering to the gasifier draws its air supply from the outlet of the gas turbine compressor.

A market prospect for non-integrated gasification is the conversion of natural gas fired turbines to operation with coal. Here, a stand-alone GCC design could provide a fuel-gas source in substitution of the original natural gas supply, with minimal modification necessary to the gas turbine and its associated steam cycle.

The conventional, fully integrated, air-blown GCC arrangement has evolved partly as a function of gas turbine compressor/expander capacity considerations and partly because of the high power requirement of compressing air from atmospheric pressure all the way up to gasifier pressure, using a stand-alone compressor. Without a complex ASU to be concerned with during start-up and part-load operation, there is no reason why air-blown gasifiers should not adopt the most efficient, fully integrated, arrangement. A conventional fully integrated air-blown gasification cycle is illustrated in Figures 4.

Partial and non-integrated design options do, however, exist and have been examined for the ABGC concept[56], revealing trade-off benefits in power output and electrical generating costs against small reductions in cycle efficiency.

The study was based upon gas turbine performance data obtained by European Gas Turbines. The Frame 9FA gas turbine is very flexible, such that its power output can be increased from its normal value, about 225MW_e, to a maximum of 275MW_e by putting additional gas flow through the expander turbine.

The base case for the ABGC study was the conventional arrangement, where 100% of the air required for the gasifier is provided by the gas turbine compressor. This air is cooled to minimise the power requirement of the booster compressor, which then further boosts the air to the 25bar pressure required by the gasifier. The air mass flow taken from the gas turbine is later returned to the turbine as fuel-gas, resulting in a gas turbine power close to the standard natural gas rated output.

Within this study the ABGC was of a configuration where all the fuel-gas desulphurisation takes place within the air-blown fluid-bed gasifier, using limestone addition. Downstream hot gas clean-up consists only of a post-gasifier cyclone and subsequent ceramic candle particulate filtration at 600°C. The 20-30% of the coal which remains as unconverted char is removed from the gasifier, cyclone and ceramic filter and burned in a CFBC, raising additional steam. Predicted overall net plant efficiency is 46.9% (LHV basis). This high cycle efficiency value is attributed by the authors to:

- no oxygen plant, consuming high auxiliary power;

- sulphur being retained in the gasifier, so the energy losses in gas cleaning are low;

- no lost energy in melting the residual ash into slag;

- overall carbon utilisation is high;

- high temperature heat is available in an oxidising environment within the CFBC, which means that a high efficiency single pressure reheat steam cycle can be used.

The other extreme from the fully integrated (conventional) case is the arrangement where all the gasifier air is taken from atmosphere and boosted directly to gasifier supply pressure in a stand-alone electrically driven compressor. This configuration allows the maximum flow through the gas turbine expansion stage, since the full flow of the low calorific value fuel-gas is added to the full flow of the gas turbine compressor discharge. The high expander mass flow increases power output, but to the extent whereby it is considered necessary to reduce the turbine firing temperature and reduce the volume of gas presented to the expansion turbine, avoiding the turbine compressor coming too close to its surge limit.

Since the gasifier air supply is separately boosted, it acquires the efficiency losses and additional parasitic power consumption associated with an electrically driven compressor (compared to the gas turbine compressor).

These counter-effects increase net power output by 3% but reduce cycle efficiency by 0.8 percentage points.

Possibly the optimum configuration is partial integration, where sufficient air is extracted from the gas turbine to avoid fears of overloading the expansion stage, so allowing the full turbine firing temperature to be used. This occurs (in the case of the Frame 9F) when around 50% of the gasifier air is extracted and boosted and 50% is supplied entirely from a stand-alone compressor. The net output from such a configuration is calculated to be 9.4% higher than the base condition, with a reduction in the cost per kW_e of exported electricity. Cycle efficiency is marginally (0.2 percentage points) lower.

The authors state their belief that such an arrangement offers the best overall option with most modern gas turbines. Total parasitic power consumption on a $555MW_e$ (gross) scheme is deduced as 9.9%, with the gas turbine extraction air boost compressor absorbing 0.5% and the stand-alone air compressor absorbing 4.6% (or 5.5% including the coal feeding air compressor). Thus, the parasitic power absorption of the air supply system totals 5.1% (or 6.0%) of the gross power generation, roughly a half the parasitic power of a similarly partially integrated ASU within an oxygen-blown IGCC plant.

ENVIRONMENTAL COMPARISON

In carrying out the comparison between KRW fluidised bed gasifiers in the air and oxygen-blown modes[33], the study team made special note of a range environmental issues which need to be considered in making the choice between air-blown and oxygen-blown gasifiers. The range of issues also encompass the choice between hot and cold gas cleaning.

Although specific to one particular comparison study, many of the issues raised are of general significance within the air-blown versus oxygen-blown debate and, being based upon the KRW fluidised bed gasifier, substantially apply to the air-blown fluidised bed gasifier within the ABGC.

Cycle Efficiency

Relative efficiency values are, of course, influenced in any study by such factors as the utilisation efficiency of the raw fuel, choice of hot or cold gas clean-up, wet or dry coal feeding, optimisation of the process flowsheet and minimisation of auxiliary (plant) power consumption (the final two will include the effect of an integrated, part integrated or non-integrated ASU). Environmentally, the highest cycle efficiency may not be the best arrangement, nor be the optimum choice from the point of view of operating economics (USc per kWh).

Comparing air-blown and oxygen-blown KRW gasifiers, the air-blown case requires 5.2% less coal to generate the same amount of electricity, so should have an accordingly reduced environmental impact (though much of this advantage is accrued from the carbon burn-up cell, which is included in the air-blown arrangement in order to oxidise the LASH bed material. Such a cell could equally be included within the oxygen-blown variant). The air-blown scheme also has the lowest electricity generation costs, so is also best from an economic point of view.

Interestingly, despite consuming less coal, the air-blown case can emit more CO_2 to atmosphere. The limestone, which it calcines for in-bed sulphur capture, can increase the CO_2 release more than the saving from reduced coal burning. Fortunately, this will only be the case for relatively high sulphur coals, where the feed rate of limestone is necessarily high.

Sulphur Dioxide Emissions

There is considerable flexibility in the amount of sulphur capture possible when comparing hot and cold gas clean-up systems. In-bed limestone feeding is capable of >90% sulphur capture without the need for further downstream processes. This simplified arrangement was exploited in the IEACR study (Section 2.1) with consequent benefits in cycle efficiency and plant costs. The addition of hot fuel-gas removal of HCl, nitrogen compounds and H_2S, improves sulphur capture to over 99% and also reduces NOx formation, but penalises cycle efficiency by possibly 0.6 percentage points.

The IEACR study limited its cold, wet gas sulphur removal to around 90% by incorporating the Purisol process. In the KRW study, wet sulphur removal is by the Selexol process, capturing an overall 96.4%. Various similar fuel-gas washing systems are available from a range of manufacturers[7].

At the extreme, cold gas washing techniques can achieve over 99% sulphur removal, though the highest values entail some penalty on the cost of electricity. Over 99% removal is obtained by incorporating a COS hydrolysis stage upstream of the H_2S desulphurisation wash. One such process (commercial name COMBISULF), combining COS hydrolysis with subsequent methyldiethanolamine (MDEA) washing for H_2S removal, is being incorporated downstream of the oxygen-blown Krupp-Koppers PRENFLO gasifier at the Elcogas plant, Puertollano[10]. Flue gas SO_2 emission is anticipated to be <4mg Sm^{-3} (15% O_2, dry, 0°C). The Rectisol process, licensed by Lurgi, is also capable of +99% capture.

Thus, both hot and cold gas clean-up methods are capable of similarly high capture efficiencies. There is, however, a potentially serious disadvantage with hot gas clean-up when the removed sulphur becomes calcium sulphide (CaS) within the gasifier's LASH bed material.

Although LASH is intended to be oxidised to environmentally benign calcium sulphate ($CaSO_4$) in the sulphator vessel (or the separate CFBC of the ABGC), achieving a high conversion is proving more difficult than anticipated[31,67]. Unless this can be overcome, the LASH solids may acquire a hazardous waste classification, increasing their disposal cost. Recent work related to the ABGC suggests it may be possible to avoid the presence of sulphide by optimising CFBC combustion conditions, sorbent type and sorbent size.

Of course, air-blown fluidised bed gasification could, as an alternative, perform desulphurisation exclusively on the fuel-gas, by either hot or cold clean-up methods, and omit limestone feeding to the bed entirely. However, both methods detract from the highest cycle efficiency (Section 3) and also increase plant costs (possibly +3%[21]).

The cold gas cleaning system, normally associated with oxygen-blown gasifiers, has no ash disposal difficulties. Indeed, depending upon the type of gasifier and the sulphur sorbent treatment system employed, oxygen-blown gasifiers can render all their ash as a glassy, unleachable, residue (entrained flow gasifiers) and can generate sulphur or sulphuric acid as saleable by-products. Nonetheless, there is little doubt that hot gas clean-up will be the eventual route chosen by gasification processes seeking to achieve the highest possible cycle efficiency.

Air-blown gasification is disadvantaged by around 2 percentage points if it uses wet, cold gas, clean-up. Thus, there is an incentive to use hot gas cleaning methods. Acknowledging the novelty of such clean-up arrangements, designers seek to minimise prototype novelty by performing all desulphurisation using in-bed limestone. It is that expedient which introduces the potential difficulty with LASH bed material disposal.

Nitrogen Oxide Emissions

Nitrogen oxides are formed in gas turbines via two mechanisms:

i. oxidising of the nitrogen in the fuel, ie the 'fuel-bound nitrogen' NO_x;

ii. the combination of oxygen and nitrogen in the oxidising air, ie 'thermal'-NOx.

In the reducing medium of the gasifier, the fuel bound nitrogen yields molecular N_2, NH_3 and HCN. With cold gas clean-up systems, as are common with oxygen-blown gasification, the HCN and NH_3 are almost completely removed by hydrolysis and wet scrubbing and appear in the sour condensate effluent. Therefore, NOx emissions from fuel bound nitrogen are largely avoided.

With air-blown gasification, a hot gas clean-up system is most likely to be applied. Although the presence of limestone in the gasifier accelerates cracking reactions, reducing the concentration of nitrogen compounds, the raw fuel-gas still contains a significant amount of NH_3. Some reduction in NH_3 can be made using advanced hot gas clean-up techniques, but the concentration which enters the turbine combustor remains relatively high, giving substantially greater NOx emissions than from cold gas clean-up.

Thermal-NOx is produced in the exhaust flue gases from the gas turbine from both oxygen and air-blown gasifiers, to a concentration which is related to the stoichiometric adiabatic flame temperature (AFT) of the gasifier fuel-gas. The stoichiometric AFT is the temperature which is achieved when a fuel is burned with no excess air and no heat exchange with the environment.

Oxygen-blown Gasification

Typical compositions of actual fuel-gases from several oxygen-blown processes, and their computed stoichiometric AFTs, are given in Table 19 assuming the gas is fed to the burner at 15bar and 300°C[40]. The stoichiometric AFT of pure coal-gas is reported to be around 2,500°C[50].

Although the calorific value of natural gas, around 31.4MJ Sm^{-3} (LHV), is much higher than gasifier fuel-gas, fuel-gas requires only one quarter the amount of combustion air (on a volumetric basis). Thus, the adiabatic combustion of natural gas at 10bar pressure and with 300°C preheat is around 2,150°C after allowing for temperature reduction through dissociation effects[57].

By using reaction kinetics to deduce the likely difference in NOx formation between natural gas and fuel-gas, flame temperature is found to be the dominant factor. Although under actual gas turbine conditions, oxygen-blown gasifier fuel-gas might give rise to an AFT only 120°C higher than that of natural gas, this still implies a NOx formation rate 2.5 times as great.

Thus, it is environmentally unacceptable to burn fuel-gas from oxygen-blown gasifiers without including some means to limit NOx production. This is conventionally achieved by mixing the fuel-gas with steam and/or nitrogen. At Buggenum, nitrogen from the integrated ASU is injected into the fuel-gas and brings the AFT down to 1,840°C[58]. Combustion calculations show that the temperature drop produced by the mixing of nitrogen with the fuel-gas is exactly a half that obtained by mixing the same mass flow of steam. One kg of steam per kg of fuel-gas produces a fall of 150°C in the AFT, whilst one kg of nitrogen per kg of fuel-gas gives a 75°C drop. It is equally possible to inject water into the gas turbine's combustion chamber, but this would have an efficiency penalty and is not considered to be a practical solution.

Non-integrated ASUs in GCC plants generate their nitrogen (and oxygen) at near atmospheric pressure in order to minimise wasted energy. The nitrogen is then vented to atmosphere, such that saturating the fuel-gas with water vapour is the only practical means of achieving AFT dilution.

Highly moisturised fuel-gas (43.5% by volume) results in a claimed[54] NOx formation of 85mg Sm^{-3} at 15% O_2 but also engenders CO emissions in the gas turbine exhaust gases. Fuel-gas saturation with moisture is therefore limited by the maximum admissible CO emission. Preheating the fuel-gas mixture before combustion (up to 300°C) reduces the likely CO emission.

When the gasifier is integrated with an ASU in an IGCC arrangement, the nitrogen from the ASU is available at an elevated pressure. The value of this nitrogen is only fully realised if it is further boosted in pressure and passed to the expander turbine. In so doing, it also suppresses NOx formation. Not only does the nitrogen inflate the volume of gas flowing through the expander, increasing gas turbine power generation, it also replaces the steam which would otherwise be used for NOx suppression. All the steam is then available to generate power via the steam turbine. In one study[54], 12% more steam turbine power was claimed to be produced through minimising the amount of heat consumed in saturating the fuel-gas.

Often, in schemes based upon integrated ASUs, the fuel-gas is partially humidified as well as receiving surplus nitrogen from the ASU. Humidifying the gas provides a heat sink for low level heat within the IGCC process that could not otherwise contribute significantly to the steam cycle.

Table 20 gives examples of oxygen-blown gasifier gas after desulphurisation, humidification and addition of excess nitrogen.

Air-blown Gasification

Thermal NOx is less of a problem with air-blown gasification, since the inherent nitrogen already dilutes the calorific value of the fuel-gas to between 4 and 5MJ Sm^{-3} (LHV), giving it a low AFT, <1,700°C. However, the problem of fuel bound NO_x production remains and will equally afflict oxygen-blown gasifiers if they move to hot gas clean-up for the benefits of higher cycle efficiency.

It would be possible to incorporate selective catalytic reduction (SCR) into the exhaust stack gases but the cost of such equipment does not encourage this option. Instead, as discussed in Section 7.2, work is in progress to trap nitrogen compounds within the hot gas cleaning system, along with HC1 and H_2S.

In Japan, research is being conducted into air-blown gasification which produces l,000vppm of ammonia in the fuel-gas. Special low heating value burners are being developed that suppress NOx formation by small additions of C_3H_y (sic: should possibly read C_xH_y). Emission levels of 170mg Sm^{-3} (expressed as NO_2 at 0°C and 15% O_2) are claimed in the waste flue gases[40].

In the case of the ABGC, 20-30% of coal combustion takes place in the CFBC char combustor. CFBC technology has an inherently low-NOx emission; <80mg Sm^{-3} is anticipated for char combustion (expressed as NO_2 at 0°C and 15% O_2). Current NO_x emission expectations for the complete ABGC concept is 170mg Sm^{-3} (at 15%O_2) but it is ultimately hoped to reduce this, by appropriate advanced hot gas clean-up developments, to <100mg Sm^{-3} (at 15% O_2)[59].

Gas Turbine NOx Emission Investigations

All the major turbine manufacturers have carried out comparative studies of the thermal NOx emission characteristics of their machines when burning gasifier fuel-gas.

General Electric carried out an EPRI sponsored study. First results were reported in 1990[60] and follow up information was reported in 1993[61]. A GE laboratory test facility was used to simulate conditions that would apply in a MS7001F machine. This work found that NOx emissions could be correlated as a simple function of fuel-gas AFT, including any fuel preheat. Fuel-gas net calorific values of 11MJ Sm^{-3} LHV down to 3.3MJ Sm^{-3} were studied and found to fit the correlation. High NOx emissions of l,000mg Sm^{-3} (15% O_2) from a fuel calorific value of 10.7MJ Sm^{-3} reduced to

<50mg Sm^{-3} (15% O_2) when the fuel-gas calorific value was diluted with nitrogen or water vapour to below 5MJ Sm^{-3}.

Testwork included fuel-gas calorific values modified by the addition of water vapour and nitrogen diluents; it also included a simulated low calorific fuel-gas as would be generated by air-blown gasification. Interestingly, emissions of CO were found to be no higher, when the fuel-gas was diluted with water vapour than when it was diluted with nitrogen, except when the H_2:CO ratio in the fuel-gas fell below 1:1. Unfortunately for oxygen-blown gasification, all the analyses shown in Tables 20 and 21 have H_2:CO ratios <1. Under these circumstances, CO emissions rose sharply as the fuel-gas net heating value fell below 5.5MJ Sm^{-3}.

The work by GE also examined the tendency for CO emissions to rise as turbine firing temperature falls at part-load. The fuel-gas was saturated with water vapour down to a net calorific value of 3.3MJ Sm^{-3} but had a H_2:CO ratio of around 3:1. With such a low calorific value, the NOx emissions became undetectable for combustor exit temperatures below 1,100°C (full firing temperature around 1,350°C) but CO emissions then rose sharply. At a firing temperature of 1,000°C the CO emission level exceeded l00vppm.

The simulated air-blown fuel-gas was contaminated with NH_3 by injecting NH_4OH solution containing 30% NH_3 by weight. The conversion rate of NH_3 to NOx was found to be considerably lower than expected, particularly for NH_3 fractions <500vppm (<40% conversion). NOx emissions were around 115mg Sm^{-3} (15% O_2). Rather than claiming credit for this via the design characteristics of the 7F multi-nozzle burner system, GE report their intention to carry out further testwork.

Similar testwork to that by GE is reported by ABB[62], who favour premix combustion as compared to conventional diffusion burner techniques. They claim premix techniques have greater potential for low-NOx emissions after only moderate nitrogen dilution. This avoids the potential problem of high CO emissions from diffusion burners when resorting to saturating the fuel-gas with water vapour to achieve the same NOx results. With a simulated oxygen-blown fuel-gas, ABB found an undiluted NOx emission of 720mg Sm^{-3} (at 15% O_2) but this fell to <50mg Sm^{-3} when the gas calorific value was diluted to 6.5MJ Sm^{-3}.

