AIR-BLOWN OR OXYGEN-BLOWN GASIFICATION FOR POWER GENERATION?

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ABSTRACT

The majority of coal gasification processes use oxygen, as opposed to air, as the main oxidant. Recent studies, however, have shown that some air-blown gasification combined cycle (GCC) power generation systems offer lower operating costs and superior thermal efficiency when compared with oxygen-blown GCC systems.

Resolving the question of which is best, air-blown or oxygen-blown GCC, is complicated by the wide variety of gasifier configurations available (with possibilities for dry or slurry coal feeding arrangements, combined with options to clean the resulting fuel-gas using relatively novel, but more efficient, methods). Another factor, specific to oxygen-blown gasification, is whether or not to integrate the air separation unit (ASU) with the gas turbine, the integrated scheme resulting in an integrated gasification combined cycle (IGCC) process.

This paper is aimed at resolving this question and reports on a recently completed study that objectively reviewed and compared the current status of air-blown and oxygen-blown technologies. Only published material has been used in this study, and overall the literature illustrates a positive trend favouring air-blown over oxygen-blown gasification. The study concludes that air-blowing is preferable to oxygen-blowing, in terms of:

- plant efficiency, and hence CO_2 emissions;
- capital and operating costs;
- ease of power plant control.

Assessing cycles on a normalised basis, air-blown GCC is found to have ~1.8 percentage points inherent thermal efficiency advantage over oxygen-blown GCC.

Plant cost estimates vary widely, but those based upon entrained-flow oxygen-blown gasifiers are consistently the highest while those based upon air-blown fluidised beds are the lowest. Overall, air-blown gasification is favoured. The estimated cost of electricity is lower with air-blown gasification, as might be expected from the lower plant costs and higher cycle efficiency. Predicted electricity costs are around 4.6USc kWh^{-1} for air-blown, and 5.8USc kWh^{-1} for oxygen-blown, GCCs.

The ease of plant control is strongly dependant on the level of complexity of a GCC power plant. Hence, IGCC plant is the most difficult, oxygen-blown GCC using a non-integrated ASU is less so, with air-blown GCC being the 'easiest'.

BACKGROUND

To date, oxygen-blown systems have been adopted in projects supported by both the US Department of Energy's Clean Coal Technology program (eg Tampa, Tom's Creek and Pinion Pine) and the European Commission's Thermie programme (eg Puertollano and the British Gas/Lurgi gasifier at Westfield in Scotland). In fact, all 17 commercial worldwide coal gasification plants use oxygen rather than air, although this is mainly because the fuel-gas is used as a feedstock for chemical production rather than

/power generation. For chemical production, oxygen-blown gasification is desirable because it • maximises the concentration of hydrogen and carbon monoxide in the fuel-gas, ie it gives a high 'cold gas' conversion efficiency. For power generation, 'cold gas' efficiency is largely irrelevant, since 'sensible' as well as potential energy can be converted into'electricity via steam and gas turbine cycles.

A factor specific to oxygen-blown gasification is whether to integrate the ASU with the gas turbine, thereby creating an IGCC process. This improves cycle efficiency, by reducing the parasitic power consumption, and reduces costs. However, it also increases operational complexity, which can cause problems with plant start-up times and availability.

This paper reports on a study commissioned by the UK Department of Trade and jndustry's Coal R&D programme, which ETSU manages. The study 'Air-blown Versus Oxygen-blown Gasification' (COAL R089) had the objective of assessing available technical and financial information on air-blown and oxygen-blown GCCs and to compare them on a normalised basis, the findings to be used to steer the Programme's support of gasification technologies. British Coal Corporation's Coal Technology Development Division, based at Stoke Orchard, was contracted to undertake the study.

Objectivity in this study was ensured by only using material available in the open literature. Data were drawn for comparison from various sources, including papers from EPRI gasification conferences, reports produced by IEA Coal Research, a NOVEM (Netherlands Agency for Energy and the Environment) study, and papers from ASME/IEEE International Power Generation Conferences and relevant IMechE conferences. A particularly relevant source of information was a Bechtel study where four gasification schemes were considered for use in India using run-of-mine high ash coal to produce electricity. This study compared air-blown and oxygen-blown GCCs for a number of types of gasifier.

No published study covers all possible GCC configurations. Apart from air-blown versus oxygen-blown operation, the three generic gasifier types have options of dry or slurry feeding systems and a choice of cold (wet) or hot (dry) fuel-gas cleaning. The normalised basis chosen for the reported study involved dry coal feeding, hot fuel-gas cleaning and, in the case of oxygen-blown gasifiers, a non-integrated ASU (this is academic for some gasifier designs where slurry feeding is used to convey the coal and therefore not an independent variable). This basis enabled an evaluation of air-blown and oxygen-blown schemes. The following criteria were used to evaluate the cycles:

- cycle efficiency;
- capital costs including consideration of ASU developments;
- electricity generation costs;
- fuel flexibility;
- load following ability and ease of operation;
- environmental performance;
- planned future development.

Availability and maturity have nut been used for the evaluation even though they are very important when considering which technology is best suited for the particular application. The reason why they have been ignored is that there are very few GCC plants in operation worldwide. Over the next 5-10 years this situation will have changed with many more GCC power plants in operation and at that time a realistic assessment of these two important factors can be made.