Tests carried out by Rolls-Royce pic for British Gas pic at Westfield using a SK30 gas turbine found water injection gave lower NOx emissions (87.5mg Sm^{-3} at 15% O_2) than steam (155mg Sm^{-3} at 15% O_2)[63]. These values compare to an uncontrolled release of 50mg Sm^{-3} at 15% O_2 from natural gas. Both water and steam injection into fuel-gas led to increased CO emissions but the CO increased less markedly with water injection.

Summary of NOx **Emission Control Strategies**

There is little doubt that the advantage in terms of NOx emissions currently lies with gasification processes working with cold gas clean-up such that fuel bound nitrogen compounds are removed prior to the gas turbine. This implies an advantage to oxygen-blown gasification. NOx emissions <50mg Sm^{-3} (15%O_2) seem available provided the fuel-gas calorific value is diluted to around 5MJ Sm^{-3} (LHV). The GE testwork suggests that dilution solely with steam is becoming an unsatisfactory technique by the time dilution has brought the fuel-gas calorific value down to 5MJ Sm^{-3}; although NOx emissions continue to fall, CO emissions rise steeply. Thus, dilution with nitrogen is ultimately preferable, suggesting the need for either nitrogen integration of the ASU or (at least) a partial IGCC arrangement. It is notable that for control of thermal NOx with oxygen-blown gasifiers, the fuel-gas calorific value needs to be reduced to the level given by air-blown gasification.

In the medium term, oxygen-blown gasifiers will equally wish to use hot gas clean-up, which must encourage the development of hot gas techniques that remove nitrogen compounds along with the HC1 and H_2S. Such developments will profit both air-blown and oxygen-blown gasification. In the meantime, of course, by remaining with cold gas clean-up, oxygen-blown designs can achieve impressively low-NOx emissions. In this respect, time is on the side of oxygen blowing.

Another influencing factor is that both UK and USA emission standards are getting ever more stringent. The USA has a New Source Performance Standard (NSPS) equivalent to 60mg Sm^{-3} (15% O_2) and the UK has guidance notes to pollution inspectors[37] which signal future intentions to limit NOx emissions from power generation plants to 45mg Sm^{-3} (15% O_2). Under these circumstances, GCC will require either:

- cold gas cleaning;
- hot gas cleaning with substantial removal of nitrogen compounds as well as H_2S;
- selective catalytic reduction (SCR) of NOx within the flue gas exhaust.

Thus, tighter emission standards mitigate against simple hot gas cleaning, ie in-bed desulphurisation by limestone injection to a (fluid-bed) gasifier followed by fuel-gas particulate capture.

Trace Element Emissions

Apart from the generally recognised gaseous and particulate pollutant emissions, emission standards are becoming increasingly stringent with respect to metals, metalloids and organic compounds.

As the fuel-gases cool, most alkali metals are expected to be condensed onto, or adsorbed by, the ash particles. This applies for most metals even at (current) hot gas filtration temperatures and will certainly be the case for cold gas cleaning, where the entire fuel-gas flow may be cooled to 40°C.

However, hot gas cleaning is less likely to remove the more volatile metal elements that exist in coal, ie selenium, zinc, cadmium, arsenic, tin and especially, mercury and lead. Also, reliance will need to be placed on destruction within the gas turbine combustor of any heavy organic compounds which escape the gasifier. This is an environmental issue which must await firm operational data. Although there is no current evidence to suggest that trace element emissions will be of concern, theory places an advantage with cold gas cleaning, which in turn tends to favour oxygen-blown gasification.

RELIABILITY AND OPERATIONAL ISSUES

In carrying out the comparison between KRW fluidised bed gasifiers in the air-blown and oxygen-blown modes[33], the study team made special note of a range of practical issues which need to be considered in making the choice between air-blown and oxygen-blown gasifiers.

Although specific to one particular comparison study, many of the issues raised are of general significance within the air-blown versus oxygen-blown debate and, being based upon the KRW fluidised bed gasifier, substantially apply to the air-blown fluidised bed gasifier within the ABGC.

Gasifiers

The study team assumed there to be little difference in the reliability or operability of oxygen-blown and air-blown gasifiers. Although the oxygen-blown case has the added complexity of an ASU, this is not considered to be of particular concern in an non-integrated GCC arrangement, especially if liquid oxygen (LOX - 0.5 day capacity) and gaseous oxygen (GOX - 20 min capacity) back-up supplies are included. These supplies can be replenished via the ASU during normal operation.

The air-blown system is believed to require four gasifier vessels against three for an equivalent oxygen-blown case, due to the greater volume of fuel-gas resulting from air blowing. Vessel sizes are ultimately limited to what is practical for factory fabrication and subsequent transportation.

The air-blown gasifier system works in conjunction with a carbon burn-up cell and bed ash (LASH) sulphator, which adds to the operational complexity, though also confers operational flexibility (especially in the case of the ABGC with its separate CFBC).

It is likely that a similar carbon burn-up facility would have to be included in any future oxygen-blown fluidised bed gasifier system in order to minimise lost carbon. This additional complexity is not required by slagging gasifiers such as the entrained flow and moving bed types.

Variability in limestone supplies, affecting sulphur capture performance, is a liability of the air-blown system, though the inclusion of secondary capture via, eg downstream zinc based sorbent, is seen as a stabilising influence.

Fuel-gas Treatment

After the fuel-gas leaves the fluidised bed gasifiers both systems use a fire-tube heat exchanger for primary fuel-gas cooling after passing through a cyclone for coarse fines removal. Thereafter, the air-blown and oxygen-blown modes differ.

The air-blown gasifier, by virtue of already achieving >90% sulphur removal via in-bed limestone feeding, logically uses a hot gas clean-up arrangement to avoid the unacceptable efficiency penalty of cold (wet) gas cleaning. The hot gas clean-up consists of a ceramic element particulate filter, chloride removal in fixed beds of nahcolite ($NaHC0_3$) and sulphur polishing in regenerable fixed beds of zinc ferrite. Four hot gas clean-up systems are required, one per gasifier train.

The oxygen-blown gasifier, as is often the case with oxygen-blown gasifiers of all generic forms, uses a cold, wet gas, clean-up system. Wet clean-up processes are chosen for historical reasons and because they are fully proven, conferring greater commercial confidence. In this instance it consists of the Selexol (acid gas capture), Claus (recovers elemental sulphur from the acid gases) and SCOT (tailgas treatment) processes. One system would cater for all three oxygen-blown gasifiers.

The wet gas processes, although commercially available, are complex and perform better on a constant flow basis, rather than in interrupted, load changing, operation. Wet gas sulphur removal has difficulties if the coal sulphur content varies, ie if a plant has to cope with coal sulphur variations of, say, 0.6 to 3.0%, together with up to 50% turndown, the acid gas flow to the Claus process would vary by a factor often. Thus, a wet process would better suit base load GCC operation.

The study team envisaged a bias against wet gas clean-up from the electrical utility industry. They anticipated it would be perceived as unfamiliar chemistry, with liquid (rather than more conventional solids) handling systems having leakage possibilities and requiring operators with chemical plant background experience.

The hot gas system, by virtue of being dry and based upon the interaction of gases with solids, was thought would be more familiar and acceptable to the electricity industry. However, reliability of components, particularly the ceramic filter, remains a major concern, and the durability and long term performance of sulphur sorbents is still unknown.

Much work is in progress around the world on the topic of hot gas clean-up of HCl, H_2S and other gaseous contaminants from coal-based GCC[64,65,66] and the best choice of materials and methods is still uncertain[67].

Gas Turbine

A key development item is the fuel-gas valve which controls the feed of fuel-gas into the gas turbine combustor. Material concerns limit the allowable temperature and the commercial limit is currently around 320°C, though ultimately 650°C may become possible (the ABGC development currently applies a 600°C limit to its fuel-gas clean-up). The oxygen-blown cold gas clean-up system results in a low fuel-gas temperature that presents no difficulties in this area.

Another area where the study team felt there was still uncertainty was the ability of low calorific value air-blown fuel-gas to be compatible, under all gasifier transient conditions, with the demands of an advanced gas turbine. This was considered to be a particular concern if the fuel-gas heating value fell below 3.7MJ Sm^{-3} (LHV; the ABGC fuel-gas has a calorific value between 4 and 5MJ Sm^{-3}). The acceptability of oxygen-blown (medium calorific value) fuel-gas raises no such questions, though their dilution with nitrogen or steam, for NOx control, substantially reduces oxygen-blown fuel-gas heating values.

LOAD FOLLOWING CAPABILITY

Part-load operation and the ability to follow changes of electrical load without significant loss of gasification efficiency are important characteristics for systems which must meet the operating requirements of electricity utilities. Flexible operation is becoming increasingly important in market driven power systems. The Central Electricity Generating Board's (CEGB's) specification for plant load changes was 5% per minute between 50% and 100% load and 3% per minute between 30% and 50% load[68]. New pf coal-fired power plants are able to satisfy or exceed these specifications. GCC plants should aim to equal or exceed the capabilities of pf-firing.

There are two significant aspects relevant to the comparison of air-blown and oxygen-blown systems:

i. the response of the oxidant supply system (booster compressors and ASU);

ii. the response of the gasifier itself.

The most significant appears to be the oxidant supply system. There is no reason to think that the use of air or oxygen per se should affect the response of the gasifier. Plants with cold gas cleaning may be less responsive due to the large number of process units, which result in longer time delays.

In fixed and fluidised bed gasifiers there is a substantial fuel inventory in the gasifier. This means that short term increases in output can be achieved independently of the response of the coal feeding system. In entrained flow gasifiers the fuel inventory is very small, so the fuel and oxidant feed rates both have to be changed in order to achieve a change in fuel-gas output. A further consideration with dry feed entrained flow gasifiers, such as those by Shell and Krupp-Koppers PRENFLO, is that the CO_2 and H_2O contents in the gasifier product are usually small (<2%), so that oxygen and coal feed rates both have to be varied at the same rate to avoid carbon production.

Integration of the ASU in an IGCC plant, with inlet air taken from the gas turbine compressor, has a significant effect on the time response of the fuel-gas production plant, through both oxygen purity and the flow of dilution nitrogen. However, these interactions do not seriously impair the achievable quality of IGCC load-following control[69]. Possibly more significant is the ability to cope with upset conditions.

In both a highly integrated IGCC and a non-integrated GCC, a gasifier 'trip' upsets both the oxygen plant and the power generating equipment. In a non-integrated scheme, as at Cool Water, the oxygen plant and power generating equipment upsets could be handled, although often with some difficulty, and the power equipment could stay on-line burning an alternative fuel. However, in a highly integrated IGCC scheme, oxygen plant and power generation equipment upsets can reinforce each other, leading to a shutdown of the entire facility[70]. Non-attributable reports from Buggenum indicate that some of their operating difficulties have been due to the high degree of integration.

BOC have stated that the increased operational complexity brought about by integration of an ASU cannot be addressed adequately by the traditional decentralised gasification combined cycle control strategy. The conventional view of a GCC as separate power and gasification 'islands' may not be appropriate as the scope of integration increases. It is their belief, however, that operational complexity can be adequately addressed by a plant wide multivariable based control system which can deal with the interactions and constraints effectively[71].

Published information on the load following capabilities of GCC plants is scarce. The Shell oxygen-blown IGCC process is able to operate at steady state down to 50% of plant capacity and to accommodate step changes of 5% in the fuel-gas demand in 30s. In response to changing power demand the control system is able to accommodate ramp changes in fuel-gas demand at a rate of 3% per minute between 100% and 50% plant capacity[72]. This satisfies the specification set by the operator at Buggenum but does not meet the CEGB specification.

The Texaco GCC gasifier at Cool Water achieved load change rates of 20% at a ramp rate of greater than 4% per minute. Tests at the HTW demonstration plant achieved load changes of 8% per minute between 60% and 100%[73].

Dynamic modelling studies on the ABGC have shown that load changing rates of 3% per minute could be achieved by the fuel-gas side of the process[74]. Also, 5% per minute could be achieved but a greatly oversized air booster compressor would be necessary. Further modelling and system development work may enable the dynamic response to be improved. The dynamics of the steam side have not yet been modelled.

A similar load changing rate of 3% per minute is quoted for a commercial GCC plant based upon the Japanese two-stage air-blown entrained flow gasifier[25].

FUTURE DEVELOPMENTS

A topic that might influence the debate on relative attributes of oxygen and air-blown gasification is the subject of future developments; certain prospects could favour one of the technologies more than the other.

Gas Turbine Cycle Improvements

Since a greater proportion of overall GCC electricity generation comes from the gas turbine cycle with oxygen-blown gasifiers (Section 1.2), any improvement in gas turbine cycle efficiency is likely to benefit oxygen-blown gasifiers more than air-blown designs.

The basic idea of the gas turbine (compression, combustor, expansion) was presented in a patent application in England in 1884. As early as 1904 gas turbines were in operation in the USA and Europe. Growth in the technology has been due to metallurgical advances and design sophistication, ie blade cooling, which has made possible the use of higher temperatures and pressures, thus improving efficiency[75]. Since the world's first combined cycle power plant was installed in 1956 at Dudelange by ABB, with a cycle efficiency of 38%, there has been steady development of the technology. Efficiency rose to 46.6% in 1978, 52% in 1989 and 55.5% in 1995[76].

For the simple open cycle arrangement, with exhaust gas heat recovery, the following factors influence design improvements:

- Increasing the turbine firing temperature increases specific power. Specific power equates to kW_e of generated power per kg s^{-1} of air flow through the turbine. The higher this value, the smaller is the gas turbine required, with consequent lower turbine capital cost.

- Increasing the pressure ratio improves thermal efficiency. The importance of this factor is self-evident in reducing turbine operating costs.

- The pressure ratio resulting in the maximum specific power and maximum efficiency, changes with the turbine firing temperature. The higher the pressure ratio, the greater is the advantage to be gained from an increased firing temperature.

Thus, the trend is towards higher values of firing temperature and pressure ratio. Table 21 compares the performance of gas turbines from several leading manufacturers.

GE are developing a gas turbine with a closed-loop steam cooling system in place of conventional air cooling - this gas turbine is designated the 9H. When conventional air film cooling is employed it lowers turbine temperature by injecting coolant into the hot gas stream and also disrupts the flow. Both factors reduce efficiency and output. The GE 9H design uses closed circuit steam (readily available in a combined cycle system) to more efficiently cool the critical gas turbine parts, while not diluting or disrupting the working fluid. The inlet temperature of the first stage of the turbine therefore operates 110°C hotter, with no increase in initial firing temperature, while the turbine blade material is protected by its internal steam circulation. The steam is returned to the reheat stage of the steam cycle.

As can be seen from Table 21, the most efficient natural gas fired combined cycle plants today run at 55% efficiency, with 58% promised when the latest generation of turbines from various manufacturers come on-line during the next 12 months. The GE H class machine, by virtue of its 1,430°C firing temperature and newly developed thermal barrier coatings for the first of the four power turbine stages, is claimed will break the 60% barrier[77].

The intercooled aero-derivative (ICAD) gas turbine is another advanced gas turbine design undergoing development[78]. The compressor intercooler reduces the temperature of the power turbine's cooling air, permitting firing temperatures up to 1,480°C with commensurate natural gas fired combined cycles of 60-63% (LHV).

Other gas turbine technologies exist[7]. Examples are:

- Steam-Injected Gas Turbines (STIG) replace the steam turbine cycle altogether. The steam is injected directly into an appropriately designed gas turbine, perhaps 3.6kg per kg of fuel-gas, which is more than five times the amount normally considered for NOx suppression. Although dispensing with the steam turbine, STIG suffers from a lower thermal efficiency than conventional combined cycle systems. Boiler feed water costs are high because of the once-through nature of the process and because steam quality must be high. Little interest has emerged in STIG for coal gasification processes.

- The Humid Air Turbine cycle (HAT) eliminates the use of an exhaust flue gas heat recovery steam generator (HRSG). In the HAT cycle, a flue gas recuperator replaces the HRSG and preheats both humidified combustion air and water. The turbines air compressor is both inter-cooled and after-cooled, with recovered heat preheating additional water. The hot water humidifies the combustion air in a multi-stage, counter-current saturator. The HAT cycle is sold commercially by Westinghouse Electric Corporation as the Cascaded Humidified Advanced Turbine cycle (CHAT). Studies have investigated its integration with Texaco and Shell oxygen-blown gasifiers[79]. Such integrated gasification plants are termed IGCHAT. Claims are made for a 15% reduction in the cost of electricity compared to conventional GCC due to significant capital cost reductions.

The ultimate advantage of an improved gas turbine efficiency will depend upon the ratio of electrical power generated by the gas turbine versus the steam turbine. As described in Section 2.1, the ratio of gas: steam turbine power varies from 2:1 with natural gas firing, to between 1.8 and 1.1:1 with oxygen-blown designs, to 0.8:1 with the air-blown ABGC. Even if oxygen-blown gasification is credited with the highest gas:steam turbine electrical generating ratio, ie 1.8:1 as the BGL process, then an increase in gas turbine combined cycle efficiency from the current 55% to the medium term 60% would bring a relative cycle efficiency gain of <1 percentage point over the ABGC. It would require a gas turbine increase from 55% to 69% to totally offset the more normally found 2 percentage point advantage to the air-blown fluid-bed gasification process. This is optimistic, both in terms of the required gas turbine performance and also the fact that even the 69% figure only applies to gasification schemes where the ratio of gas turbine to steam turbine power is at the 1.8:1 maximum. It also pre-supposes that there will be no equivalent improvement in the steam turbine cycle efficiency.

Steam Turbine Cycle Improvements

Within the UK there has been a steady improvement in station (net) thermal efficiency of steam cycle power generation, with values increasing from 26% (LHV) for $30MW_e$ plant in the late 1940s to over 38% (LHV) for present day stations equipped with $660MW_e$ turbines. These figures are reliant upon an increase from 34% to 44.5% in steam turbine heat rate efficiency, ie based upon steam turbine electrical output versus steam heat input, ignoring boiler efficiency losses and station electricity usage. This improvement has derived from higher steam conditions, which have advanced from 41bar, 454°C to 159bar and 566/566°C (superheat/reheat). Reheat was introduced with $120MW_e$ sets at the end of the 1950s, using 103bar steam and 538/538°C, giving station efficiencies up to 36%[80].

Pioneering plants in Europe and the USA have operated with significantly higher steam temperatures and pressures; the first boiler plant with a steam supply temperature in excess of 600°C was commissioned in 1951. By the mid-1950s some boilers were operating with supercritical steam conditions (pressures above 220bar_g). The largest such plant was that at Eddystone in the USA, built in the late 1950s, which used a steam temperature of 650°C at a supply pressure of nearly 350bar.

The adoption of such advanced steam conditions presents greater design problems today, since modern day steam generators are sized to give considerably higher steam outputs and need to be capable of providing other than base-load operation. They are also constructed using membrane furnace wall technology, rather than being of brick set open tube construction, and are considerably taller. These factors cause additional boiler design stresses which makes the limits of 280bar, 560°C/580°C preferable[81]. Such parameters provide a steam turbine heat rate efficiency of 45-46%

Continued development of the basic steam cycle power generating system is taking place in Japan and Denmark[21]. While the new Japanese plants use predominantly natural gas, with some coal firing, the Danish plants are substantially designed for pf coal firing. The Danish work has been reported in the literature[82,83]. The Danish ELSAM programme started in 1984 with the 352MW_e Studstrup unit 3, followed by the 384MW_e Fynsvaerket unit 7 in 1991 and the 383MW_e Esbjerg unit 3 in 1992. The commissioning of Alborg (Nordjyllandsvaerket unit 1) is due in 1997 and a station efficiency approaching 48%, including FGD, is expected.

Table 22 shows the efficiencies expected to be achieved by the Danish development programme to the year 2015. Cycles make use of double reheat, low stack temperatures (105-110°C) and low condenser pressures (0.015-0.018bar_a, 5-10°C). Double reheat produces an efficiency improvement of around 2 percentage points, but entails the penalties of additional capital cost and control complexity. The current use of such plant is for base load operation.

Although attainment of such high steam cycle efficiencies as in Table 22 requires the application of sophisticated alloy steels for parts of the boiler construction as well as the turbine rotors, the Danish experience suggests these problems should be no more insuperable than the equivalent difficulties encountered in development of advanced gas turbines. Major development projects are underway, particularly in Denmark, Germany, Japan and the USA, investigating advanced materials. One material, recently introduced, is P91, a high strength ferritic alloy. With the adoption of this and similar materials, the current steam temperature maximum is about 593°C, with 350bar and 650°C already realistically foreseen.

An alternative approach to improving the Rankine steam cycle, which places less emphasis on development of higher pressure/temperature components, is that based upon the work of Dr Kalina. The Kalina Cycle achieves a 15% improvement on what is attainable by the equivalent steam cycle turbine. When operating at temperature and pressure levels within current utility practice (165bar_g and 565°C) a heat rate of 48.5% is claimed[84]. Power generation by the Kalina cycle requires a boiler-turbine system as well as a system to distil and condense the working fluid. The working fluid is a mixture of ammonia and water which displays variable temperature boiling, depending upon the proportion of the two components. By using fluid mixtures it becomes possible to transfer heat from low to high pressure, a process which cannot be achieved using pure component fluids.

The boiler-turbine system is similar to that used in a conventional steam cycle, but the condenser is replaced by the distillation-condensation system. The distillation equipment facilitates production of an approximately 50:50 mixture of ammonia and water from an initial 70:30 mixture. The 50:50 mixture is the boiler and turbine working fluid, which is expanded through the turbine to a pressure whereby it cannot be directly condensed at ambient temperature. To condense the mixture it is necessary to mix it with a water-rich stream from the distillation system, which re-produces the initial 70:30 mix. This mixture ratio can be completely condensed. The 70:30 liquid mix is then pumped to an intermediate pressure and heated sufficiently to partly vaporise the ammonia component, which is then mixed with part of the remaining water-rich stream to re-produce the 50:50 boiler-turbine working fluid again.

From the foregoing, it is clear that the long term prospects for improvements in cycle efficiency within a combined cycle power generating plant is equally possible by enhancements to the bottoming cycle as to the gas turbine topping cycle. The ABGC should not, therefore, despite its greater reliance upon the bottoming steam cycle, suffer any developmental disadvantage. Indeed, because its CFBC provides greater flexibility on the choice of steam cycle parameters, the ABGC is in a better position to obtain the best advantage from developments in both the steam and gas turbine power cycles.

Fuel-gas Cleaning System

At the current stage of GCC plant development, the various proprietary forms of cold gas clean-up system are considerably more commercially proven than dry, high temperature, processes.

Purveyors of oxygen-blown gasifiers have thus far accepted the cycle efficiency disadvantage of cold gas cleaning as being less disadvantageous than working with system components which have had no previous commercialisation. Cold gas cleaning also confers higher environmental standards than hot gas cleaning at the present stage of development; residual environmental emissions and solid waste for disposal are both higher from the hot gas process. With the British Gas/Lurgi process, fuel-gas leaves the gasifier around 500°C and is laden with tars and heavy hydrocarbons. These factors necessitate use of cold gas cleaning with this oxygen-blown GCC system.

One of the incentives for development of air-blown fluidised bed gasification is the ability for low cost desulphurisation, with little efficiency penalty, through feeding limestone into the fluidised bed. With this ability, there is little justification in adopting cold gas scrubbing for H_2S removal from the fuel-gas, so designs have evolved on the basis of merely capturing particulates, to enable the fuel-gas to be fired within the gas turbine. More stringent environmental requirements are now leading to advanced hot gas cleaning processes for H_2S polishing and nitrogen compounds removal.

The adoption of hot gas cleaning has given the air-blown fluidised bed gasifier a cycle efficiency advantage, compared to oxygen-blown gasifiers equipped with cold gas cleaning. Although estimates vary of the efficiency penalty suffered by the various GCC concepts by adopting cold gas instead of hot gas filtration (Section 3), it is evident that the hot gas cleaning concept provides the ABGC gasifier with a gain of two to three percentage points when comparing full conventional cold gas cleaning with mere hot gas removal of particulates. This advantage, together with the facility of high efficiency steam cycle parameters afforded by the CFBC, plus the low auxiliary power consumption of air blowing, combine to elevate the cycle efficiency of the ABGC above its oxygen-blown rivals. The CFBC component also provides an efficiency advantage against other air-blown fluidised bed gasifiers, such as the KRW gasifier, but at the expense of additional operating complexity.

The disadvantage to the use of hot gas cleaning, especially when this term refers only to collection of particulates, is the relatively high emission level of gaseous pollutants. In this context, the ABGC's char combustor is a disadvantage, since no matter how efficient hot or cold gas cleaning systems become, there will always be a base line emission level originating from the CFBC. Table 23 compares anticipated emissions from an ABGC system[36,59] equipped with 'simple', particulates only, hot gas filtration and advanced hot gas filtration. A comparison is also made with cold gas cleaning, applied to a Shell gasifier. Results from the New University of Ulster are averages from ten coals.

The higher level of CO_2 release from the Shell gasifier with cold gas cleaning, g kWh^{-1} (see Table 23), reflects the lower cycle efficiency of the process.

In the medium term, as environmental standards from dry, hot gas cleaning methods approach more closely those attainable from wet, cold cleaning processes, and as reliability of hot processes are demonstrated in commercial prototypes, it is likely that hot gas cleaning will become more widely adopted by competing oxygen-blown GCC processes, as they seek to maximise their cycle efficiency. This will reduce the cycle efficiency advantage of the ABGC. Further advantage will be lost as environmental standards force the early adoption of advanced hot gas cleaning in order to meet ever more stringent Western emission standards; advanced cleaning will impose higher capital and operational costs compared to the original ABGC concept.

The ABGC (and similar air-blown fluidised bed gasifiers) might find an advantageous marketing niche in parts of the world, such as India and China, where environmental emission legislation is less onerous than in the West, but where the rate of economic development is calling for considerable increases in coal-fired power generating capacity. The 'simple' ABGC is of low specific cost, is high efficiency, captures >90% sulphur and inherently controls NO_x emissions to 400-450mg Sm^{-3} (at 15% O_2). These factors should make it preferable to pf boilers equipped with FGD and low-NOx burners.

9.4 Plant Integration

The two major disadvantages to oxygen-blown gasification are the cost and power consumption of the ASU. Section 4 has described how the cost of an ASU for GCC applications has halved, in real terms, during the period since 1980. Parasitic power consumption has also reduced, both by design and by the adoption of the ASU into an IGCC concept. Several of the studies referenced within this report anticipate that the highest cycle efficiency will ultimately come from oxygen-blown gasification with a fully integrated ASU and equipped with a hot gas clean-up arrangement. Such design concepts are already being studied[6].

As operating experiences from demonstration plants such as Buggenum and Puertollano generate a practical understanding of operational control of fully integrated plants, it is likely that full or partial integration of the ASU with the gas turbine will eventually become the norm with oxygen-blown gasifiers. However, information from Buggenum suggests that full integration has a significant impact on plant complexity, particularly during the start-up sequence, and equally impacts on plant availability. After 33 months of commissioning only 2,500h of coal gasification were recorded[85] and it is likely that only part of those hours were obtained in an integrated mode. Thus, although a theoretically attractive concept, with comparative studies indicating a 2-3% improvement in cycle efficiency and a reduction in capital cost per kW_e, general adoption of the concept into oxygen-blown GCC projects must still be many years away.

Only in the medium to long term, therefore, is there any risk that the ABGC will be overtaken in cycle efficiency by the fully integrated IGCC concept equipped with hot gas cleaning.

FUEL FLEXIBILITY

The University of Ulster has carried out a study[36] which explores the variations in gasifier performance to be expected from a range of coals traded in Europe for power generation. It seeks to optimise what GCC system would best suit a particular type of coal; alternatively, what coal is best suited to a particular GCC system. Initially 167 different coals were considered, from lignites to anthracites, but these were eventually reduced to ten coals which reasonably spanned the desired chemical component, and other characteristic, parameters. Table 24 shows minimum and maximum values of relevant parameters for the selected coals.

Of the many gasifier types considered within the study, the authors finally chose to concentrate their attention upon the air-blown British Coal fluidised bed gasifier used in the ABGC, the oxygen-blown BGL moving bed gasifier, the Shell entrained flow, oxygen-blown gasifier, and the Texaco entrained flow, oxygen-blown slurry coal feed gasifier. The Texaco gasifier's slurry coal feed is usually 60-65% solids by weight. The water in the slurry acts as a transport medium, a temperature moderator and a reactant.

The gasifiers were all considered with hot and cold gas cleaning systems (see Section 3 for a description) except for the BGL process, where the moving bed, counter-flow, gasification method produces a low temperature fuel-gas, around 550°C, which contains tars and heavy hydrocarbons. These require to be cold condensed at the gasifier exit (being recycled back to the gasifier), so making subsequent cold gas cleaning the only logical approach.

The study found that environmental performance and cycle efficiency varied considerably with the coal feedstock used, the type of gasifier system and the method of fuel-gas cleaning. Considering the individual gasifier concepts, the study reached the following conclusions:

Fluidised Bed Gasifier

- Caking coals are a potential disadvantage to a fluidised bed, as any tendency towards particle agglomeration can result in the loss of fluidisation. The same concern applies to the feed size of the fuel entering the gasifier, with the added concern regarding fine coal being carried unreacted from the gasifier freeboard. It is acknowledged, however, that the British Coal ABGC pilot scale gasifier has been successfully tested with a wide range of coals with BS Swelling Numbers up to 5½. The fuel injection design is critical to successful operation on caking coals.

- The U-Gas and KRW gasifiers require a controlled degree of bed ash agglomeration in order to encourage it to sink through the bed to a suitable extractor. This imposes an additional operating constraint which worsens fuel flexibility.

- The degree of conversion of the coal to gas depends upon the volatile content of the coal. In the case of anthracite coals, their very low level of volatile matter results in a low calorific value fuel-gas, possibly unsuitable for firing in a gas turbine.

- Low volatile coals place greater reliance upon the CFBC char combustor for achieving high thermal efficiency. This, however, worsens overall emissions since the CFBC contributes a significant proportion of the total pollutants emitted.

- The percentage fuel conversion to gas, and hence the amount of residual char going to the CFBC, are both related to the volatile content and char reactivity of the coal. However, the difference in fuel conversion for most power generation coals is small and so an existing plant, with fixed sizes of gasifier, CFBC, gas turbine and steam turbine, will be adaptable to switching between one coal type and another. The capacity required of the CFBC and its associated steam cycle, relative to the gasifier and its gas turbine, will have a small contingency to be able to accommodate the widest range of world traded coals.

- Limestone feeding to the fluidised bed generates a discharge of CaO, CaS and coal ash (LASH) which is less environmentally acceptable than the glassy coal ash slag coming from the slagging gasifiers. The higher a coal's ash content and sulphur content, the greater the amount of LASH requiring to be disposed. Of course, a higher ash content fuel fed to any combustion equipment will generate more solid waste for disposal.

- Superficially it would appear that for the ABGC, cycle efficiency is maximised with high volatile coals of low ash content and low C:H ratio. By relying more on the hydrogen component of the coal to form its fuel-gas, with the fixed carbon being more likely to end up in the CFBC, the efficiency is improved by a high hydrogen content fuel since more energy goes to the gas turbine cycle.

- Very low ash contents impose an efficiency penalty on the cycle because the mineral matter in the coal helps to cool the gasifier to its operating temperature; with low ash materials, this cooling duty needs to be performed by additional steam taken from the steam cycle. The mineral matter 'deficit' could be replaced with limestone in the case of low ash, but high sulphur fuels. The ABGC cycle efficiency is optimised when the gasifier conversion is ~80% (daf basis). This is because some fuel needs to be left for combustion in the CFBC to contribute to the steam cycle and to provide the environment for the CaS oxidation. Additionally, the British Coal fluidised bed gasifier operates with a char bed with no requirement for supplementary bed material. A higher conversion than the 80% would result in a difficulty maintaining a full bed inventory.

- When using the simplest ABGC concept, ie in-bed desulphurisation and downstream particulate removal by a ceramic filter, a coal low in chlorine, nitrogen and sulphur is advantageous.

- An inherently low sulphur coal makes least demands upon the less effective in-bed desulphurisation method. The same hydrogen sulphide level in the fuel-gas is achieved regardless of the fuel sulphur content, but the mass of solid waste generated is lower with low sulphur fuels.

- The fuel-gas fuel-nitrogen content, largely in the form of ammonia, increases with the nitrogen content of the coal fed to the gasifier. Clearly a lower coal-nitrogen content results in lower overall NOx emissions from the ABGC. A low chlorine content would reduce risk of alkali metal chlorides causing problems in the gas turbine.

Moving Bed, Oxygen-blown, Gasifier

- Coal selectivity can be a problem, as the gasifier may encounter problems with strongly caking coals or coals with a large proportion of fines, which would need briquetting. It is also unsuited to lignite coals.

- The study found a strong correlation between cycle efficiency and ash content of the coal. This is a direct consequence of slagging operation, where heat is wastefully used in melting coal ash. Furthermore, the slag which is removed from the gasifier is subjected to water quench with no heat recovery. This makes the overall system efficiency susceptible to a high ash content in the coal.

- The relatively low slagging temperature makes it necessary to add a fluxing agent to the fuel, usually 0.32kg limestone per kg of ash. Further energy loss, proportional to coal ash content, is wasted in melting this flux.

- Coals of high water content are not expected to impact on cycle efficiency, provided they can be fed reliably. The water merely absorbs latent heat of vaporisation as the coal moves downward within the gasifier, counter-flow to the hot, rising, fuel-gases. Thus, the coal's moisture pre-cools the exiting fuel-gas. The fuel-gas is subsequently quench cooled downstream of the gasifier, by application of cold gas cleaning; unavoidable for moving bed gasifiers because of the carry-over of tars and heavy hydrocarbons. Hence, the moisture in wet fuels could be regarded as a pre-cooling stage of the cold gas cleaning system.

Dry Coal Feed, Entrained Flow, Oxygen-blown Gasifier

- In the Shell configuration of gasifier, the crushed coal feedstock is pre-dried to around 5% moisture before entering the gasifier. This confers greater fuel flexibility, since all coals are brought to a more uniform physical condition. Although drying is a fairly energy intensive process, which exacts an efficiency penalty, this is more than recovered by the lesser amount of evaporative heat subsequently taken from the gasifier.

- High ash content coals require additional electricity consumption in feeders and crushers and also extract greater latent heat of fusion from the gasifier when slagging the ash. The problem is made worse by the addition of limestone flux, but less so than with the moving bed gasifier since the higher temperature in the entrained flow case reduces the dependence on flux to the lower level of 0.1-0.15kg limestone per kg ash.

Slurry Coal Feed, Entrained Flow, Oxygen-blown Gasifier

- The Texaco process exhibits low cycle efficiencies due to its slurry feeding system. It is most helped by coals of high C:0 ratio. Results in the literature show that coals high in carbon tend to produce higher levels of CO, whereas coals high in oxygen tend to produce more water vapour. With a surfeit of water vapour in the gasifier from the slurry feed, any additional amount does not help gasification and merely contributes bypass heat to the lower efficiency steam cycle.

- The Texaco process is penalised by high ash contents since the ash as well as the organic material also needs to be slurried and has water associated with it. This additional water takes no part in the gasification reactions, but merely burdens the gasifier which requires more combustion reactions to maintain operating temperature, resulting in a lower heating value gas and a larger fuel-gas volume for the same energy output.

- Coals of high surface moisture content delivered to the power plant would be least likely to cause a noticeable fall in efficiency, since the slurry feed effectively masks any comparatively small additional moisture quantities. However, a high inherent moisture would have a detrimental effect on cycle efficiency since inherent moisture is not available for the slurrying process.

- Particularly (physically) difficult coals are probably best gasified in a slurry fed system.

- Coal surface properties affect the ability to create a slurry from some fuels.

In general, it seems that all the GCC systems have a reasonable fuel flexibility which will suit most coals under most circumstances.

The Shell type, dry feed, oxygen-blown, entrained flow, gasifier is possibly the most flexible of all; by grinding the coal to a fine powder (100µm top size, 90% <90µm[86]), pre-drying it and gasifying it in an entrained stream, the Shell gasifier avoids possible coal caking difficulties and reduces physical property variations. A possible problem is with the grinding equipment when used with high ash coals where the ash is hard and abrasive. This will cause accelerated wear of the grinding equipment, necessitating larger capacity grinding mills, more spare parts and increased maintenance.

The Shell Oil Company have reported their own verdict on fuel property effects on their gasification process[87]. The basis was a design for a nominal $500MW_e$ fully integrated plant and considered five coals, from bituminous to lignite.