EVALUATION CRITERIA

Cycle Efficiencies

With the cycles 'normalised', the cycle efficiencies of all the air-blown gasifiers are in reasonably close agreement at 46.3%, net electrical output and fuel lower heating value (LHV) basis. The cycle efficiencies of the oxygen-blown gasifiers are also in reasonable agreement at 44.5%. On the normalised basis, air-blown gasification offers typically 1.8 percentage points advantage over oxygen-blown gasifiers equipped with non-integrated ASUs. Integrated ASUs generally show a 1.1 percentage points gain over non-integrated systems, leaving air-blown gasification with a clear efficiency advantage of about 0.7 percentage points. Comparing an air-blown, dry coal feed, gasifier equipped with hot fuel-gas clean-up, with a conventional oxygen-blown GCC arrangement, where the gasifier receives a dry coal feed but has cold fuel-gas clean-up, the advantage to air blowing is some 2.6 percentage points.

All cycle efficiencies were calculated at constant temperature (Temperature?). All cycles will be detrimentally effected with increases in ambient air temperature however oxygen-blown schemes are likely to be most detrimentally effected because of the penalty associated with cooling the ASU and the liquid-gas storage facility.

<u>Efficiency Advantage of Simplified Hot Gas Clean-up</u>

The simplest form of hot gas clean-up, as originally developed for the UK Air Blown Gasification Cycle (ABGC), relies upon fuel-gas desulphurisation by in-bed feeding of limestone. The only downstream clean-up process is particulate capture using a ceramic filter. This simple arrangement saves around 0.5 percentage points of cycle efficiency over more advanced hot fuel-gas cleaning methods which also remove hydrogen chloride and gaseous nitrogen compounds. The cycle efficiency advantage of the ABGC with simplified hot gas clean-up, compared to an oxygen-blown GCC with cold gas cleaning, is —3.1 percentage points.

Capital Costs

Most recent projections of future costs for IGCC plants, such as are being installed at Buggenum, the Netherlands, and Puertollano, Spain, are US$1700 $\mathrm{kW_c}^{1}$ (1995 dollars) for a $500\mathrm{MW_C}$ (net electrical output) plant and US$1870 $\mathrm{kW_c}$"^{1} (1991 dollars) for a $300\mathrm{MW_C}$ plant, respectively. These figures should be a good guide to entrained flow, oxygen-blown, IGCC plant costs, since they represent predictions based upon installations already (or nearly) constructed. GCC plants with integrated ASU are expected to show a 2-3% cost advantage compared with non-integrated plants.

A 1990 study of air-blown and (non-integrated ASU) oxygen-blown Kellogg-Rust-Westinghouse (KR.W) fluidised bed gasifiers is arguably the best indication of relative capital costs. This study goes into considerable detail and predicts costs for $400\text{-}450\mathrm{MW_e}$ air-blown and oxygen-blown installations of US$1042 $\mathrm{kW_c}^{1}$ and US$1218 $\mathrm{kW_c}$~ respectively. These costs are lower than would normally be expected because they are based upon the development of an existing electric utility site already planned for extension (Plant Wamsley site of Georgia Power Company). Unfortunately, the study unintentionally penalises the oxygen-blown case by omitting a carbon burn-up cell. If the oxygen-blown case is compensated by an increase of 2 percentage points of cycle efficiency, the additional net power output reduces the plant's specific cost to US$1160 $\mathrm{kW_e}^{1\backslash}$ The air-blown KRWs specific cost represents 90% of the oxygen-blown cost.

ASU Developments since 1980

Since 1980, oxygen-blown gasification has benefited from the development of ASUs specifically suited to GCC installations. ASU suppliers have worked with gasifier developers to engineer cost-effective and power-saving versions of their ASUs, cutting away peripheral equipment normally included. The purity of the oxygen has also been allowed to fall, often to 95% and sometimes to 85% (eg Puertollano, Spain).

A 1980 ASU design typically produced 98% purity oxygen and had a capital cost of US$28,200 (1994 figures) per US ton of oxygen per day. A 1994 ASU having an oxygen purity of 90-95% has been estimated to cost almost exactly 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 modem 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%.

Electricity Generation Costs

Electricity generation costs favour air-blown gasification, as would be expected from the lower plant costs and higher cycle efficiencies. For studies involving European and American coals die cost of electricity is estimated around 4.6USc kWh^1 for air-blown gasifiers and 5.8USc kWh^1 for oxygen-blown gasifiers. Absolute values are affected by the economic ground rules (including die land, labour, capital, fuel costs, etc) applicable to the location considered, but the relative benefit of air-blown over oxygen-blown gasification is clear. The Bechtel study estimated consistendy higher electricity prices than other studies using die same GCC schemes, due to the efficiency loss attendant on using high ash coal. Nonedieless, air-blown gasifiers again had lowest electricity costs, with the air-blown fluidised bed being lowest of all, at 5.6USC kWh'^{1} (the Shell GCC was calculated to 7.9USc kWh"^{1} and the Texaco GCC 9.3US? kWh"^{1}).

Fuel Flexibility

All GCC technologies show good fuel flexibility, being able to handle a range of feed stock including, coal, petroleum coke, refinery bottoms sewage and refuse derived fuels. Dry coal feed, entrained flow, gasifiers may be the most flexible in terms of being able to receive fuel with a wide range of characteristics, since pre-drying and fine milling brings them into a common physical form, while the operating conditions in terms of temperature and oxygen partial pressure are aggressive enough for even the most unreactive fuels. Slurry fed, entrained flow gasifiers might benefit with physically difficult feedstocks, where forming them into a slurry should ease feeding problems. This advantage is, however, offset by a noticeable cycle efficiency penalty, especially with high ash fuels. Non-slagging gasifiers, such as air-blown fluidised beds, show a cycle efficiency benefit when the feedstock is a high ash fuel, since no heat is lost melting the ash into slag.

Load Following Ability and Ease of Operation

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 UK's Central Electricity Generating Board (no longer in existance) for conventional fossil-fired power stations. The load following requirements were; 5% per minute increase in load between 50-100% load and 3% per minute increase in load between 30-50% load. In this respect, therefore, neither air-blown nor oxygen-blown gasification has yet demonstrated an advantage.