For most bituminous coals the Shell GCC performance was not highly sensitive to the coal selected. Capital costs of 1,260 to 1,330 mid-1991 US$ per kW_e were estimated for three bituminous coals, with cycle efficiencies (HHV) of 43.6 to 42.3%. In general, the higher sulphur coals produced the lower efficiencies. The lower rank coals noticeably reduced cycle efficiency and increased costs, blamed on higher moisture and ash contents. Texas lignite, with 17%) ash and *30.1%* moisture, reduced plant efficiency to 39.0% and increased costs to US$1,500 per kW_e.

The Texaco, slurry coal feed, oxygen-blown, entrained flow gasifier, grinds its coal to a coarse specification, so is less susceptible to mill wear. Texaco claim[5] to have ground, slurried and gasified coals from around the world in various Texaco commercially operating plants. They also claim to have gasified bituminous coals, sub-bituminous, lignites and petroleum cokes at their Montebello laboratory pilot rig.

Since the amount of slurry water fed to the gasifier increases with the ash content, high ash coals place a heavy evaporative load upon the gasifier. This imposes a significant deficit on the cycle efficiency, which increases the amount of coal needing to be gasified, which further increases the ash throughput. For this reason Texaco technology did not fair well in the comparative study which tried to identify the best GCC technology for high ash Indian coals[35]. The 'design' coal had an ash content of 35% and a moisture content of 18% (allowing for the monsoon season). Texaco advised that they would grind the coal to <420µm and form a 68% solids content slurry. They also elected to operate at high gasifier temperature to avoid injecting a fluxing agent, since this would have further increased the inerts loading and necessitated additional slurry water.

Despite making efforts to limit the amount of inerts entering the gasifier, Texaco found themselves limited by the capacity of their slag lock-hopper valve, which limited the capacity of each gasifier to 1,800 tonnes per day of coal. This necessitated the use of seven gasifiers in order to provide the total $600MW_e$ (net) electrical output. For comparison, Shell were able to offer four gasifiers while the air-blown KRW fluidised bed system required six.

Although open to criticism for its deleterious effect on cycle efficiency, the Texaco slurry system is a safe, efficient and cost effective way of providing the gasifier with a pressurised, milled coal supply, so cannot be simply dismissed. It also helps eliminate variations in coal characteristics, making it especially suited to gasifying 'difficult' coals.

Coal ash content was a recurring theme within the University of Ulster study. They found that coals with a high ash content had a negative effect on the efficiency of most of the systems, but this was particularly true of the (oxygen-blown) slagging gasifiers, especially when direct water quenching was used in slag solidification and disposal, with no heat recovery. The problem is particularly exacerbated when the ash has a high melting point, since slagging gasifiers often require a fluxing agent in order they can form a free-flowing slag.

The BGL gasifier is affected more by high ash contents than the others, since its lower operating temperature requires greater amounts of fluxing agent. Also the process is less flexible than the other oxygen-blown processes, being unsuitable for strongly caking coals or coals with a large proportion of fines. Ideally, it needs a graded coal supply, which might influence fuel costs. It is also unsuited to gasifying lignites. Its inherent inability to be developed in conjunction with a hot gas cleaning system might also impose a limit on its application in the medium term, if such a cleaning system becomes more widely adopted on other GCCs for the sake of enhanced cycle efficiency. Its strength is its high cold gas efficiency (Section 1.2), which makes it especially suited to the supply of town-gas or chemical by-products.

Air-blown fluidised bed gasifiers, such as the ABGC and the KRW system, are particularly advantaged by coals of medium to low sulphur, low chlorine and inherently low nitrogen content. These characteristics increase the possibility of limiting hot gas cleaning to simple in-bed desulphurisation and downstream particulate capture, a design option which maximises cycle efficiency and minimises cost, but sacrifices best possible emissions control. High ash coals exert a lower efficiency penalty than with slagging gasifiers, since the ash is not slagged (it is partly agglomerated in the KRW design) and so accounts for less heat loss. There is also less power consumed in coal preparation, since milling is usually only to <6mm. However, high ash coals will produce large quantities of bed LASH, with its uncertain environmental characteristics which might entail additional disposal costs. However, it is important to appreciate that the components which might entail additional disposal costs (eg, CaO, CaS if present) will exist in lower concentrations.

11. DISCUSSION

Oxygen-blown gasifiers, as currently applied to GCC power generation, derive mainly from earlier developments intended for manufacturing town-gas or chemical by-products. In this former role the need for a high 'cold gas' conversion efficiency, ie a high conversion of potential energy from coal to potential energy in fuel-gas, was paramount. Continued use of the same design concepts for GCC makes commercial sense, since it capitalises on the developer's prior design and operational experience. However, such gasifier configurations, and the use of oxygen blowing, are not necessarily best for GCC, where sensible as well as potential energy in the fuel-gas can be converted into electricity via the gas and steam turbine cycles. Nonetheless, with gas turbines having a higher cycle efficiency (~55%) than steam turbines (~45%), there is a theoretical advantage in maximising the conversion to fuel-gas total energy in order to maximise use of the gas turbine.

Air-blown gasifiers, by necessity of having to promote additional combustion reactions to heat the nitrogen of the air, have lower cold gas efficiencies than oxygen-blown gasifiers of the same type, which leads to lower gas turbine: steam turbine power ratios. This can largely be resolved by judicious choice of gasifier, since certain designs are more appropriate to air blowing while others more naturally suit oxygen blowing. It is incongruous to use air blowing with a single-stage entrained flow gasifier fed with coal in the form of a slurry. Such an arrangement must rely excessively on combustion reactions to evaporate and superheat the slurry moisture, heat the attendant nitrogen in the air and slag the coal ash. Similarly, oxygen blowing a low temperature, non-slagging, gasifier results in a surfeit of heat which has to be attemperated by steam or indirect cooling.

Sensible energy in the fuel-gas which enters the steam cycle upstream of the gas turbine, ie by fuel-gas cooling, is termed bypass heat. Bypass heat is inevitable in the design of coal-fired GCC processes, particularly in the case of (i) gasifiers which operate at a high fuel-gas exit temperatures, such as single-stage, entrained flow, ash slagging, designs, where much fuel-gas cooling is required at the gasifier exit, and (ii) gasifiers which fail to achieve complete carbon conversion and so require a separate combustion stage for the residual char, as is the case with single-stage, low temperature, fluidised bed processes. As bypass heat increases, the ratio of gas turbine: steam turbine power generation falls.

With natural gas-fired combined cycle power generation the gas turbine:steam turbine power ratio is around 2:1, but with all GCC plants the ratio is lower. Typical figures for oxygen-blown gasifiers are 1.8:1 for moving bed, 1.6:1 for two-stage, entrained flow, 1.3:1 for single-stage entrained flow receiving a dry coal feed and 1.1:1 for single-stage entrained flow receiving a slurry coal feed.

The lowest gas turbine:steam turbine power ratio, around 0.8:1, occurs with the British Coal (air-blown) ABGC gasifier. Char, resulting from incomplete carbon conversion in the gasifier, combusts to bypass heat within a purpose designed CFBC boiler and so enters the steam cycle. Because the CFBC operates at (metallurgically) benign oxidising conditions of 900°C, it enables the use of high efficiency steam cycle parameters, rather than the steam cycle being compromised by gasifier conditions. Thus, although more power is generated by the ABGC's steam cycle than would have been the case with an oxygen-blown gasifier that left no residual char, the ABGC's steam cycle compensates by operating at a higher efficiency. This efficiency benefit and the ABGC's lower parasitic power consumption compared to an oxygen-blown gasifier with its ASU, more than compensates for the apparently less advantageous gas turbine:steam turbine power ratio.

All GCC process designers strive to optimise the overall gas turbine plus steam turbine cycle efficiency, but are constrained in that optimisation by process requirements and constructional limitations. The flowsheet variations adopted by GCC designers to overcome these limitations, and the general complexity of GCC schemes, makes it difficult to anticipate what arrangement will achieve the ideals of highest efficiency, lowest cost and lowest electricity price. Resolving the question whether to air blow or oxygen blow GCC gasifiers is further confused by the wide variety of gasifier configurations. Some use a dry coal feed while some use a slurry. Then some designers opt for the more novel, but more efficient, hot (dry) fuel-gas cleaning process, while others stay with the long proven cold (wet) fuel-gas cleaning method.

Other factors which also enter the debate are:

- Integration of the ASU with its oxygen-blown gasifier improves cycle efficiency, by reducing parasitic power consumption, and reduces plant costs. However, an integrated ASU also increases operational complexity, which impacts upon plant start-up times and general availability.

- Certain gasifier types and fuel-gas cleaning methods are already commercially available, while others are still developmental.

- Environmental considerations may favour certain GCC configurations more than others by virtue of stack emission, solid waste and liquid effluent disposal legislation.

With so many factors influencing the final decision on what is 'best', it is possible to bias comparative studies, knowingly or unknowingly, in a certain direction by means of the ground rules of the study. Thus, protagonists of oxygen-blown gasification can make a strong case for their technology by drawing comparisons based upon a gasifier design which is inappropriate for air blowing. In order to make an impartial judgement on the relative merits of air-blown versus oxygen-blown gasification, therefore, this report is not based upon a new study carried out specifically for its preparation. Instead, the public literature has been studied and all relevant information has been extracted and compared.

Studies that enable comparisons to be drawn between air-blown and oxygen-blown gasification were found to have been carried out by the IEACR (UK), CRIEPI (Japan), EPRI (USA), RWE Energie AG (Germany), CSIR (India) and the University of Ulster (N Ireland). As expected, no one study covers all the possible gasifier technologies and gas cleaning methods. Indeed, some are very specific to just one gasifier type (which might be more appropriate to one blowing medium than the other). Further differences in the studies include the gas turbine and steam turbine cycle parameters, plant total electrical output and degree of residual pollution control. All of these variations mean that the studies cannot be directly compared, one with another, but must be viewed independently. Collectively, however, they illustrate a trend between air-blown and oxygen-blown GCC processes.

Efficiency Variations between Air-blown and Oxygen-blown Cycles

Table 25 compares cycle efficiency values for all the air-blown versus oxygen-blown studies. The stated cycle efficiency value is the highest found in each study, while the other values are reductions in efficiency (in percentage points) that are reported for the competing technologies. It is worthy of note that all but one of the maximum values appear in the air-blown side of the table, while the solitary maximum on the oxygen side (the Krupp-Koppers PRENFLO study) draws advantage from the type of gasifier technology and from its ASU being fully integrated with the gas turbine. While the balance appears strongly to favour air blowing, the various factors mentioned above can influence which technology delivers the highest cycle efficiency. Thus, a fair comparison requires the studies to be normalised with respect to these factors before the true effect of air blowing or oxygen blowing can be seen.

Information from within Table 25 has been extracted and compared in order to isolate the advantages conferred by hot (dry), rather than cold (wet), fuel-gas cleaning (see Table 26), and the use of a dry, rather than a slurry, coal feed (see Table 27). Elimination of both these factors is necessary in assessing just the relative merits of the blowing medium. The choice of hot or cold fuel-gas cleaning is one which is applicable to all gasifiers (apart from with the moving bed design, which needs to use cold cleaning). The choice of whether to use a dry or a slurry coal feed, however, is not a variable which affects all designs. It is specific to certain gasifier developers (mainly Texaco and Destec) who have made it an intrinsic part of their technology.

Table 26 assesses the cycle efficiency deficit which accompanies changing from hot to cold fuel-gas cleaning, while other technology options are kept constant. Each value in the table reflects the disadvantage of moving from hot to cold fuel-gas cleaning. Values indicate that air-blown gasification is disadvantaged more by cold gas cleaning than is oxygen-blown gasification. This is logically what would be expected, since cold gas cleaning air-blown derived fuel-gas entails losing sensible and latent energy from a larger fuel-gas volume.

Table 26 also attempts to distinguish between two levels of hot fuel-gas cleaning. 'Simple' cleaning refers to fluidised bed gasifiers using in-bed limestone for fuel-gas desulphurisation followed, downstream, by just ceramic filter particulate capture. 'Advanced' cleaning allows for the further efficiency loss which accompanies additional fuel-gas processing. In the extreme this might include HCl gas removal, sulphur capture polishing (with or without in-bed limestone) and gaseous nitrogen compound removal.

Comparing advanced hot gas cleaning with cold gas cleaning for a dry coal feed air-blown gasifier, the loss is a near consistent 1.8 percentage points. The only data for cold gas cleaning fuel-gas from a slurry coal fed air-blown gasifier shows 2.6 percentage points loss. The equivalent oxygen-blown results show a relatively wide spread, with apparent sensitivity to coal type. This might equally apply to air blowing, but sufficient data are not available to evaluate this. For the oxygen-blown, dry coal feeding case, a typical value of 0.8 percentage points advantage to hot gas cleaning will be presumed within this study (the >2.3 percentage point claimed in one literature source'[401] is considered very speculative). When oxygen blowing a slurry coal feed, a typical advantage of 1.6 percentage points to hot gas cleaning will be assumed (the IEACR information is suspect, being surprisingly low compared to values from the other studies).

The greater efficiency loss which occurs when cold gas cleaning fuel-gas from a slurry coal fed gasifier reflects the greater amount of water vapour within the fuel-gas. During cold gas cleaning, water vapour condenses and the latent heat bypasses to the lower efficiency steam cycle. With hot gas cleaning, the water vapour remains uncondensed and expands through the gas turbine. Thus, the greater the percentage water vapour in the fuel-gas the greater is the amount of bypass heat which condenses during cold gas cleaning and enters the steam cycle.

The University of Ulster anticipate the ABGC to be disadvantaged by 2.4 percentage points (% LHV) when changing from 'simple' hot gas cleaning to cold gas cleaning. Therefore, compared to the typical 1.8 percentage point advantage of 'advanced' hot gas cleaning over cold gas cleaning, the advantage of 'simple' compared to 'advanced' hot gas cleaning is around 0.6 percentage points. A current CTDD study suggests a deficit to the ABGC of 0.4 percentage points when moving from 'simple' to 'advanced' hot gas cleaning. In recognition of both studies, an average 0.5 percentage point value will be presumed.

Table 27 indicates the effect of moving from dry coal feeding to slurry coal feeding while keeping other technology aspects constant. There is evidence that the deficit to slurry coal feeding becomes greater when the resulting fuel-gas is subjected to cold, rather than hot, clean-up. There is no strong evidence, however, that the effect is different between air-blown and oxygen-blown gasification. Thus, within this study it will be presumed that the use of slurry coal feeding, compared to dry coal feeding, both with hot fuel-gas clean-up, causes a deficit of 2 percentage points. The use of slurry feeding compared to dry coal feeding, both with cold fuel-gas clean-up, causes a deficit of 2.6 percentage points,

ASU Costs, Development and Integration

Oxygen-blown gasification has benefited from the development, over the last 10 to 15 years, of ASUs specifically suited to GCC installations. Various ASU suppliers have worked with gasifier developers to engineer cost effective and power saving versions of their ASUs, cutting away unnecessary peripheral equipment that is normally included for the co-production of argon, krypton, xenon and neon. The purity of the oxygen has also been allowed to fall, often to 95% but also, as at Puertollano, to 85%.

A 1980 ASU design typically produced 98% purity oxygen and had a capital cost of US$28,200 per US ton per day (corrected to 1994 US$) of oxygen produced. A 1994 ASU having an oxygen purity of 90-95% has been estimated to cost almost exactly a half that amount while consuming about 10% less energy. The 1980 ASU was expected to consume some 367kWh per US ton of oxygen at 95% purity, hence the 10% saving anticipated for a modern ASU would reduce this to 330kWh per US ton.

The 1980 ASU might have inflated the costs of an oxygen-blown gasifier installation by 15-20%, whereas the modern ASU is more likely to inflate costs by 7-15%.

Parasitic power consumption of an ASU can vary, depending upon the manufacturer. Praxair (formerly the Linde Division of Union Carbide) suggest a modern ASU might consume 9-13% of the GCC gross system output. However, there is a trade off between capital cost and power consumption. A 1987 EPRI study found 348kWh per US ton for a BOC Cryoplants ASU with a capital cost of US$13,855 per US ton per day of oxygen, while an Air Products ASU consumed 311kWh per US ton but had a capital cost of US$22,165 per US ton per day of oxygen. However, the Air Products ASU was found to have the lower lifetime cost (US$10,300 per US ton per day), because of its low (6.1% out of a total of 9.5%) parasitic power and low cooling water consumption. Capital cost of this ASU was 13.5% of total plant construction costs.

Apart from ASU developments that have taken place to produce designs optimised to the needs of GCC plants, another technique which further lowers ASU costs and parasitic power consumption is the integration of the ASU with the gas turbine. Various forms of integration exist, viz. nitrogen only integration and air plus nitrogen integration. GCCs with an air integrated ASU are termed IGCCs within this report.

The air delivery and oxygen pressurisation compressors account for more than half the capital cost of an ASU. With a fully integrated ASU, receiving air from the gas turbine's compressor, the ASU does not require an air compressor, which itself represents over 1/3 the cost. Also, the oxygen compressor can be smaller because oxygen emerges at a higher pressure when the ASU receives air from the gas turbine rather than from a stand-alone compressor. The nitrogen from the ASU similarly emerges at a higher pressure and is viably fed to the gas turbine's combustion chamber where it boosts power output and also suppresses NOx formation. The cost of the integrated ASU is further reduced because the higher working pressure enables the ASU distillation vessels to be manufactured smaller for the same duty. The 85% oxygen purity integrated ASU at Puertollano contributes only 5.2% to total investment costs, including product compressors and back-up storage of nitrogen and oxygen. This percentage figure is misleadingly low, however, due to total costs at Puertollano being relatively high.

The Krupp-Koppers PRENFLO study in Table 25 is advantaged, compared to the other oxygen-blown gasifiers within that table, because its ASU is fully integrated with the gas turbine. The first commercial demonstration plant to use full air and nitrogen integration is the Shell gasifier currently being commissioned at Buggenum, The Netherlands; reports imply operational difficulties with this concept, which may slow down universal adoption of ASU integration.

In order to ascertain the cycle efficiency benefit of integrating the ASU with the gas turbine a further literature search was made for comparative studies. Table 28 shows the result, with data taken from the IEACR (UK) and four studies sponsored by the EPRI (USA). There is general agreement that cycle efficiency improves by around 1.1 percentage point (a 2-3% improvement) by integrating the ASU. IGCC plant costs are also felt likely to reduce by 2-3% by virtue of integrating the ASU.

Normalisation of Cycle Efficiency Variations

Table 29 summarises the cycle efficiency deficits taken from Tables 26, 27 and 28. These indices are applied in Table 30 to all of the study results previously shown in Table 25. Table 30, therefore, brings all the study results to a common format of hot (advanced) fuel-gas cleaning and dry coal feeding, such that air-blown and oxygen-blown gasification cycle efficiencies can be directly compared without confusion from other technology options.

Table 30 clearly shows the benefits of air blowing for the best cycle efficiency. When considering the data in Table 30, it is proposed to disregard the SCS KRW study (a detailed study but fundamentally flawed by the omission of a carbon-burn-up cell in the oxygen-blown case) and the high ash Indian coal study (biased against entrained flow, ash slagging, gasifiers and especially against the slurry coal feed Texaco design due to its use of quench fuel-gas cooling).

Taking a simplistic approach to the data in Table 30 and simply comparing the IEA, CRIEPI, HTW and University of Ulster studies, the air-blown gasifiers have an average cycle efficiency around 46.3% when designed for dry coal feeding and advanced hot fuel-gas cleaning (+2.2, -2.0 percentage points), while the oxygen-blown gasifiers average 44.5% on the same design basis (+2.1, -1.7 percentage points). Thus, using this simple method of analysis, air-blown gasifiers would seem to offer 1.8 percentage point advantage over oxygen-blown gasifiers. Table 29 shows that 1.1 percentage points of this advantage could be regained to oxygen-blown gasification by the adoption of an integrated ASU (IGCC).

The highest cycle efficiency within Table 30 is that reported in the Krupp-Koppers PRENFLO study. However, the figures in both the air-blown and oxygen-blown cases are suspiciously high, especially considering that a single-stage entrained flow gasifier is somewhat inappropriate for air blowing. In Table 28, the similar (but later) study comparing oxygen-blown Krupp-Koppers PRENFLO gasifiers with and without integrated ASUs, shows a cycle efficiency of 44.4% for the integrated case. Comparing this to the Table 25 result of 46.9% (common basis: cold fuel-gas cleaning) suggests that the results reported in Table 25 (and modified into the values in Table 30) may be in error by +2.5 percentage points. This would make the Table 30 air-blown Krupp-Koppers PRENFLO result a more typical 46% (basis: hot gas cleaning).

It is not clear why, in the IEACR study (Tables 25 and 30), the air-blown KRW suffers a -2.6 point deficit compared to the ABGC. This may partly be accountable to the ABGC having a higher efficiency steam cycle and partly to it having a simpler form of hot gas cleaning; the IEACR study does not make clear exactly what hot gas cleaning system has been modelled with the ABGC. The oxygen-blown KRW result shows a -3.2 point deficit, though 1.8 points of this would be expected via the analysis described above. This leaves 1.4 percentage points unaccountable deficit, compared to the ABGC.

It would be expected that fluidised bed gasifiers (air-blown or oxygen-blown) would have a small efficiency advantage over entrained flow, ash slagging, designs due to the latter's greater heat loss to sensible and latent heat in the emergent molten slag. Within the sparse data available, this is not evident, though the Indian coal study shows that for very high ash coal a noticeable difference should appear.

Table 29 shows a 0.5 percentage point advantage to the ABGC (and probably the KRW) when equipped with 'simple' hot fuel-gas clean-up. This is additional to the 1.8 percentage point advantage of air-blown compared to oxygen-blown gasification, as deduced above. The acceptability of such an arrangement is, however, susceptible to tightening environmental regulations.

Environmental Considerations

Aggressive marketing claims from competing technologies using cold gas cleaning are driving down legislative emission targets in the UK and the rest of Europe. Although air-blown GCC is consistently favoured in terms of cycle efficiency, plant costs and electricity generation costs, it generally achieves these using hot fuel gas cleaning. Current hot fuel-gas cleaning technologies need further demonstration to prove their capability to meet all the latest emissions criteria.

Environmental emissions from the most 'advanced' hot fuel gas cleaning arrangements already offer equal desulphurisation performance to commercially proven cold fuel gas cleaning, but other emissions are less good. This is especially the case for NOx emissions.

The ABGC with hot fuel-gas cleaning offers significantly lower environmental impact than pulverised fuel-fired power stations fitted with flue gas desulphurisation. However, the circulating fluidised bed combustion (CFBC) component of the ABGC cycle emits a flue gas which prevents the overall ABGC achieving the best emission performance attainable by hot fuel-gas cleaning. Further work is underway to reduce the CFBC emissions.

Table 31 summarises the relative cycle efficiencies of air-blown and oxygen-blown gasification, and draws comparison between the environmental capabilities. The ABGC is given a +0.5 percentage point advantage for 'simple' rather than 'advanced' hot fuel-gas cleaning but, thereafter, no distinction is made between the ABGC and other 'generalised' air-blown gasifiers. In reality, the ABGC would be expected to show an advantage over other air-blown designs because of the greater steam cycle flexibility afforded by its CFBC, but it is difficult to extract such subtlety from the data. As noted above, the IEACR study shows the ABGC to have between 1.4 and 2.6 percentage point advantage over the similar KRW fluid-bed gasifier which has no CFBC. However, the Krupp-Koppers PRENFLO study (when reduced by 2.5 percentage points, see above) and the CRIEPI study both show similar cycle efficiencies to the ABGC, even when air blowing single-stage and two-stage entrained flow gasifiers.

Capital and Operating Costs

Table 32 compares costing data for various GCC technologies and also shows calculated costs for electricity. Costing data is more sparse than cycle efficiency data and authors admit to high percentage uncertainties. The Indian coal study is not generally applicable, being influenced by the high ash coal and also Indian labour rates etc. There is a clear trend in favour of air-blown gasification.

The footnotes to Table 32 contain most recent projections of future costs for plants such as are being installed at Buggenum[85], US$1,700 per kW_e (1995 dollars) for a $500MW_e$ plant, and Puertollano[46], US$1,870 per kW_e (1991 dollars) for a $300MW_e$ (net electrical output) plant. These figures should be a reasonable guide to entrained flow, oxygen-blown, IGCC gasifiers, since they represent predictions based upon similar plant already (or nearly) constructed. IGCC plants are expected to show a 2-3% cost advantage compared to GCC (Table 28).

The 1990 SCS-KRW study is arguably the best indication that air-blown gasification costs may be less than oxygen-blown. This study considered one technology (the KRW fluidised bed gasifier) in considerable detail, rather than considering a range of technologies more superficially. Although the predicted costs, US$1,042 per kW_e for the air-blown KRW and US$1,218 per kW_e for the oxygen-blown KRW, seem surprisingly low, this is explained by the study being based upon an existing electric utility site already planned for extension (Plant Wansley site of Georgia Power Company). Thus, capital costs are reduced through sharing existing facilities. The reported air-blown cost is 86% that of oxygen-blown. This is somewhat unfair, however, since the study unintentionally penalises the oxygen-blown gasifier by omitting a carbon burn-up cell, thus reducing its cycle efficiency. If the oxygen-blown cycle is compensated by raising it 2 percentage points, the additional net power output reduces the plant specific cost to US$1,160 per kW_e. The air-blown specific cost is then 90% of the oxygen-blown specific cost.

Load Following

The ability of GCC and IGCC plants to follow changes in electrical load still requires development. No plant has yet achieved all the load following requirements previously written within specifications from the CEGB for pf fired boilers. In this respect, therefore, neither air-blown nor oxygen-blown gasification has yet demonstrated an advantage.

Future Developments

Future technological developments which will provide an efficiency boost for both the gas turbine and steam turbine parts of the overall combined cycle are already being reported. The ABGC, by having relatively benign, oxidising, 900°C CFBC flue gases, should be able to accommodate increases in steam cycle efficiency as well as gas turbine cycle efficiency, whereas other gasifier designs, both air-blown and oxygen-blown, may be limited to gas turbine cycle improvements. The IEACR study anticipates that the ABGC could use advanced supercritical steam conditions and suggests a commensurate +1.2 percentage point advantage (over and above the figures quoted in Table 31).

Fuel Flexibility

Fuel flexibility seems reasonable for all GCC technologies. Dry coal feed, entrained flow, gasifiers may be able to receive coals with a wide range of characteristics, since pre-drying and fine milling should bring them into a common form. Slurry fed, entrained flow, gasifiers might have an advantage with physically difficult feedstocks, where coarse grinding and forming them into a slurry should ease fuel feeding. This advantage is, however, paid for by a noticeable cycle efficiency penalty. Non-slagging gasifiers, such as air-blown fluidised beds, are expected to show a cycle efficiency benefit when the feedstock is a high ash coal, since no heat is lost melting the coal ash into slag. It is also easier to recover heat from the granular bed solids than is the case when molten ash is quench cooled.

India and China require major increases in power generating capacities, and the high cycle efficiency of an ABGC equipped with hot fuel-gas cleaning, ie +3.1 percentage points when compared to an oxygen-blown gasifier with cold fuel-gas cleaning and a non-integrated ASU is an attractive proposition. High-ash coal, as is typically available in these countries, will further increase the cycle efficiency advantage of the ABGC over competing technologies.

CONCLUSIONS

- Judicious selection of plant configuration can introduce a bias in favour of air- or oxygen-blown gasification. To avoid this, public domain studies have been reviewed rather than a new study being commissioned.

- Cold gas efficiency and fuel-gas calorific value are irrelevant when comparing gasification technologies for power generation application.

- When GCC studies are normalised and compared on the common basis of having a dry coal feed and advanced hot gas cleanup, air-blown GCC has an inherent 1.8 percentage point cycle efficiency advantage over oxygen-blown GCC. If an oxygen-blown gasifier has an integrated ASU, forming an IGCC, the cycle efficiency advantage to air-blown gasification is 0.7 percentage points.

- An air-blown GCC with advanced hot gas cleaning has an efficiency advantage of 2.6 percentage points over a conventional oxygen-blown GCC equipped with cold gas cleaning. Advanced hot gas cleaning incurs an efficiency penalty of 0.5 percentage points over simple in-bed limestone feeding for SO_2 emission reduction followed by downstream particulate capture. Thus, the simplest ABGC arrangement has a 3.1 percentage point cycle efficiency advantage over an oxygen blown GCC with cold gas cleaning.

- Capital costs for air-blown GCC are about 90% of non-integrated oxygen-blown GCC. Integrating the ASU confers a 2-3% capital cost reduction over a non-integrated arrangement.

- The capital cost of ASUs for GCC application has halved over the past 15 years, with an accompanying 10% reduction in their operating costs. ASUs inflate oxygen-blown GCC plant costs by 7 to 15%, with integrated ASUs being at the lower end of this range.

- The cost of electricity from air-blown GCC is some 20% cheaper than from oxygen-blown GCC. The cost varies depending on location, fuel etc.

- High ash content coals decrease the cycle efficiency of both air-blown and oxygen-blown GCCs. However the impact on low temperature fluidised bed systems is less pronounced. Slurry fed GCC systems are penalised more by higher ash contents.

- The ABGC is likely to benefit from advances in both steam and gas turbine technology. Other GCCs will need to rely on advances in gas turbine technology.

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ABBREVIATIONS

ABGC	Air-blown gasification cycle
AFT	Adiabatic flame temperature
ASU	Air separation unit
bar, bar_a, bar_g	Commonly applied unit of pressure = 100kN m^{-2}. When given a subscript, eg bar_a, bar_g, these denote that the original referenced text stated whether the pressure was absolute or gauge. In many instances references do not give a subscript, hence the use of bar in this report. Conventionally, bar is taken to mean bar_a.
BGL	British Gas\Lurgi
BHEL	Bharat Heavy Electricals Ltd (India)
CCPGG	Clean Coal Power Generation Group
CEGB	Central Electricity Generating Board (in the UK, now no longer in existence)
CFBC	Circulating fluidised bed combustion
CHAT	Cascaded humidified advanced turbine (cycle)
CRIEPI	Central Research Institute of Electric Power Industry (Japan)
CSIR	Council of Scientific and Industrial Research (India)
CTDD	Coal Technology Development Division (British Coal Corporation)
DTI	UK's Department of Trade and Industry
DCF	Discounted cash flow
EP-ASU	Elevated-pressure ASU
EPRI	Electric Power Research Institute, Palo Alto, California, USA
FGD	Flue gas desulphurisation
GCC	Gasification combined cycle
GE	General Electric Company of the USA

HAT	Humid air turbine (cycle)
HHV	Higher (gross) heating value
HRSG	Heat recovery steam generator
HTW	High Temperature Winkler
ICAD	Intercooled aero-derivative (gas turbine)
IEACR	IEA Coal Research Ltd
IGCC	Integrated gasification combined cycle
IGCHAT	Integrated gasification cascaded humidified advanced turbine (cycle)
kW_e	Kilowatt = Watt x 10^3. In all instances in this report, the use of the abbreviation kW implies an electrical output and the kW has a subscript in recognition of this, ie kW_e.
kWh	A kilowatt for an hour. In this report all use of kWh refers to the passage of electrical power, rated in kW_e, for one hour.
PRENFLO	Krupp-Koppers' Pressurised entrained-flow gasification process
KRW	Kellogg-Rust-Westinghouse
LASH	Sulphided limestone and coal ash solids from a fluidised bed gasifier
LHV	Lower (net) heating value, being the fuel calorific value which is expressed on the basis that indigenous water attached to a fuel (as with wetted coal) and water formed from hydrogen during combustion, both exist as vapour after combustion and remain as vapour, ie their latent heat of vaporisation is not recovered. This report tries to standardise on the use of LHV, but some referenced texts give only the HHV (see above) and do not always give sufficient information to enable a conversion to LHV.
LP-ASU	Low pressure ASU
m^3, Nm^3, Sm^3	Cubic metre. In the context of air, fuel and flue gas volumes, the prefixes N and S denote Normal and Standard m^3 respectively, and are repeated in this report exactly as in the original referenced texts. In the metric system, Nm^3 has been displaced by Sm^3, meaning a gas volume corrected to 0°C. In the USA (and sometimes incorrectly in the UK) Nm^3 is still used to

	(generally) denote 0°C. Where m^3 (alone) was stated in the original texts and refers to flue gas emission values, this report substitutes Sm^3, in order to imply the anticipated 0°C condition.
METC	Morgantown Energy Technology Centre (US DOE)
MHI	Mitsubishi Heavy Industries Ltd (Japan)
MITI	Japanese Government - Ministry of International Trade and Industry
MW_e, MW_{th}	Megawatt = Watt x 10^6. The subscripts denote whether the value is an electrical output (MW_e) or a thermal heat flow (MW_{th}).
NEDO	Japanese Government - New Energy and Industrial Technology Development Organisation
O&M	Operating and maintenance
pf	Pulverised fuel
PFBC	Pressurised fluidised bed combustion
SCS	Southern Company Services Inc (USA)
SGP	Shell gasification process
STIG	Steam injected gas turbine
TAG™	EPRI (see above) Technical Assessment Guide
TVA	Tennessee Valley Authority, USA
US$	United States of America, Dollars Currency
USc	United States of America, Cents Currency
USA	United States of America
US AID	United States of America Agency of International Development

Table 1 Comparison of Cold Gas Efficiencies (% lhv)

	Oxygen Blown	**Air Blown**	
Oxidant temperature	200°**C**	300 °C	500 °C
Dry coal feed entrained, single-stage (cooled)	81		
Dry coal feed entrained, single-stage	84	69	73
Dry coal feed entrained, two-stage	88	80	83
Wet coal feed entrained, single-stage	75	55	60
Wet coal feed entrained, two-stage	81	69	**73**
Dry coal feed, ABGC, no char credit		**59**	
Dry coal feed, ABGC, char subtracted from coal		78	

Notes:

The dry feed, single-stage entrained, cooled vessel, oxygen blown case is similar to Shell and Prenflo.

The wet feed, single-stage entrained, oxygen blown case is similar to Texaco.

The wet feed, two-stage entrained, oxygen blown case is similar to Destec.

The dry feed, two-stage entrained, air blown case is similar to Mitsubishi, except it is uncooled.

Table 2 Comparative GCC Efficiencies as Estimated By IEA Coal Research in Report IEACR/55

Case No. (IEACR/55 Numbering)		12	13	14	15	16	17	18	19	20	21	22	23	24	25
Blowing Medium		Air			Air	Oxygen	Oxygen	Oxygen							
Gasifier Type		Fluid-Bed with 20% Carbon to Separate CFBC Char Combustor			Fluid-Bed with 5% Carbon to Burn-up Cell		Moving Bed (Fixed Bed)	Single Stage, Entrained Flow.							
Gas Turbine Class		III	III	III	II	II	II	II	II	II	I	II	III	III	III
Design Variation		Cold Gas Cleaning	Hot Gas Cleaning	Hot Gas Cleaning + Supercritical Steam Cycle	Both Include Hot Gas Cleaning + Sulphur Polishing By Zinc Ferrite		Cold gas Cleaning	Dry Coal Feed, Cold Gas Cleaning	Dry Coal Feed, Hot Gas Cleaning	Wet Coal Feed, Hot Gas Cleaning	Wet Coal Feed, Cold Gas Cleaning	As Case 21 But Class II GT	As Case 22 But Class III GT	As Case 23 But Supercritical Steam Cycle	As Case 23 But Integrated ASU
Coal Input	t/h	163.8	132.4	132.4	83.3	78.8	79.8	88.9	84.7	92.6	54.9	97.3	136.3	136.3	124.7
	MW_{th} (hhv)	1133.6	916.5	916.5	576.5	545.3	552.3	615.1	586.2	641.0	380.1	672.8	943.1	943.1	862.8
Gas Turbine Power	MW_e	228.3	219.7	219.7	146.7	152.2	152.0	154.0	153.0	165.6	87.4	159.2	230.2	230.2	204.7
Steam Turbine Power	MW_e	296.5	219.0	232.5	111.7	97.0	86.3	133.1	120.0	123.8	72.4	143.4	200.6	212.5	186.8
Gross Power Output	MW_e	524.8	438.7	452.2	258.4	249.2	238.3	287.1	273.0	289.4	159.8	302.6	430.8	442.7	391.5
Auxiliary Power	MW_e	29.4	22.2	25.7	14.9	21.9	25.1	35.7	33.5	38.3	23.4	41.7	58.6	62.0	42.0
Net Power Output	MW,	495.4	416.6	426.5	243.5	227.3	213.2	251.4	239.5	251.1	136.4	260.9	372.2	380.7	349.5
Cycle Efficiency	%hhv	43.7	45.4	46.5	42.2	41.7	38.6	40.9	40.9	39.2	35.9	38.8	39.5	40.4	40.5
	%lhv	45.8	47.6	48.8	44.3	43.7	40.5	42.8	42.8	41.1	37.6	40.7	41.4	42.3	42.5
CO, emitted	gC/kWh	208	201	196	204	206	229.0	220	216	228	250	232	227	222	222