Full integration, or partial integration, of the ASU with the GCC scheme adds additional complexity to the power plant. This has the effect of making operation, especially start-up and shut-down more difficult. A non-integrated ASU, although more expensive in terms of capital and operational costs, Simplifies operation of the GCC plant. The oxygen for the gasifier is effectively 'on tap' in much the same way the coal. Air-blown systems require only compressor air for the gasifier which may be supplied by the gas turbine's compressor or by an independent compressor during start-up and shut-down making it the least complex, and hence improving the operability of the plant.

Environmental Considerations

Although air-blown GCC is consistendy favoured in terms of cycle efficiency, plant costs, and electricity generation costs, it generally achieves these using hot fuel-gas cleaning. Environmental emissions from hot fuel-gas cleaning arrangements under development are not as good as can be achieved with cold fuel-gas cleaning with respect to NO_x, HC1 and other gaseous emissions. Advanced hot-gas 'polishing' technologies offer improvements in the environmental emissions from air-blown systems - making diem comparable to oxygen-blown systems.

Planned Future Development

Current technology developments will provide an efficiency boost for both the gas turbine and steam turbine parts of the combined cycle. The ABGC readily offers increases in both the steam cycle efficiency and gas turbine cycle efficiency. Whereas, other gasifier designs, both air-blown and oxygen-blown, may be limited to gas turbine cycle improvements. A 1.2 percentage point enhancement in cycle efficiency would possible with the use of supercritical steam conditions.

DISCUSSION

The comparison of 'normalised' cycle efficiencies shows that air-blown GCC with hot gas clean-up has a significant advantage of oxygen-blown systems. Table 1 shows the normalisation indices deduced from the various studies. The effect of different GCC schemes on thermal efficiency are considered in more detail in Table 2 along with an environmental summary.

Table 3 clearly shows that air-blown GCCs have a higher thermal efficiency the oxygen-blown systems when compared on a normalised basis. Integration of thee ASU has the effect of increasing the efficiency of the oxygen-blown system but does not bring it up to the level of the air-blown systems. Figure 1 compares the effect of using hot gas clean-up on GCC systems - clearly hot gas clean-up offers advantages to both air-blown and oxygen-blown systems. The figure also shows the effect of using cold gas cleaning and of slurry feeding. Both effects are detrimental to the efficiency of the plant but again air-blown gasification still has the advantage over oxygen-blown systems.

Relative capital costs are reduced by air-blown gasification not requiring a cosdy ASU, but are increased due to the larger volumes of the fuel-gas (the fuel-gas contains large quantities of nitrogen). In addition, specific gasification rates are lower with air blowing, which also increases the relative size and cost of the gasifier. Conventional dunking anticipates that savings on the ASU will equitably offset the increased cost of the gasifier and fuel-gas vessels - this study has establish that this is not the case. The additional cost of the ASU out-weighs any of the additional costs associated with air-blown plant. An integrated ASU is more advantageous than a non-integrated ASU because it reduces the capital costs associated with the ASU and increases overall thermal efficiency. Full integration has the disadvantage of increasing plant complexity and is likely to make operation difficult compared to non-integrated systems. Table 3 shows the specific and electricity costs for five GCC plants normalised to $300MW_C$, it shows that the two air-blown GCC schemes cost less and produce lower cost electricity.

It has been shown that all GCC technologies can handle a range of fuels. However, air-blown fluidised bed GCC technologies are able to handle high ash fuels that would be difficult in any other system. This makes them particularly suitable for use at, or near, a mine as minimal coal preparation is needed. In addition, the size range of coal suitable for this technology is wider than for oxygen-blown entrained flow or fixed-bed gasifiers.

Environmentally the air-blown schemes are not as inherendy 'clean' as the oxygen-blown systems, this is because of the use of cold gas clean-up on oxygen-blown GCC plants. Air-blown GCCs can be made to be clean as oxygen-blown GCCs by using cold gas clean-up although this will cause a decrease in plant efficiency. Nevertheless, air-blown systems with cold gas clean-up are still more efficient than oxygen-blown systems, making them more acceptable when CO_2 emissions are considered. In the longer-term hot gas clean-up will be developed to a stage where it is as clean as cold gas clean-up.

Air-blown GCC schemes will benefit from advances in both the steam and gas turbines - whereas oxygen-blown schemes are able to benefit only from advances in gas turbine technologies. This is especially true for plant where mere is a separate combustor for the residual char (eg ABGC and High Temperature "Winkler) where high quality steam can be produced giving an option for a supercritical steam cycle to be introduced into the scheme. The air-blown system ako has the added advantage of being able to handle 'difficult' fuels, from sewage to high ash coals.

CONCLUSIONS

- The cost of electricity from air-blown GCCs is some 20% cheaper than from oxygen-blown GCCs.
- Capital costs for air-blown GCCs are around 90% of non-integrated oxygen-blown GCCs.
- Air-blown GCC schemes are inherendy less complex than oxygen-blown GCCs - what ever the level of integration.
- Air-blown GCCs have has an inherent 1.8 percentage point cycle efficiency advantage over oxygen-blown GCCs.
- Air-blown GCC schemes are likely to benefit from advances in gas turbine technology.
- Air-blown GCC schemes will benefit from increase steam conditions if they include a high-temperature heat recovery zone (such as a char combustor).
- Air-blown GCC schemes, especially fiuidised bed-based schemes, are particularly well suited to high ash content coals.
- Air-blown GCC schemes are as environmentally acceptable as oxygen-blown GCC schemes, and more so when CO_2 emissions are considered.

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BIBLIOGRAPHY

Adlhoch, W. et al. *From Research to Reality: the KoBra IGCC Project.* Sydney, Australia: IEA/OECD Conference. 18-21 November, 1991.

Dawes, S.G. et al. *Air-bloum Gasijication for Power Generation.* London: Institute of Chemical Engineers Conference 'Gasification: an Alternative to Natural Gas'. November 1995.

De Winter, H.M.J., Willeboer, W. *Design, Construction and Startup of the 250MWe IGCC Plant at Buggenum.* San Francisco, CA, USA: 13th EPRI Conference on Gasification Power Plants. Session 2, October 1994.

Fisher, MJ. *Air-bloum Versus Oxygen-blown Gasification.* Coal R&D programme Report No. COAL R089. 1996

Gallaspy, D.T. and Tsatsaronis, G. *A Comparison of Alternative IGCC Power Plant Designs: Air-Blown IGCC unth Hot Gas Clean-up vs Oxygen-Blown IGCC with Cold Gas Clean-up.* San Diego, CA, USA: Joint ASME/IEEE International Power Generation Conference. October, 1991.

Good, D.W. and Olson Jr, R.R. *Air Separation Unit Cost Reduction.* San Francisco, CA, USA: 13th EPRI Conference on Gasification Power Plants. October 1994.

Higginbotham, E.B. and Motter, J.W. *Pinon Pine Power Project: Project Status.* San Francisco, CA, USA: 13th EPRI Conference on Gasification Power Plants. October 1994.

Ishikawa, H. et al. *Japanese Current Status of IGCC Development: Clean and Efficient Power Generation From Coal for the Coming Decades.* VGB Kraftwerkstechnik 72, No 5, 1992.

Iyengar, R.K. et al. *Use of High Ash Indian Coal for IGCC Plant.* Palo Alto, CA, USA: 9th EPRI Conference on Gasification Power Plants. October 1990.

Klosek, J. et al. *Air Versus Oxygen for Gasification Combined Cycle Power.* VGB Kraftwerkstechnik 74, No 12, 1994.

Maude, C. *Advanced Power Generation: A Comparative Study of Design Options for Coal.* London: IEACR. Report IEACR/55. March 1993.

Minchener, A.J. *Development of the CRE Gasifier for Use in Integrated Partial Gasification Combined Cycle Systems.* San Francisco, CA, USA: 12th EPRI Conference on Gasification Power Plants. October 1993.

Schippers, K. et al. 'KoBra Will Demonstrate High Temperature Winkler IGCC. *Modern Power Systems.* February 1993.

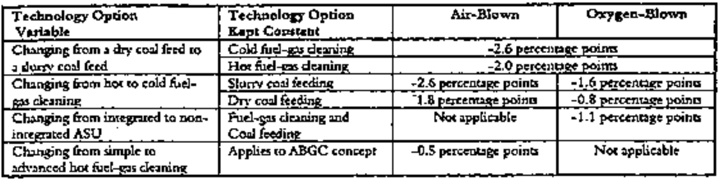
Sendin, A. et al. *Puertollano IGCC Project: Status and Research Work to Improve the Current and Future IGCC Plants.* San Francisco, CA, USA: EPRI Conference on New Power Generation Technology. October 1995.

Sendin, A. et al. *The Puertollano IGCC Project: a 335MW Demonstration Power Plant for the Electricity Companies in Europe.* San Francisco, CA, USA: 13th EPRI Conference on Gasification Power Plants. Session 2, October 1994.

Takematsu, T. and Maude, C. *Coal gasification for IGCC Power Generation.* London: IEACR. Report IEACR/37. March 1991.

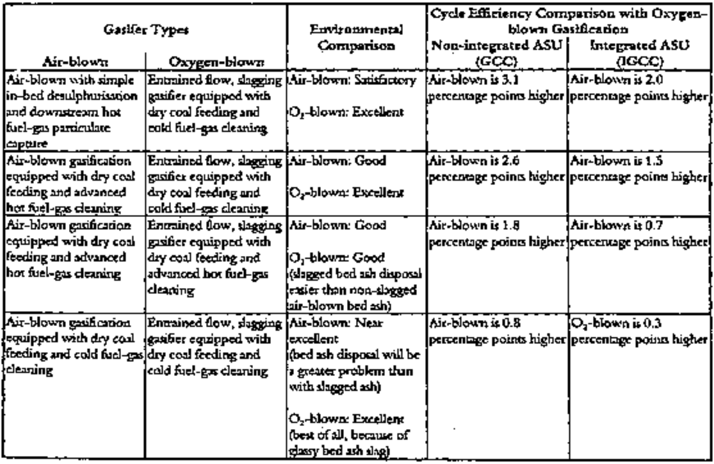
Todd, D.M. and Joiner, R.J. *IGCC Cost Study.* San Francisco, CA, USA: 13th EPRI Conference on Gasification Power Plants. October 1994.

Yanome, S. et al. 'Development of Coal Gasification Technology'. *IHI Engineering Review.* Vol 25, No 1/1-6. January 1992.