Table 3 Cost Comparison of GCC Systems by IEA Coal Research

Case No. (IEACR/55 Numbering)	12	13	15	16	17	18	19	20	21	22	23	24
Blowing Medium	Air		Air	Oxygen	Oxygen	Oxygen						
Gasifier Type	Fluid Bed with 20% Carbon to Separate CFBC Char Combustor		Fluid-Bed with 5% Carbon to Burn-up Cell		Moving Bed (Fixed Bed)	Single-Stage, Entrained Flow.						
Gas Turbine Class	III	III	II	II	II	II	II	II	I	II	**III**	**III**
Design Variation	Cold Gas Cleaning	Hot Gas Cleaning	Both Include Hot Gas Cleaning + Sulphur Polishing By Zinc Ferrite		Cold gas Cleaning	Dry Coal Feed, Cold Gas Cleaning	Dry Coal Feed, Hot Gas Cleaning	Wet Coal Feed, Hot Gas Cleaning	Wet Coal Feed, Cold Gas Cleaning	As Case 21 But Class II GT	As Case 22 But Class III GT	As Case 23 But Supercritical Steam Cycle
Net Power Output MW_e	495.4	416.6	243.5	227.3	213.2	251.4	239.5	251.1	136.4	260.9	372.2	380.7
Total capital Cost US$ Million	603.2	524.3	349.1	351.5	404.6	491.1	471.4	478.5	311.4	499.1	649.5	700.2
Specific Cost US$/$kW_e$	1218	1259	1434	1546	1933	1954	1968	1906	2283	1913	1745	1839

Table 4 IEA Coal Research Cost Comparison of GCC Systems, Normalised to a 300 MW_e Plant Net Electrical Output

Case No. (IEACR/55 Numbering)		12	13	15	16	17	18	19	20	21	22	23	24
Net Power Output	MW,	300											
Specific Cost	US$/$kW_e$	1402	1380	1353	1430	1748	1859	1848	1812	Omitted	1840	1854	1966
Cycle Efficiency	(hhv) %	43.7	45.4	42.2	41.7	38.2	40.9	40.9	39.2	Case 22 Substitutes	38.8	39.5	**40.4**
(As in Table 2)	(lhv) %	45.8	47.6	44.3	43.7	40.1	42.8	42.8	41.1		**40.7**	**41.4**	**42.3**

Table 5 IEA Coal Research Comparison of Electricity Costs, USc/kWh

IEA Case No.	Blowing Medium	Gasifier Type	Electricity Costs USc/kWh
13	Air with Steam	Fluid-Bed (as ABGC), Dry Feed, Hot Gas Cleanup	4.21
15		Fluid-Bed (as KRW), Dry Feed, Hot Gas Cleanup	4.75
17	Oxygen with Steam	Moving Bed, Dry Feed, Cold Gas Cleanup	6.00
18		Entrained Flow, Dry Feed, Cold Gas Cleanup	5.94
22		Entrained Flow, Wet Feed, Cold Gas Cleanup	5.95

Table 6 Comparative Cycle Efficiencies (%lhv) From CRIEPI/MHI Study (1982-85)

Gasifier Arrangement	Fuel Gas Cleaning Method	Air or Oxygen Blown	Net Cycle Efficiency %lhv	Change in Cycle Efficiency Due to Cold Fuel Gas Cleaning Method
Two-Stage Entrained Flow, Dry Coal Feed.	Hot (dry)	Air	45.4	-1.9
	Cold (wet)	Air	43.3	
Two-Stage Entrained Flow, Slurry Coal Feed.	Hot (dry)	Air	43.4	-2.6
	Cold (wet)	Air	42.4	
Single-Stage Entrained Flow, Dry Coal Feed.	Hot (dry)	Oxygen	43.6	-0.9
	Cold (wet)	Oxygen	42.0	
Single-Stage Entrained Flow, Slurry Coal Feed.	Hot (dry)	Oxygen	41.0	-0.9
	Cold (wet)	Oxygen	41.1	

Table 7 KRW Fluidised Bed Gasifiers in Air Blown and Oxygen Blown Modes

Variable		Air Blown	Oxygen Blown
Gasifier Operating Temperature	°C	1038	1010
Coal Flow Rate	t/h	143.3	145.3
Air to Gas Turbine	t/h	2711.5	2735.5
Gasifier Air Requirement	t/h	487.1	not applicable
Gasifier Oxygen Requirement	t/h	not applicable	88.7
Gasification Pressure	bar_a	27.6	32.1
Gas Turbine Firing Temperature	°C	1224	1224
Gas Temperature at Turbine Exit	°C	592	601
Main Steam Pressure	bar_a	100	93.1
Main Steam Temperature	°C	538	521
Reheat Steam Pressure	bar_a	24.1	24.1
Condenser Pressure	bar_a	0.12	0.12
Steam Turbine Gross Power	MW_e	182.2	159.4
Gas Turbine Gross Power	MW_e	311.6	298.8
Total Gross Power	MW_e	493.8	458.2
Auxiliary Power Consumption	MW_e	35.4	44.7
Net Power Output	MW_e	458.4	413.5
Cycle Efficiency	(hhv) %	39.7	37.7
	(lhv) %	41.6	39.3

Table 8 Cost Comparison for KRW Air and Oxygen Blown Gasifiers

	Air Blown US$ Million (mid-1990)	Oxygen Blown US$ Million (mid-1990)
Coal Reception, Handling and Preparation	15.90	16.88
Limestone Reception and Handling	9.07	not applicable
Boost Air Compressor	8.82	not applicable
Air Separation Unit (ASU)	not applicable	55.36
Gasification	55.16	86.47
Recycle Gas Compression	13.94	not stated
Gas Conditioning	35.21	not applicable
Zinc Ferrite Sulphur Polishing	20.17	not applicable
Sulphided Lime Sulphator	13.09	not applicable
Acid Gas Removal (Selexol^{R})	not applicable	10.65
Sulphur Recovery (Claus)	not applicable	5.88
Tail gas Treatment (SCOT)	not applicable	5.30
Sour water Stripping	not applicable	3.82
Waste-water Treatment	not applicable	4.44
Gas Turbine System	70.78	147.97
Steam Turbine System	46.27	
Heat recovery Steam Generator (HRSG)	24.93	
Ash and Fines Handling/Disposal	5.79	3.19
General Facilities and Fees	58.95	64.79
Contingencies	99.4	99.0
Total Capital Cost US$ Million (mid-1990)	477.5	503.8
Net Power Output MW_e	458.4	413.5
Specific Capital Cost US$/$kW_e$	1042	1218

Table 9 Comparison of Electricity Production Costs from KRW Study

(10 year levelised, 65% capacity factor, mid-1990 US Dollars)

	Air Blown	Oxygen Blown
Fuel Costs USc/kWh	1.32	1.39
Capital Charges USc/kWh	2.67	3.08
Fixed O&M Costs USc/kWh	0.67	0.77
Variable O&M Costs USc/kWh	0.24	0.1
Total USc/kWh	4.91	5.34

Table 10 Comparison of Various Gasifiers Using Indian High Ash Coal

	Air Blown		Oxygen Blown	
	KRW: Fluidised Bed Dry Coal Feed Hot Gas Cleaning	CSIR Design: Moving Bed Dry Graded Coal Cold Gas Cleaning	Shell: Entrained Flow Dry Coal Feed Cold Gas Cleaning	Texaco: Entrained Flow Slurry Coal Feed, Cold Gas Cleaning
Coal Feed t/d	8,890	10,768	9.964	12,576
Thermal Input MWth (hhv)	1,435	1,738	1,608	2,030
MWth (lhv)	1,364	1,652	1,529	1,930
Limestone t/d	280	0	618	0
Raw Water m^3/h	1,019	1,698	1,324	2,224
Steam Turbine Power MW_e	280.2	271.8	308.4	258.0
Gas Turbine Power MWe	350.4	368.3	392.2	396.5
Gross Power MW_e	630.6	640.1	700.6	654.5
Auxiliary Power MW_e	53.4	67.6	136.2	158.3
Net Power MW_e	577.2	572.5	564.4	496.2
Net Thermal Efficiency (hhv) %	40.2	32.9	35.1	24.4
(lhv) %	42.3	34.6	36.9	25.7
Sulphur By-product t/d	0	32	29	37

Table 11 Comparison (Indian Basis) of Capital and Electricity Generation Costs

High Ash Indian Coal Study

Gasifier Type	Total capital US$ Million	Specific Cost US$/$kW_e$	Electricity Cost (6,000 h/y equiv. 68.5% load factor) USc/kWh
KRW Fluidised Bed, Air Blown	859	1,488	5.6
CSIR Moving Bed, Air Blown	899	1,535	6.2
Shell Entrained Flow, Oxygen Blown	1,209	2,143	7.9
Texaco Entrained Flow, Oxygen Blown	1,211	2,442	9.3

Table 12 Comparison by IEA Coal Research of IGCC Efficiencies with Hot and Cold Fuel Gas Cleaning

Gasifier Configuration	Case No. (IEACR/55 Numbering)	Fuel Gas Cleaning Method	Overall Cycle Efficiency %lhv	Change in Cycle Efficiency Due to Cold Fuel Gas Cleaning Method
Air Blown, Fluidised Bed with CFBC Char Combustor (ABGC)	13	Hot	47.6	-1.8
	12	Cold	45.8	
Oxygen Blown, Entrained Flow with Dry Coal Feed	19	Hot	42.8	0
	18	Cold	42.8	
Oxygen Blown, Entrained Flow with Wet Coal Feed	20	Hot	41.1	-0.4
	22	Cold	40.7	

Table 13 Comparison by The New University of Ulster of Hot and Cold Fuel Gas Cleaning

	Air Blown, Fluidised Bed, Dry Coal Feed, ABGC Gasifier.				Oxygen Blown, Entrained Flow, Dry Coal Feed, Shell Gasifier.		Oxygen Blown, Entrained Flow, Slurry Coal Feed, Texaco Gasifier.	
	Hot Gas Particulate Filtration Only		Advanced Hot Gas	Cold Gas Compared to Advanced Hot Gas	Advanced Hot Gas	Cold Gas Compared to Advanced Hot Gas	Advanced Hot Gas	Cold Gas Compared to Advanced Hot Gas
	At 650°C	At 250°C						
Cycle Efficiency on Australian (Belambi) Coal %lhv	46.6%	-1.0	46.0%	-1.8	46.2%	-0.6	43.0%	-1.8
Cycle Efficiency on UK (Point of Ayr) Coal %lhv	47.3%	-0.9	46.7%	-1.8	46.3%	-1.6	42.0%	-1.6
Range for ten coals studied %lhv	45.4% to 49.9%	-0.7 to -1.4	not reported	not reported	45.7% to 47.0%	0 to -1.6	41.6% to 44.4%	-1.6 to -2.8

Table 14 Comparison of ASUs from BOC Cryoplants and Air Products (Dec 1987 US$)

	BOC Cryoplants	**Air Products**
Capital Cost US$ X 10^6	27.5 (13,855 US$/USt per day)	44.00 (22,165 US$/USt per day)
Power Requirement MW_e	28.80 (348 kWh/USt)	25.69 (311 kWh/USt)
Cooling Water l/m	37,120	25,015
Annual O&M Costs US$ X 10^6		
Maintenance costs	0.55	0.88
Administrative and support	0.07	0.11
Electricity @ 5.5USc/kWh	12.49	11.14
Cooling water @ 0.lUS$/1,000 l	1.85	1.25
TOTAL	14.96	13.37
Financial Factors		
Constant Dollar discount rate	0.10	0.10
O&M cost levelising factor	1.20	1.20
Lcveliscd Cost US$ x 10^6		
Capital contribution	2.75	4.40
O&M contribution	17.95	16.04
TOTAL	20.70 (10,430 US$/USt per day)	20.44 (10,300 US$/USt per day)

Table 15 Types of ASU Integration Intended for Demonstration

Plant Location	Gasifier Type	How Air is Supplied to the ASU	Whether Nitrogen is Supplied to the Gas Turbine
Buggenum, The Netherlands	Shell entrained flow	From the gas turbine	Yes
Puertollano, Central Spain	Krupp Koppers entrained flow	From the gas turbine	**Yes**
Wabash River West Terre Haute, Indiana, USA	Destec, two-stage, entrained flow	From a standalone compressor	No
Tampa Electric Polk County, Florida, USA	Texaco, entrained flow	From a standalone compressor	Yes

Table 16 TAG[IM] 1993 Comparison Data for Different Degrees of ASU Integration

	Non-Integration	Partial Integration	Full Integration
ASU cost US$/kW$_e$	146	148	**134**
Total Plant Cost US$/kW$_e$	1686	1467	1439
Total Capital Requirement US$/kW$_e$	1951	1695	1663
Cycle Efficiency	39.26	39.08	40.33

Table 17 Performance (32°C ambient) of Integrated and Non-Integrated K-K PRENFLO Schemes

	Non-Integrated ASU	Integrated ASU	Data Comparison
Coal Feedrate (dry) metric t/h	117.0	110.9	
Gas Turbine Output MW$_e$	317.8	262.2	The difference in performance reflects the different gas mass flows passing through the compressor turbine (326.9 vs 362.6 kg/s) and power turbine (382.8 vs 376.3 kg/s)
Steam Turbine Output MW$_e$	188.2	213.0	Although the net plant steam generation is similar in both cases, the non-integrated scheme consumes more sensible heat in the saturator to moisture the fuel gas. Thus, the integrated case produces 13% more power.
Gross Plant Output MW$_e$	506.0	475.2	
Auxiliary Power MW$_e$	66.1	45.4	The ASU of the integrated case consumes 43 % less auxiliary power than the non-integrated requires with its stand-alone air compressor. Overall, the integrated case has 30% less parasitic power.
Net Plant Output MW$_e$	439.9	429.8	
Net Plant Efficiency %(HHV)	41.6	42.8	A significant benefit of the integrated ASU scheme is a 2-3 % improvement in net plant heat rate, primarily a result of the lower auxiliary power consumption.
%(LHV)	43.1	44.4	
Plant Costs mid-1990 US$	654.3 million	622.3 million	The integrated ASU case (on a **US$/kW$_e$** basis) shows an advantage of about 3 % compared to the non-integrated case.
Specific Costs US$/kW$_e$	1,487	1,448	
Total capital required (excl. AFUDC) mid-1990 US$	684.8 million	651.3 million	
Specific Costs US$/kW$_e$	1,557	1.515	
30 year Levelised Generating Costs			The cost of electricity from the integrated ASU scheme is about 5 % lower than from the non-integrated scheme.
65% capacity factorUSc/kWh	5.9	5.6	
75% capacity factor USc/kWh	5.3	5.1	
85% capacity factor USc/kWh	4.9	4.7	

Table 18 Performance (15°C ambient) of Integrated and Non-Integrated Destec Schemes

		Non-Integrated **ASU**	**Integrated ASU**
Coal Feedrate (dry)	t/h	192	191
Gas Turbine Output	MW_e	384	384
Steam Turbine Output	MW_e	220.4	247.6
Gross Plant Output	MW_e	604.4	631.6
Auxiliary Power	MW_e	73.1	90.1
Net Plant Output	MW_e	531.3	541.6
Net Plant Efficiency	%(hhv)	39.1	40.2
	%(lhv)	40.8	41.9
Water Consumption	l/kWh	1.6	1.37
Total Plant Costs	Jan-1990 US$	665 million	679 million
Cost of the ASU (% of total plant cost)	Jan-1990 US$	90 (13.5%)	93 (13.7%)
Specific Costs	US$/$kW_e$	1,251	1,253
Levelised Generating Costs Based upon coal at 1.5US$/GJ (hhv)		4.76	4.71

Table 19 Composition of Dry Fuel Gas at the Outlet of Oxygen Blown Gasifiers

Component		Texaco Gasifier	Shell/ K-K PRENFLO	BGL
CO	% by volume	53.3	63.9	57.7
H_2	% by volume	33.9	30.1	26.1
CH_4	% by volume	0.2	0.03	6.3
C_xH_y	% by volume	-	-	3.1
CO_2	% by volume	10.3	1.3	2.8
N_2 / Ar	% by volume	2.0	4.35	3.6
H_2S	% by volume	0.3	0.3	0.3
COS	% by volume	0.01	0.02	0.02
Adiabatic Flame Temperature	°C	2,260	2,348	2,239
Fuel Gas Net Calorific Value (lhv)	MJ/Sm^3	10.5	11.4	19.0

Table 20 Cleaned Fuel Gas from Oxygen Blown Gasifiers, Diluted for NO_x Suppression

Component		Texaco Gasifier	Shell/ K-K PRENFLO	BGL
CO	% by volume	19	27	25
H_2	% by volume	12	11	11.5
CH_4	% by volume	-	-	2.8
C_xH_y	% by volume	-	-	1.4
CO_2	% by volume	7	1	1.2
N_2 / Ar	% by volume	57	56	53
H_2O	% by volume	5	5	5.1
Adiabatic Flame Temperature	°C	1,568	1,747	1.869
Fuel Gas Net Calorific Value (lhv)	MJ/Sm^3	3.7	4.5	7.4

Table 21 Latest Heavy Duty Gas Turbines Firing Natural Gas

Manufacturer	Simple Cycle Capacity MW_e	Simple Cycle Turbine Efficiency %	Combined Cycle Efficiency %	Firing Temperature °C	Compressor Pressure Ratio
General Electric 9F	226	35.7	55	1240	15:1
Siemens V94.3	211	36.1	54	1245	15.6:1
ABB GT26	240	37.8	58	1235	30:1
General Electric 9G Also being developed as the 9H using closed circuit steam cooling (see text Section 9.1)	282	39.5	58 (9H = 60%)	1430	23:1

Table 22 Danish Development of Steam Turbine Power Generation, 1995 - 2015

Steam Parameters	Year of Commissioning	Station Net Efficiency	Turbine Heat Rate Efficiency
290 bar, 580/580/580°C	1998	~ 47 %	~ 53.5 %
300 bar, 600/600/600°C	2000	~ 48 %	~ 55 %
325 bar, 610/630/630°C	2005	~ 50 %	~ 57 %
375 bar, 700/700/700°C	2015	• 52 %	~ 59.5 %

Table 23 Comparison Emissions for Hot Gas and Cold Gas Cleaning

(References: The New University of Ulster and CTDD)