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 and Coal feeding	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 1 Cycle efficiency percentage point variations dependent upon technology options



Gasifer Types		Environmental Comparison	Cycle Efficiency Comparison with Oxygen-blown Gasification	
Air-blown	Oxygen-blown		Non-integrated ASU (GCC)	Integrated ASU (IGCC)
Air-blown with simple in-bed desulphurisation and downstream hot fuel-gas particulate capture	Entrained flow, slagging gasifier equipped with dry coal feeding and cold fuel-gas cleaning	Air-blown: Satisfactory O_2-blown: Excellent	Air-blown is 3.1 percentage points higher	Air-blown is 2.0 percentage points higher
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: Good O_2-blown: Excellent	Air-blown is 2.6 percentage points higher	Air-blown is 1.5 percentage points higher
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: Good O_2-blown: Good (slagged bed ash disposal easier than non-slagged air-blown bed ash)	Air-blown is 1.8 percentage points higher	Air-blown is 0.7 percentage points higher
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) O_2-blown: Excellent (best of all, because of glassy bed ash slag)	Air-blown is 0.8 percentage points higher	O_2-blown is 0.3 percentage points higher

Table 2 Air-blown versus oxygen-blown gasification

Gasifier type	Air-blown with steam		Oxygen-blown with steam		
	Fluidised-bed, dry feed, hot gas clean-up (eg ABGC)	Fluidised-bed, dry feed, hot gas clean-up (eg KRW)	Moving bed, dry feed, cold gas clean-up (eg British Gas/Lurgi)	Entrained flow, dry feed, cold gas clean-up (eg Shell)	Entrained flow, slurry feed, cold gas clean-up (eg Texaco)
Net power output (MW_e)	416.6	243.5	213.2	251.4	260.9
Cycle efficiency (%)					
HHV	45.4	42.2	38.2	40.9	38.8
LHV	47.6	44.3	40.1	42.8	40.7
Total capital (US$ millions)	524.3	349.1	404.6	491.1	499.1
Specific cost (US$ kW_e^{-1})					
Non-adjusted	1259	1434	1933	1954	1913
Normalised to 300MW_e	1380	1353	1746	1859	1840
Electricity cost (US¢ kWh^{-1})	4.21	4.75	6.00	5.94	5.95

Table 3 Cost comparison of five GCC systems

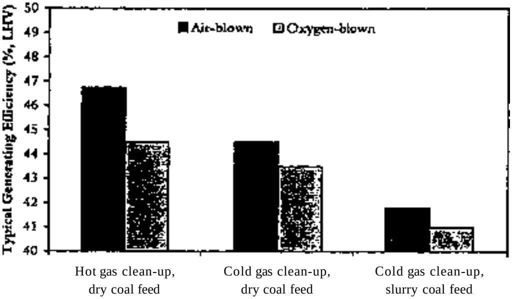


Figure 1 Comparison of Cycle Efficiencies

IGCC - TREADING THE PATH BETWEEN OPTION AND ACCEPTANCE

(or: WE ARE ALL IN THE SAME BOAT)

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INTRODUCTION

When, at the end of the War of Liberation in 1815 Napoleon arrived at the French camp at Waterloo the traditional salute was not fired. When he asked why, one of his Generals replied as follows:

- Many soldiers have died or have been taken prisoner
- We did not know exactly when you were arriving
- Our soldiers are totally exhausted
- The Lieutenant responsible is reported missing
- We have run out of ammunition

Somewhat annoyed, Napoleon interrupted his General's outpouring of excuses:

"Thank you General, the last reason would have sufficed!"

This anecdote has a direct connection with our conference, because ...

DEFINITION OF SUBJECT

... the advance notice for this year's "Gasification Technology in Practice" Conference mentioned nine separate questions to be discussed at the Conference. However, only one of them is really relevant:

"Reliability, availability and funding in the market".

This title provides a cover for an extremely sensitive problem area, namely acceptance of the technology and its introduction to the market.

The answers to all other questions can be briefly summarised as follows:

- There are gasifiers available for all feedstocks such as solid, liquid and fossil fuels, and waste products.

- There is a wide range of proven gas cleaning methods available for all kinds of untreated gases and which meet stringent purity requirements.

- Clean gases can converted using known technologies for use in any number of applications in the chemical industry, iron ore reduction processes, heat and power generation.

- There is no such thing as the most economic gasification process for universal use. On the contrary, an individual, optimum process has to be worked out for each application, by means of a study taking account of

  - feedstock
  - quality requirements for clean gas
  - quality of waste products
  - operating characteristics

- site-specific conditions
- environmental legislation
- and other significant parameters

All current consumption needs can be met using the possibilities available today. It is generally only engineering companies who have the best overview about the full range of various process alternatives and variants available. They alone are therefore in the position to establish the best process variant for a particular application.

Nevertheless engineering companies are also making the discovery that detailed process studies and optimisations are not enough to break down the reservations and inhibitions of prospective process users towards innovative technologies. As long as there are no hard proofs that the techno-economic requirements can be fulfilled - and this is what the process users believe - potentials and options are to be regarded merely as interesting but unreliable hypotheses. The prototype plants in Wabash River, Tampa, Buggenum or Puertollano fail to provide them with sufficient confidence that initial teething problems can be solved. Very often this reluctance stems from a need to protect oneself from a fear of the unknown.

As is so often the case, there are two sides to the coin here too: the seller and the buyer. The seller is offering the customer bananas, but he does not like bananas and prefers oranges instead.

After this detailed introduction to the problem the following questions arise:

1. Why should a client, contrary to his interests at this point in time, buy innovative IGCC plants, when conventional technology satisfies his requirements?

2. How can the client be won over to the benefits of innovative IGCC technology?

SOLUTION TO PROBLEM
Energy consumption scenario

Regarding question 1 - why?
Fig. 1 shows UNESCO's forecast of the world population development for the period 1850 - 2120. Between 1850 and 2000 the earth's population grew from 1.3 to 5.5 billion people. In the next 50 years it will double again to 11 billion. Europe, North America and the former Soviet Union will only take a small part in this process, with the greater population development taking place in Eastern and Southern Asia, Africa and South America.

Current world energy consumption is 13 TW, of which about 50% is used by 15% of the world's population in Europe and the USA (see Table 1). On average the per capita consumption in these regions is about 6.25 times higher that in the rest of the world. Over the entire globe the spread ranges from 80 W per person in the poorest countries to 11,000 W per person in the USA. Average per capita consumption in Europe is roughly in the middle at 6,000 W.

According to Professor Durr[1], member of the Club of Rome, the limit at which in all probability there is likely to be no damage to the biosphere is around 9 TW. Today, however, world-wide consumption of 13 TW is already about 45% too high. And this does not take account of the fact that

- Developing countries are striving for a higher standard of living associated with greater energy consumption and

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[1] Den unbekannten Pfad der Zukunft betreten. Int. Kongr. des Forum Verlages 8.-10.11.96

- a doubling of the world population must be expected even within the next fifty years.

Presupposing the current energy consumption of the industrialised countries across the board for the entire world population in 2050. this would exceed-the limit for the biosphere (approx. 9 TW) established by Prof. Durr by a factor of 19.

These figures show that current energy consumption, which is already too high, will increase even more substantially in the future. The next generations, and the poorest countries in the world, would quite rightly question our credibility if there is no change. But it is precisely this credibility that we need if we want to open up the markets of the industrialising countries.

The challenge to our generation is not only to cut back its energy consumption but over and above this, to use it more efficiently. In this conjunction "cut back" means to restrain energy consumption where it is not absolutely essential, and "use more efficiently" means to convert primary energy to achieve the highest possible yield. The costs of energy conversion may only take second place to the concerns of environmental protection and conservation of resources. All the wealth in the world is of no use to us one day if our resources have been exhausted.

To recap: if the maximum load on the biosphere restricts per capita consumption to 1.5 kW (> 13 TW total consumption), we Central Europeans have to reduce our requirement of 600 W by 75%. This is possible, according to Prof. Durr, if we reduce end-user energy consumption by 50% and double the efficiency of end-user energy production. Using the technical facilities we have at our disposal today, and which we will have all the more in the future, we would revert to the standard of living enjoyed by Switzerland in 1961. And no-one will seriously claim that this standard of living was bad!

The appeal made by the next generation and the poorest countries in the world is therefore:
"More economic and more efficient use of energy".
Is then this demand unsocial or even unfair? Hardly! Not even if the deregulation of the electricity sector with all too cheap natural gas is hindering the innovation of coal technologies. This impediment can only be short-lived, as resources will be used up far quicker in the wake of all this cheap natural gas than currently forecast for static consumption. At an annual energy consumption of 13 TW

Gas will last for another 65 years
Oil for another 45 years and
Coal for another 200 years.

If consumption is multiplied - the borderline case may be about 20 times in 50 years - the range of resources available will be reduced accordingly.

The research and development sector is being charged with a specific task here, and that is:

> To develop processes for more efficient utilisation of primary energy. Any extra cost involved is simply the sacrifice we have to make for the benefit of the next generations and the poorest countries in the world.

Ladies and gentlemen, we have already carried out a major part of this directive. Excellent preliminary achievements have been made; only they fail to find sufficient recognition, particularly by power plant operators. The potential is certainly acknowledged, as it cannot be objectively denied (see Fig. 2). Although Carnot's theorem does not allow the maximum achievable efficiency rate in power plant processes to exceed the Carnot point (curve in bold print), it is possible to get as close to it as you wish with increasing technical effort. An increase in mean working temperature and an improved process design help to increase the maximum achievable efficiency.

The comparison of conventional and innovative IGCC power plant technology shows that in the temperature range implemented today IGCC can achieve an efficiency that is 5 percentage points higher than a conventional power plant; and in the future, with 80% approximation to the Carnot point, a total of 15 percentage points, equivalent to about 25% greater efficiency. Compared with the current standard in power plants (45%) a 50% better energy utilisation is actually achieved, and this is only possible using innovative technology. Prof. Durr's requirement therefore is not unrealistic.

And the costs?

Assuming that a conventional power plant costs 900 US$/kW and has an efficiency of 40%, then in the USA it produces electricity at a price of 100% (see Fig. 3). Innovative power plants with a higher efficiency, approximately 45, 50 or 55%, are then located in a parameter field whose upgrade depends on the specific plant investment. Power plants with higher efficiency than the conventional power plant might as a consequence be specifically more expensive. How much dearer they might be (acceptable price) can be seen on the abscissa of the intersection between the electricity generation costs for the conventional power plant (COE = 100%) and the costs of the innovative power plant (h = 45/50%).

How high are the actual costs?

Let us look at an IGCC power plant (Fig. 4) which is currently being built in Puertollano, Spain. It has a capacity of 300 MW_{net}, using a very high ash fuel, a mixture of an indigenous coal with 45% ash and a petroleum coke with a very high S content. This power plant, financed by the EU and a total of 8 European power generating companies, will go on stream in the course of 1997.

Gasification is according to the Krupp Koppers (now Krupp Uhde GmbH) PRENFLO principle, and electricity is produced using a Siemens Combined Cycle (CC).

The specific plant investment (see Fig. 5) is 1840 US$/kW, whereas the specific total investment is approximately 3240 US$/kW. Only the last figure is known and is generally considered prohibitively high. And rightly so, for it contains very high Owner's costs of 45%, which cannot be influenced by the plant builder.

Aware of this problem EPRI proposed a standardised economic efficiency calculation, the Revenue Requirement Method, which for example standardises specific investment costs and thus the Owner costs too.

If we apply this method, and also take account of the following:

- a good coal, for instance Pittsburgh No. 8, effects a considerable efficiency improvement
- a normal plant size of 450 MW (instead of 300 MW_{ne}i) is predetermined by the new generation of Siemens turbines, which results in a progressive reduction of unit cost, and
- a further enhancement of efficiency is achieved by optimising the thermo-economy of the IGCC plant

this results in an IGCC power plant which could be ordered as early as 1998 at a specific price of 1436 US$/kW. This is already below the acceptable costs - determined as shown before - of 1600 US$/kW. This means that

Theoretically, IGCC will be economic by 1998 at the latest.