Pollutant		ABGC Hot Gas Cleaning. 90-95 % in-bed desulphurisation via limestone feeding		Shell Cold Gas Cleaning (M-Sulphinol)
		Hot Gas Particulate Removal Only (In-Bed Desulphurisation)	Advanced Hot Gas Cleaning; Includes Part Removal of N_2 Compounds and H_2S Polishing	
CO	mg/Nm^3	8	not reported	13
HC1	mg/Nm^3	140	40	0
NO_x	mg/Nm^3	415	<230	21
SO_2	mg/Nm^3	145	<75	32
CO_2	g/kWh	710	<700	736

Table 24 Parameter Range of Coals Studied by the University of Ulster

		Minimum Value	Maximum Value
Proximate Analysis			
Water content	% as received	0.70	9.34
Ash content	% dry, basis	4.20	25.78
Volatile matter	% dry, basis	5.20	39.59
High (gross) Heating Value (hhv)	MJ/kg, dry, ash free, basis	33.0	36.4
Low (net) Heating Value (Ihv)	MJ/kg, dry, ash free, basis	32.3	35.1
Ultimate Analysis			
Carbon	% dry, ash free	81.06	93.22
Hydrogen	% dry, ash free	3.10	5.70
Oxygen	% dry, ash free	1.81	10.46
Nitrogen	% dry, ash free	0.88	2.1
Sulphur	% dry, ash free	0.75	3.07
Chlorine	% dry, ash free	0	0.98

Table 25 Comparison of Cycle Efficiencies

Air-Blown versus Oxygen-Blown Gasification

Air Blown										Oxygen **Blown**								Comments
Fluidised Bed				Mov'g Bed	Entrained Flow					Fluidised Bed			Mov'g Bed	Entrained Flow				The maximum net % cycle efficiency value is shown for each study, followed by the percentage point changes on that value which apply to the alternative technologies.
ABGC		KRW	HTW	CSIR	Two-stage				1-stage	KRW		HTW	BG/L	Single-stage				
Dry Coal Feed							Slurry Coal Feed		Dry Coal Feed	Dry Coal Feed						Slurry Coal Feed		
Fuel Gas Cleaning Method																		
Hot	Cold	Hot	Cold	Cold	Hot	Cold	Hot	Cold	Cold	Hot	Cold	Cold	Cold	Hot	Cold	Hot	Cold	
46.9	-1.8	-2.6								-3.2			-6.4	-4.1	-4.1	-5.8	-6.2	IEA study
					45.4	-1.9	-1.8	-4.4						-2.1	-3.0	-3.4	-4.3	CRIEPI study
									-0.2						46.9			KK PRENFLO study
			45.5									-0.5						**HTW** KoBra study
		41.6									-2.3							SCS KRW study
		42.3		-7.7				________							-5.4		-16.6	Indian coal study
46.4	-1.8												-0.6	-0.1	-1.3	-3.9	-5.6	Univ. of Ulster

The table summarises results from each of the reported studies. The highest cycle efficiency reported within each study is stated, and percentage point variations are then shown for the cycle efficiency of the other competing concepts. Also note:

- Values are consistent within each study (across a row) but cannot necessarily be compared vertically (within a column).
- Values of cycle efficiency are all based upon fuel lhv.
- In the IEA study, the base value, ie, the cycle efficiency of the ABGC gasifier, has been reduced by 0.7 percentage points over the value shown in Table 2, to compensate for being originally based upon a Class III gas turbine, while all other values arc based upon a Class II.
- The KK PRENFLO study is biased against the air-blown concept by being based upon an inappropriate (for air) single-stage entrained flow gasifier, while the oxygen version is advantaged by having an integrated ASU, ie, an IGCC.
- The SCS study is biased against the oxygen-blown concept because the oxygen version does not have a carbon burn-up cell.
- The University of Ulster study uses a constant steam cycle efficiency for all concepts, which disadvantages the ABGC while flattering other concepts, particularly the British Gas/Lurgi. The base value of 46.4% is for the ABGC equipped with 'advanced' hot gas cleanup. All values are %lhv, converted from %hhv in the original reference.

Table 26 Comparison of Hot (Dry) and Cold (Wet) Fuel Gas Cleaning

Air-Blown versus Oxygen-Blown Gasification

Air Blown				Oxygen **Blown**		Comments
Fluidised Bed		Entrained Flow		Entrained Flow		See following Notes.
ABGC		Two-stage		Single-stage		
Dry Coal Feed			Slurry Coal Feed	Dry Coal Feed	Slurry Coal Feed	
From 'Simple' Hot to Cold Gas Cleanup	From 'Advanced' Hot to Cold Gas Cleanup	From 'Advanced' Hot to Cold Gas Cleanup	From 'Advanced' Hot to Cold Gas Cleanup	From 'Advanced' Hot to Cold Gas Cleanup	From 'Advanced' Hot to Cold Gas Cleanup	
	-1.8			0	-0.4	1EA study
		-1.9	-2.6	-0.9	-0.9	CRIEPI study
-2.4	-1.8			0 to -1.6 range over ten coals. Average -0.8	-1.6 to -2.8 range over ten coals. Average -2.2	University of Ulster
				-1.3 [6] to -2.3 [40]		Dutch studies **[6,40]**

Notes:

The table summarises results from each of the studies. Values are percentage point cycle efficiency changes attributable to moving from hot (dry) to cold (wet) fuel gas cleaning for each specified gasifier technology.

Table 27 Comparison of Dry Coal Feeding with Slurry Coal Feeding

Air-Blown versus Oxygen-Blown Gasification

<table>
<tr><th colspan="2">Air Blown</th><th colspan="2">Oxygen Blown</th><th>Comments</th></tr>
<tr><td colspan="2">Entrained Flow</td><td colspan="2">Entrained Flow</td><td rowspan="4">See following Notes.</td></tr>
<tr><td colspan="2">Two-stage</td><td colspan="2">Single-stage</td></tr>
<tr><td>Advanced Hot Gas Cleanup</td><td>Cold Gas Cleanup</td><td>Advanced Hot Gas Cleanup</td><td>Cold Gas Cleanup</td></tr>
<tr><td>Dry Coal Feed to Slurry Coal Feed</td><td>Dry Coal Feed to Slurry Coal Feed</td><td>Dry Coal Feed to Slurry Coal Feed</td><td>Dry Coal Feed to Slurry Coal Feed</td></tr>
<tr><td></td><td></td><td>-1.7</td><td>-2.1</td><td>IEA study</td></tr>
<tr><td>-1.8</td><td>-2.5</td><td>-1.3</td><td>-1.3</td><td>CRIEPI study</td></tr>
<tr><td></td><td></td><td>-1.8 to -4.8
range over ten coals.
Average -3.2</td><td>-2.7 to -6.0
range over ten coals.
Average -4.5</td><td>University of Ulster</td></tr>
</table>

Notes:

The table summarises results from each of the studies. Values are percentage point cycle efficiency changes attributable to moving from dry coal feeding to slurry coal feeding for each specified gasifier technology.

Table 28 Comparison of Cycle Efficiencies From Integrated and Non-Integrated ASUs in IGCC and GCC Systems

Integrated ASU			Non-Integrated ASU			Comments
Entrained Flow Gasifiers			Entrained Flow Gasifiers			The net % cycle efficiency value is shown for the Integrated ASU scheme, with the change in percentage points of efficiency for the corresponding Non-Integrated ASU scheme appearing in the same row. Below the Integrated ASU % cycle efficiency value is given the change in plant specific costs attributable to moving from an Integrated to a Non-Integrated scheme.
Single-stage		Two-stage	Single-stage		Two-stage	
Dry Coal Feed	Slurry Coal Feed	Slurry Coal Feed	Dry Coal Feed	Slurry Coal Feed	Slurry Coal Feed	
Cold Gas Cleanup	Cold Gas Cleanup	Cold Gas Cleanup	Cold Gas Cleanup	Cold Gas Cleanup	Cold Gas Cleanup	
	42.5% Cost change not reported			-1.1		IEA study
	40.33% Costs -14.65%			-1.07		TAG™ guide
	38.7% 39.1% Costs -2.8%			-0.64 [51] -1.07 [53]		Union Carbide / GE studies 1987 [51] 1991 [53]
44.4% Costs -2.7%			-1.3			KK PRENFLO study
		41.9% Costs; No change			-1.1	Fluor Daniel study (partial integration)

Table 29 Cycle Efficiency Percentage Point Variations Dependent Upon Technology Options

Air-Blown versus Oxygen-Blown Gasification

Technology Option Variable	**Technology Option Kept Constant**	**Air Blown**	**Oxygen Blown**
Changing from a Dry Coal Feed to a Slurry Coal Feed	Cold Fuel Gas Cleaning	**-2.6** percentage points	
	Hot Fuel Gas Cleaning	-2.0 percentage points	
Changing from Hot to Cold Fuel Gas Cleaning	Slurry Coal Feeding	-2.6 percentage points	-1.6 percentage points
	Dry Coal Feeding	-1.8 percentage points	-0.8 percentage points
Changing from Integrated to Non-Integrated ASU	Fuel Gas Cleaning Method Coal Feeding Method	Not applicable	-1.1 percentage points
Changing from 'Simple' to 'Advanced' Hot Fuel Gas Cleaning	Applies to **ABGC** Concept	-0.5 percentage points	Not applicable

Table 30 Comparison of GCC Technologies 'Corrected' (as necessary) to Dry Coal Feeding and Hot Gas Cleaning

Air-Blown versus Oxygen-Blown Gasification

Air Blown									Oxygen Blown							Comments
Fluidised Bed				Entrained Flow					Fluidised Bed			Entrained Flow				See following notes
ABGC		KRW	HTW	Two-stage				1-stage	KRW		HTW	Single-stage				
Dry Coal Feed						Slurry • Dry Coal Feed		Dry Coal Feed	Dry Coal Feed					Slurry • Dry Coal Feed		
Hot Gas Clean	• Hot Gas Clean	Hot Gas Clean	• Hot Gas Clean	Hot Gas Clean	• Hot Gas Clean	Hot Gas Clean	• Hot Gas Clean	• Hot Gas Clean	Hot Gas Clean	• Hot Gas Clean	• Hot Gas Clean	Hot Gas Clean	• Hot Gas Clean	Hot Gas Clean	• Hot Gas Clean	
46.9	46.9 (0)	44.3 (-2.6)							43.7 (-3.2)			42.8 (-4.1)	43.6 (-3.3)	43.1 (-3.8)	44.9 (-2.0)	IEA study
				45.4 (-0.8)	45.3 (-0.7)	45.6 (-0.6)	46.2					43.3 (-2.9)	43.2 (-3.0)	44.0 (-2.2)	45.3 (-0.9)	CRIEPI study
								48.5					46.6 (-1.9)			KK PRENFLO study
			47.3								45.8 (-1.5)					HTW KoBra study
		41.6								40.1 (-1.5)						SCS KRW study
		42.3											37.7 (-4.6)		29.9 (-12.4)	Indian coal study
46.4	46.4 (0)											46.3 (-0.1)	45.9 (-0.5)	44.5 (-1.9)	45.0 (-1.4)	Univ. of Ulster

All net % cycle efficiency values are shown, modified as necessary so that all technologies are based upon 'advanced' hot gas cleaning and a dry coal feed. The value in parenthesis is the reduction from the maximum value (in percentage cycle efficiency points). Also note:

- Values are consistent within each study (across a row) but cannot necessarily be compared vertically (within a column).
- The moving bed gasifier design results have been omitted as such a design is inappropriate for hot gas cleaning.
- The SCS study is biased against the oxygen-blown concept because the oxygen version does not have a carbon burn-up cell.
- The High Ash Indian coal study is biased against slagging gasifiers because of heat lost as molten slag. Also, the slurry fed (Texaco) design uses low efficiency fuel gas quench cooling.

Table 31 Typical Cycle Efficiency Advantages Deduced for Air Blown over Oxygen Blown Gasification

Comparison			Cycle Efficiency Comparison with Oxygen Blown Gasification	
Technologies Being Compared		Environmental Comparison	Non-Integrated ASU (GCC)	Integrated ASU (IGCC)
Air Blown	Oxygen Blown			
The ABGC with 'simple* in-bed desulphurisation and downstream hot fuel gas paniculate capture.	Entrained flow, slagging gasifier (as Shell) equipped with dry coal feeding and cold fuel gas cleaning	ABGC: Medium Oxygen Blown: Excellent	The ABGC is +3.1 percentage points higher	The ABGC is +2.0 percentage points higher
Generalised Air Blown gasification equipped with dry coal feeding and 'advanced' hot fuel gas cleaning	Entrained flow, slagging gasifier equipped with dry coal feeding and cold fuel gas cleaning	Air Blown (incl. ABGC): Good Oxygen Blown: Excellent	Air Blown is +2.6 percentage points higher	Air Blown is +1.5 percentage points higher
Generalised Air Blown gasification equipped with dry coal feeding and 'advanced' hot fuel gas cleaning	Entrained flow, slagging gasifier equipped with dry coal feeding and advanced hot fuel gas cleaning	Air Blown (incl. ABGC): Good Oxygen Blown: Good (slagged bed ash disposal easier than non-slagged air blown bed ash, especially LASH)	Air Blown is +1.8 percentage points higher	Air Blown is +0.7 percentage points higher
Generalised Air Blown gasification equipped with dry coal feeding and cold fuel gas cleaning	Entrained flow, slagging gasifier equipped with dry coal feeding and cold fuel gas cleaning	Air Blown: Near Excellent (bed ash disposal will be a greater problem than with slagged ash) ABGC: Good (suffers due to CFBC emissions and bed ash disposal) Oxygen Blown: Excellent (best of all, because of glassy bed ash slag)	Air Blown is +0.8 percentage points higher	Oxygen Blown is +0.3 percentage points higher

In the above table, generalised air blown gasification is placed on a par with the ABGC, whereas the ABGC is expected to have the advantage due to its CFBC giving greater steam cycle parameter flexibility and, hence, enhanced overall cycle efficiency. If this is the case, then generalised air blown gasification in the above table should suffer an equivalent percentage point reduction in its advantage over oxygen blown gasification.

Table 32 Comparison of Specific Plant Costs and Electricity Prices

Air-Blown versus Oxygen-Blown Gasification

<table>
<tr><td colspan="9">Air Blown</td><td colspan="9">Oxygen Blown</td><td>Comments</td></tr>
<tr><td colspan="4">Fluidised Bed</td><td>Moving Bed</td><td colspan="4">Entrained Flow</td><td colspan="3">Fluidised Bed</td><td>Moving Bed</td><td colspan="5">Entrained Flow</td><td rowspan="5">Values shown are respectively:

specific plant costs, US$/kW$_e$

cost of electricity, USc/kWh</td></tr>
<tr><td colspan="2">ABGC</td><td>KRW</td><td>HTW</td><td>CSIR</td><td>Two-stage</td><td colspan="3">Single-stage</td><td colspan="2">KRW</td><td>HTW</td><td>BG/L</td><td>Two-stage</td><td colspan="4">Single-stage</td></tr>
<tr><td colspan="7">Dry Coal Feed</td><td colspan="2">Slurry Coal Feed</td><td colspan="7">Dry Coal Feed</td><td colspan="2">Slurry Coal Feed</td></tr>
<tr><td colspan="18">Fuel Gas Cleaning Method</td></tr>
<tr><td>Hot</td><td>Cold</td><td>Hot</td><td>Cold</td><td>Cold</td><td>Hot</td><td>Cold</td><td>Hot</td><td>Cold</td><td>Hot</td><td>Cold</td><td>Cold</td><td>Cold</td><td>Hot</td><td>Hot</td><td>Cold</td><td>Hot</td><td>Cold</td></tr>
<tr><td>1380

4.21</td><td>1402</td><td>1353

4.75</td><td colspan="6"></td><td>1430</td><td></td><td></td><td>1748

6.00</td><td></td><td>1848</td><td>1859

5.94</td><td>1812</td><td>1840

5.95</td><td>IEA study
mid-1991 US$
All at 300 MW,</td></tr>
<tr><td colspan="2"></td><td>1042

4.91</td><td colspan="6"></td><td></td><td>1218

5.34</td><td colspan="7"></td><td>SCS KRW study
mid-1990 US$
Average 436 MW$_e$</td></tr>
<tr><td colspan="2"></td><td>1488

5.6</td><td></td><td>1535

6.2</td><td colspan="4"></td><td colspan="6"></td><td>2143

7.9</td><td></td><td>2442

9.3</td><td>Indian coal study
end-1989 US$
Nominal 600 MW$_e$</td></tr>
</table>

The table summarises specific cost information, in US$/kW$_e$ (net electrical output), and costs of electricity, in USc/kWh, as reported in the various studies. Also note:

- Values are consistent within each study (across a row) but cannot necessarily be compared vertically (within a column).
- The SCS study is biased against the oxygen-blown concept because the oxygen version does not have a carbon burn-up cell. If the cycle efficiency is compensated by an increase of 2 percentage points, the specific cost falls to 1,160 US$/kW$_e$, but this may then be a small underestimate, as it does not include the capital cost of the burn-up cell.

Information from elsewhere [34] reports an average specific cost for oxygen blown gasifier GCC installations (not fluidised bed) of 1,401 US$/kW$_e$ for 500 MW$_e$ plants, based upon end-1994 US$

Current (mid-1995) Dutch information, based upon experiences of installing the plant at Buggenum [85], is the anticipation of 1,700 US$/kW$_e$ for a future 600 MW$_e$ installation.

Current (mid-1995) information from Puertollano [46] is for the existing 300 MW$_e$ demonstration plant to have cost 1,858 US$/kW$_e$ (gross electrical output, equivalent to 2,075 US$/kW$_e$ net electrical output), but a future coastal plant is estimated likely to cost 1,675 US$/kW$_e$ (gross electrical output, equivalent to 1,870 US$/kW$_e$ net electrical output).

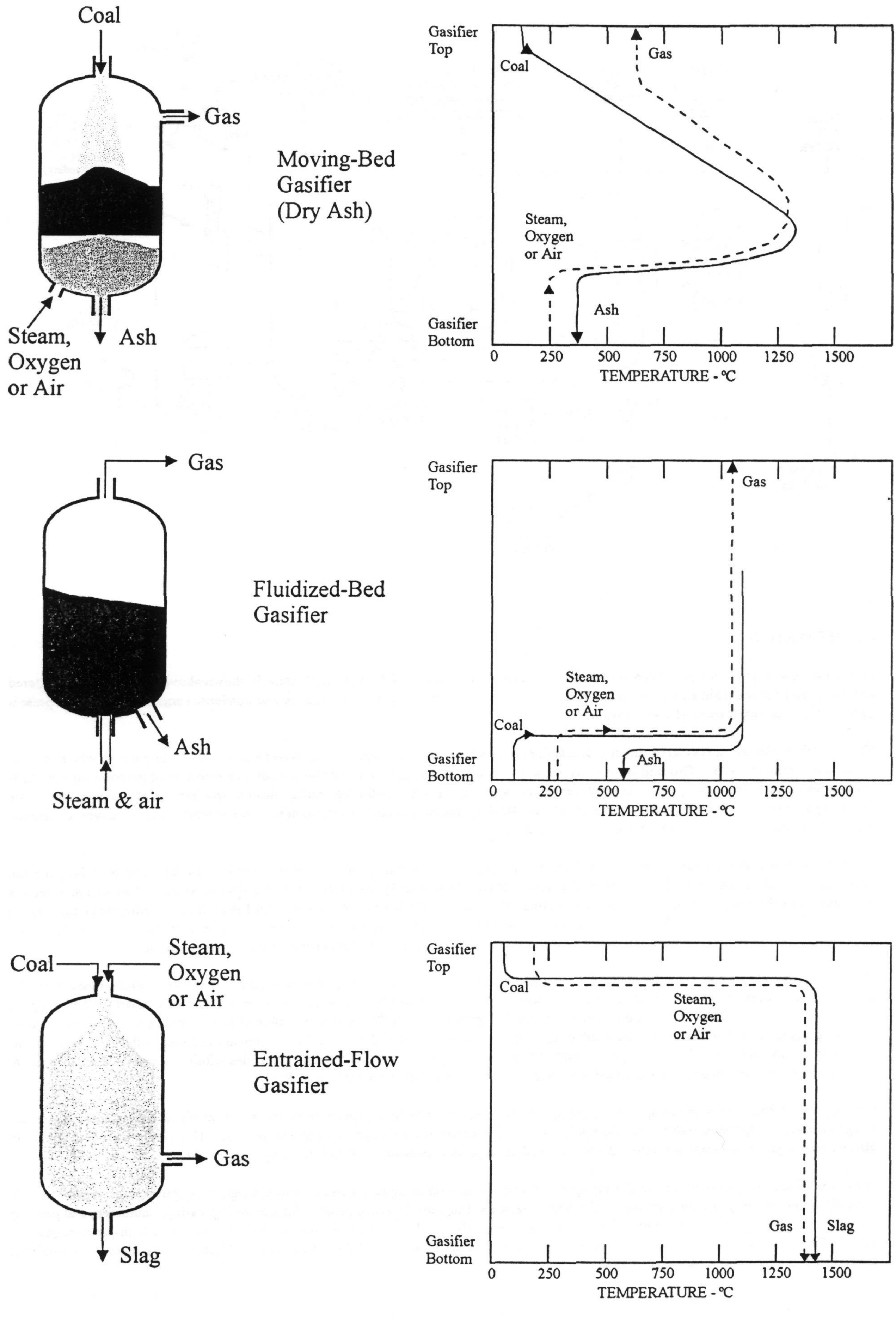
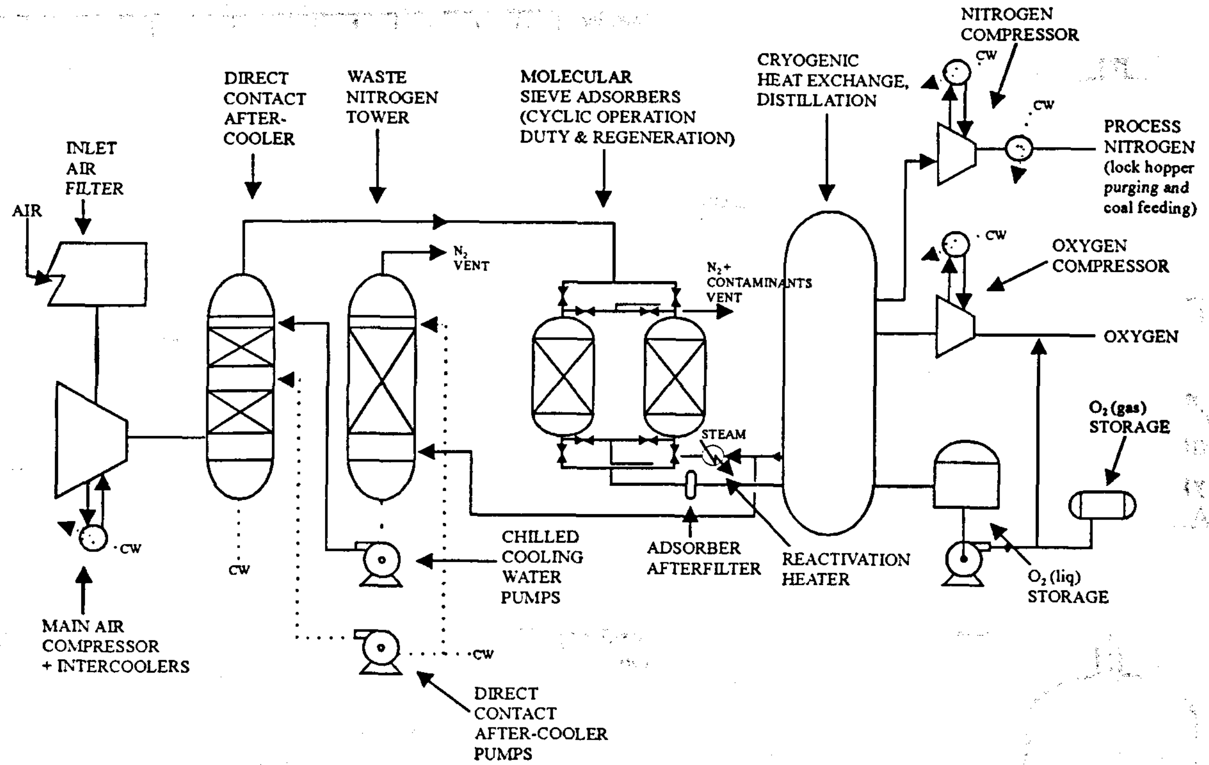


Figure 1 Generic Coal Gasification Technologies



AIR SEPARATION

A process flow diagram for an air-separation plant designed to produce 95% O_2 in a single train is shown above. Ambient air is filtered and compressed in the main air compressor. Water-cooled intercoolers reduce the temperature and condense some of the moisture present in the air between each stage of compression.

The partially cooled air next enters a direct-contact aftercooler where it contacts both cooling water and chilled cooling water, which reduces the air temperature to 40°F. This unit is typically a tower with two packed sections. The hot air enters the bottom of the tower and contacts cooling water. Chilled cooling water is produced in the waste nitrogen tower through humidification and heating of the cold, dry waste nitrogen gas leaving the cold box. Dry waste gas at near 40°F is introduced onto the bottom of the packed waste nitrogen tower, contacting and chilling cooling water which enters the top of the tower.

The residual water vapour, carbon dioxide, and other atmospheric contaminants that could freeze out in the cold sections of the plant are removed from the cooled air exiting the top of the direct contact aftercooler by the molecular sieve absorption unit. Two or more vessels containing absorbent are used in a cyclic process lasting from two to eight hours. While one vessel is on line purifying incoming air, the other vessel is first heated with dry waste gas to remove adsorbed contaminants and cooled to operating temperature before being placed back in services. The reactivation heater uses steam to increase the temperature of regeneration gas to the proper level.

Dry, carbon dioxide-free air from the molecular sieve adsorption unit is cooled to cryogenic temperatures in banks of single-block, plate-fin heat exchangers against oxygen, nitrogen, and waste streams leaving the distillation section. The dry air stream is separated into oxygen, nitrogen, and waste gas by cryogenic distillation. Since low-purity oxygen (95 percent by volume) is required, a variation of the dual reboiler process is used in the cryogenic distillation. Cold boxes surround and support the equipment contained within and insulate the system from heat leak. The boxes are purged with dry nitrogen to prevent condensation of impurities within the insulated enclosure. A portion of the compressed gases is expanded to provide refrigeration to make up for heat leak.

Oxygen, which has been warmed against incoming air, is compressed to final-product pressure in a centrifugal compressor. Since the oxygen is used at high temperature, no aftercooler is provided between the compressor and the pipeline. The nitrogen product stream is also warmed against incoming air before being compressed to product pressure in a reciprocating compressor.

A portion of the oxygen produced in the cryogenic distillation section is liquefied and sent to the liquid oxygen storage tank, which is typically a low-pressure, vacuum-insulated, flat-bottom tank used to store liquefied product for use during startup, shutdown, and peaking requirements. For instantaneous supply of oxygen product, regardless of the cause of an air-separation plant outage, a high-pressure gaseous oxygen (GOX) storage tank is provided to supply 20 min of product flow. The GOX storage tank is filled by pumping and vaporizing a portion of the stored liquid oxygen product.

Figure 2 Non-integrated ASU Flow Diagram

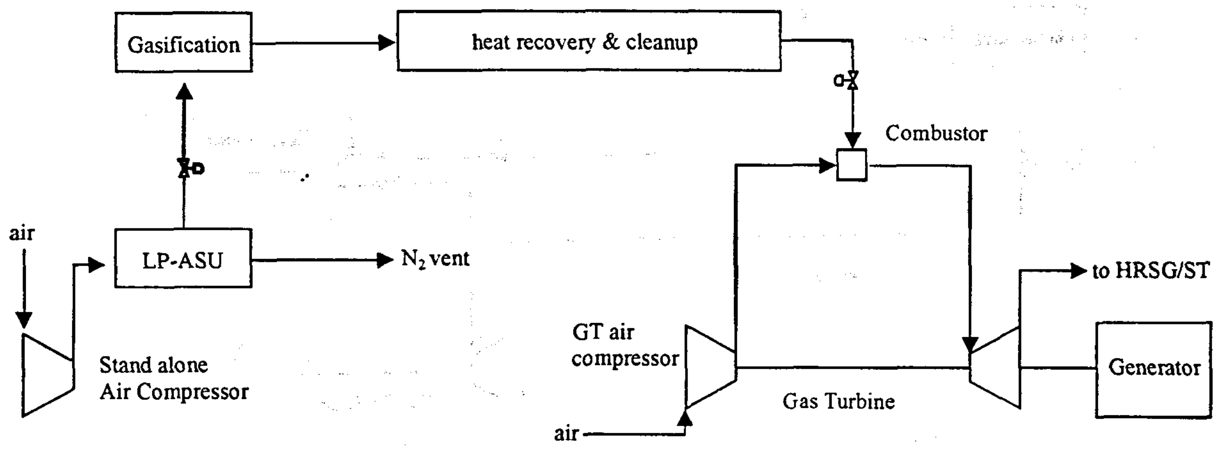


Figure 3a Oxygen Blown GCC with Stand-Alone LP-ASU

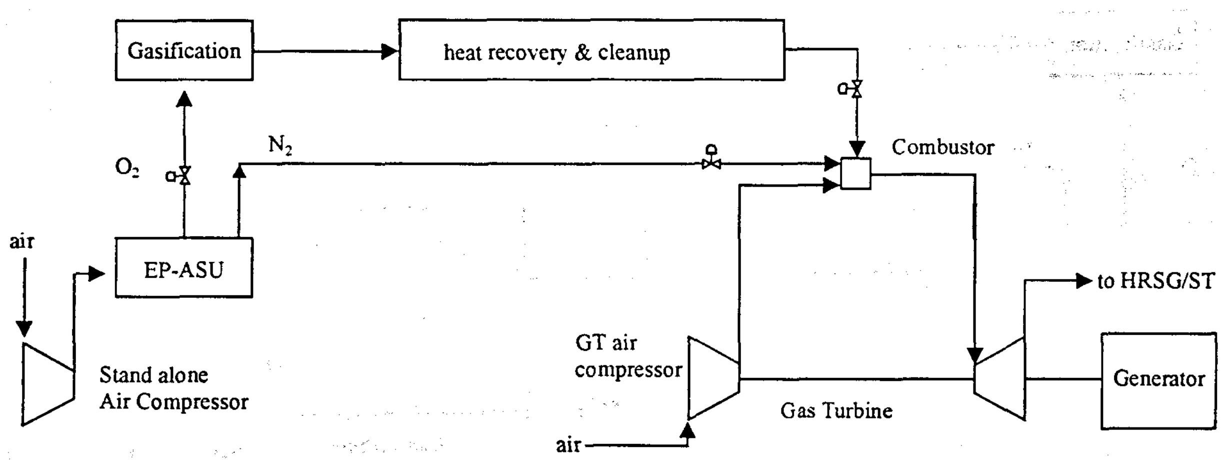


Figure 3b Oxygen Blown GCC with Nitrogen Integrated EP-ASU

Note: LP-ASU = Low Pressure (traditional) Air Separation Unit
EP-ASU = Elevated Pressure Air Separation Unit
HRSG/ST = Heat Recovery Steam Generator and Steam Turbine

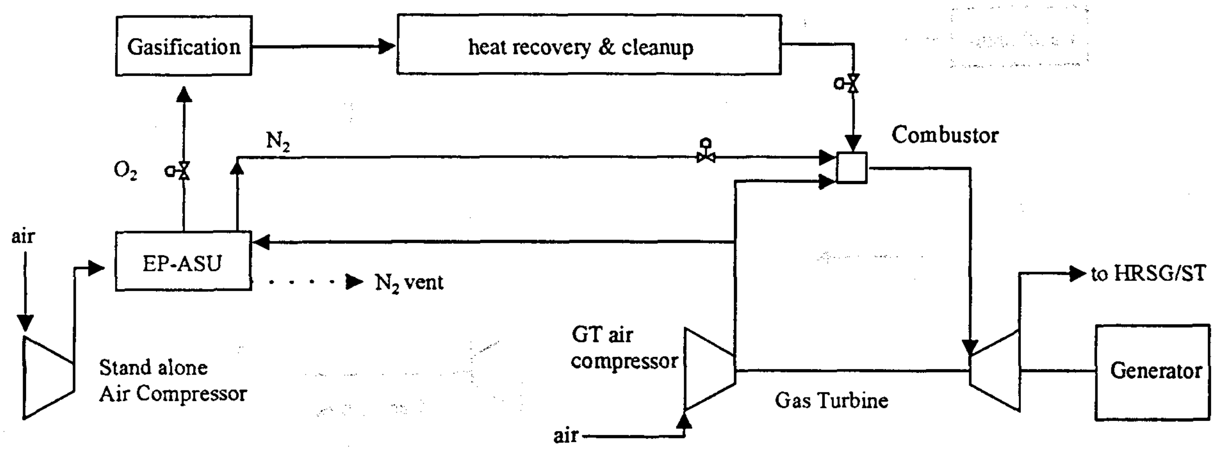


Figure 3c Oxygen Blown IGCC with Partial Air and Full Nitrogen Integrated EP-ASU

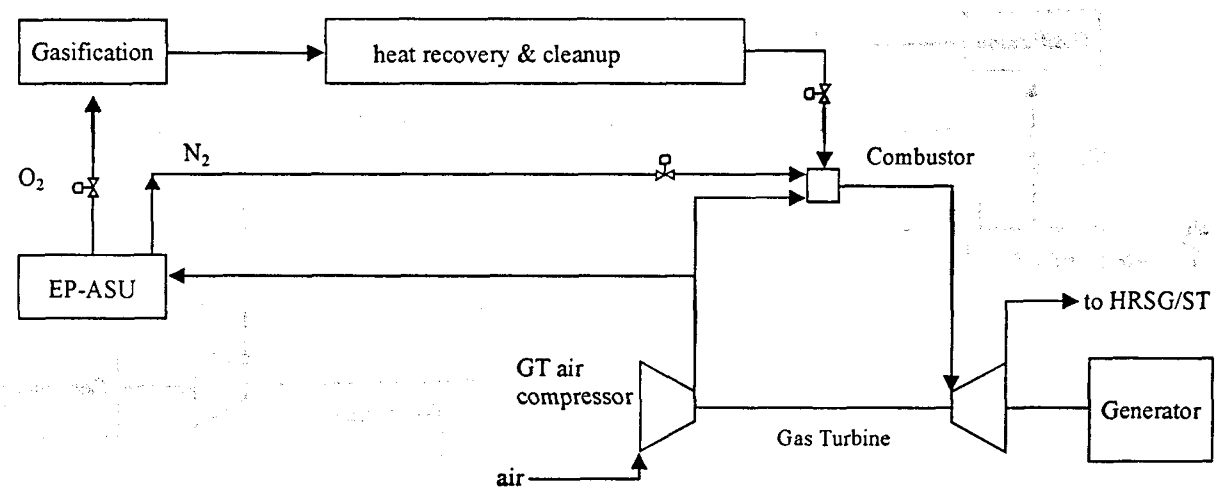
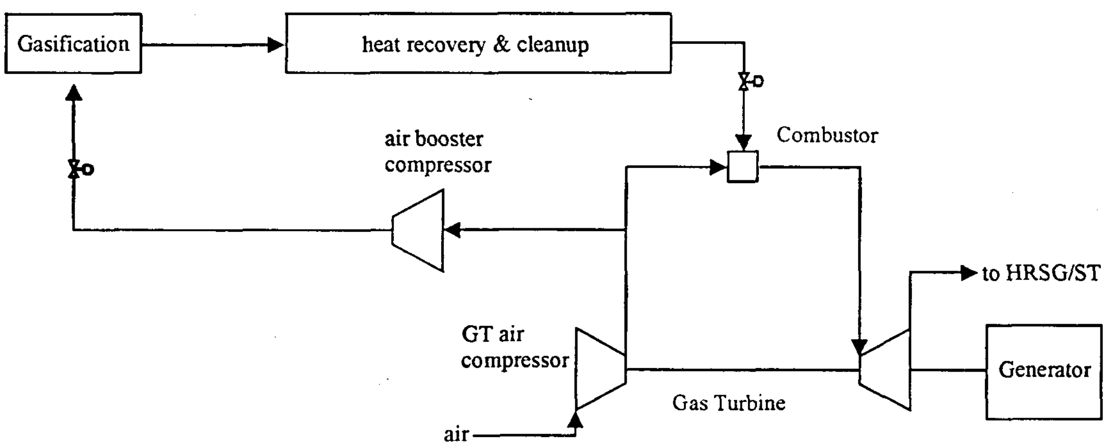


Figure 3d Oxygen Blown IGCC with Full Air and Nitrogen Integrated EP-ASU

Note: EP-ASU = Elevated Pressure Air Separation Unit
HRSG/ST = Heat Recovery Steam Generator and Steam Turbine



Note: HRSG/ST = Heat Recovery Steam Generator and Steam Turbine

Figure 4 Conventional Air Blown GCC