Strategy for entering the market

Regarding question 2 - how can we break into the market?

One cannot help asking why then IGCC technology has not already been introduced on a large scale. The answer has already been given at the beginning. No-one is going to believe the figures until they have been proven by the practical operation of a prototype plant.

How is this to be done?

Here too, experience has provided the key (see Fig. 6). The specific total investment costs for an innovative technology were plotted as a function of the number of units built. In the R&D phase plant costs can only be estimated, and they are indeed always higher than forecast.

In the subsequent commercial testing phase the specific investment costs can be continuously reduced due to increasing familiarity with the process, resulting in improvements and simplifications. Only after more than 4 follow-on plants after the first commercial plant do the specific investment costs reach a more or less final level. In the meantime a period of about 35 - 40 years will have passed.

It is now more than 25 years since the development of IGCC technology was first started in 1971. We are now at the culmination of our learning curve and forecast a positive operating result for the second plant (as previously shown), a result which will be further improved in successive plants. Nevertheless - and this fact should not be concealed - there is a risk of loss of income due to temporary operational malfunctions. This of course puts an incalculable strain on the operating result.

Assuming that the potential of IGCC technology will have to be utilised in the future, then the risks of the follow-on plants also have to be borne - by somebody. There is a lot to be said for the public purse covering these risks, ie. out of general tax income or special taxes and levies. But there is also a lot to be said for these being paid by the established utilities from their profits (already paid for by the consumer) or from a combination of various possibilities. This is certainly worthy of reflection, but not for too long because:

- the demands for environmental protection and conservation of resources are already on the table

- the R&D input for IGCC to date has taken 25 years; in case of an irreversible interruption, this time and the billions in cost will be lost if the necessary consecutive costs are not planned for now.

- given the known contigencies today national governments will be passing laws which, as past experience has shown, will impose considerably more expensive constraints at the wrong time, e.g. by levying energy or environmental taxes and introducing permitting restrictions for new plants, than would be the case if there were to be a voluntary initiative in good time.

Most industrialised countries have very stable electricity grids. Therefore it ought not to be a problem for them to compensate for the risks of plant failure, which lie exclusively in the reimbursement of fixed costs. In industrialised countries therefore the consecutive costs for the remaining development work would be the lowest. The industrialised countries are therefore enjoined to cooperate in the further development of IGCC technology. Only by doing so do they contribute towards:

- environmental protection and conservation of resources

- in the light of anticipated statutory restrictions making provision for their own future competitiveness

and last but not least

- demonstrating their seriousness of purpose and credibility in respect of those who are dependent on our experience and know-how, namely the next generations and the poorest countries in the world.

SUMMARY

1. The compulsion towards an energy technology which conserves the environment and resources is unavoidable, and IGCC plays a key role.

2. IGCC technology is available now. However, the process users still have to gain confidence in its operational reliability.

3. Confidence in this technology will only come through practical operation. The microeconomic risk has to be covered. If the operators alone cannot do this it will fall to the public purse (state) to step in.

4. Industrialised countries have the best opportunities for systematic testing. They not only have a duty to do so (by setting a good example), it is also in their own self-interest (plant operation and sale of such technology).

5. The developing countries have to be interested in the application of innovative energy technology, in order to preserve their own resources and minimise investment requirements.

While the interests of industrialised and developing countries may differ, it is important to remember:

We are all in the same boat.

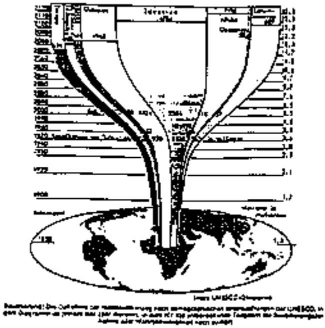


Fig. 1

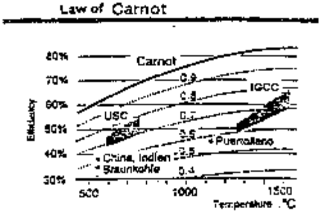


Fig. 2

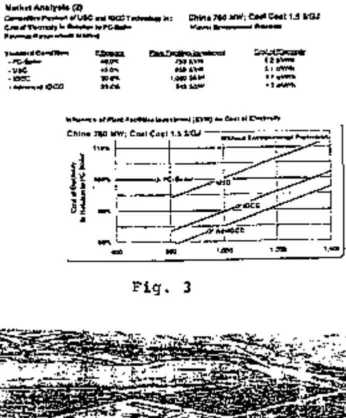
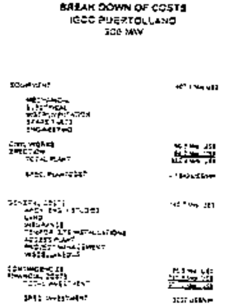


Fig. 3



BREAK DOWN OF COSTS
IGCC PUERTOLLANO
300 MW

Fig. 5

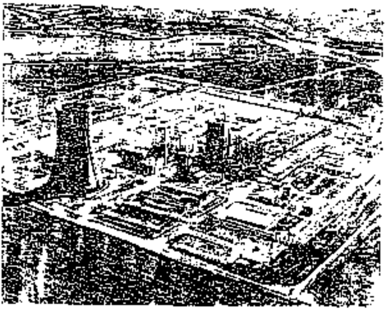


Fig. 4

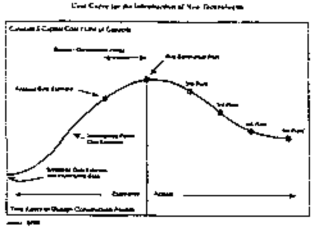


Fig. 6

Tab. 1

	Population [Mrd.]		Energy Consumption [TW]	
	2000	2050	2000	2050
Europe N-America	901	1.146	6.450	?
ROW	4.609	9.454	6.550	?
Σ	5.510	10.600	13,000	?

CONCEPT AND REALISATION OF THE SCHWARZE PUMPE, FRG 'WASTE TO ENERGY AND CHEMICALS CENTRE'

H. Hirschfelder - Lurgi Umweit GmbH, FRG
B. Buttker - SVZ Schwarze Pumpe, FRG
G. Steiner - Mannesmann Seiffert (MSE), FRG

1 HISTORY OF SCHWARZE PUMPE

The Schwarze Pumpe site (the literal translation means 'Black Pump') is located about 150 km south-east of Berlin, Germany.

Large lignite reserves are being found in the area around Schwarze Pumpe. Since lignite represented the only major energy source of the German Democratic Republic, a number of large chemical, coking and power plants was erected in this region. The town gas plant Schwarze Pumpe started production in 1964; with 24 fixed bed gasifiers Schwarze Pumpe finally supplied about 75 % of the total town gas consumption of the GDR. After reunification the towngas from lignite was replaced stepwise by natural gas until 1996, when town gas production as such ceased completely. The question therefore arose of how to best use the existing plants, which were largely in good condition, and how to exploit the extended know-how and long experience of the engineers and scientists of Schwarze Pumpe.

Early on therefore the idea was pursued to use the existing facilities for the conversion of all sorts of wastes, contaminated solids and liquids and other difficult materials into useful products in an environmentally friendly manner. Sucessful tests were carried out immediately upon the granting of the relevant permits from the authorities with such materials as contaminated solids/liquids, sewage sludge, plastics, rubber and other difficult wastes.

In 1996 Schwarze Pumpe - now 'SVZ Sekundarrohstoff Verwertungszentrum Schwarze Pumpe' (Centre for the Re-use of Secondary Raw Materials) was acquired by BWB-Berliner Wasserbetriebs (Berlin Water Authority) from the German Treuhand, the government agency in charge of privatization of former state owned industry.

Recovery of mineral matter in the waste as slag {e.g., in the BGL process)

Environmentally friendly

High thermal efficiency

Feedstock flexibility (solid and liquid wastes)

Favorable economics.

SVZ's choice was based both on the above and on the proven track record of most units as make up the gasification process chain.

3 DESCRIPTION OF THE SV2 PLANT AND OPERATIONAL RESULTS

3.1 OVERALL PLANT DESCRIPTION (Fig.1)

The SVZ plant as operating at present consists of

Receiving bunkers for sewage sludge, plastics, contaminated wood, fluff, waste pellets, others;

Agglomeration facilities for sewage sludge (piston press briquetting) and plastics (extruders);

7 fixed bed gasifiers;

2 types of entrained flow gasifiers;

Partial gas conditioning (CO-shifting);

Rectisol gas purification/sulfur removal.

Until start-up of the new methanol synthesis and combined cycle power plant, the gas is being used as fuel in an adjacent power plant.

3.1.1 DESCRIPTION OF THE GASIFICATION PLANT

Waste gasification represents the heart of the SVZ re-utilization centre of secondary raw materials. Three types of gasifiers are being used in the present configuration.

Solid wastes such as plastics, sewage sludge, rubber, fluff, contaminated wood, residues of paint, household wastes etc. are processed in seven fixed bed gasifiers (figure 2). The gasifiers have an inner diameter of 3.6 m and operate at a pressure of 24 bar. Gasification agent is a mixture of steam and oxygen. Gasifier capacity ranges from 8-14 t/hr depending on feedstock composition. Since fixed bed gasifiers require fuel in lump form (e.g 20 - 80 mm) prior agglomeration - briquetting, pelletizing or some other method - is required for the counter-current operation of descending fuel and rising gasification agent/produced gases.

Liquid wastes - such as tar/oil from above fixed bed gasifiers, used oils, solvents, oil/water emulsions - are gasified in a refractory lined, entrained flow gasifier into which a former fixed bed gasifier has been converted. This type of entrained flow gasifier has been in operation since 1969 for the gasification of tar and oil from lignite coking.

Recovery of mineral matter in the waste as slag (e.g., in the BGL process)

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The former GSP entrained flow reactor for dry lignite gasification has been modified to process liquid feedstocks - e.g. oils and slurries. The reactor is equipped with a cooling system for operation at temperatures between 1600 -1800^{0}C. The ash leaves the reactor in liquid form.

The gases from the entrained flow reactors are partly shifted prior to Rectisol gas purification.

3.2 OPERATIONAL RESULTS

In the years 1992 to begin of 1996 a large number of waste materials (fig. 3) with a total quantity of almost 600.000 have been gasified, with waste plastics and sewage sludge being the main solid feedstocks. Addition of coal (lignite briquettes) was required to comply with the permit issued by the environmental agency, Freiberg/Saxonia.

Late in 1994 discussions between SVZ and Lurgi were resumed (they had already started in 1990) with a view to use the BGL (British Gas/Lurgi) gasifier as replacement for the SVZ grate gasifiers, which were already in successful operation for more than two decades. When compared to the grate gasifiers, the BGL gasifier offers the advantages of double to treble specific throughput, significantly lower gasification steam requirement, production of completely molten slag, possibility of partial/total recycle of tars/oil to the gasifier and other positive features.
Following various stages of discussions and studies, Lurgi in 1996 received the order for the basic and permitting engineering, followed by the order for the detailed engineering and supply including erection and start-up for one BGL-gasifier (phase I of the SVZ extension plan).

'Menus' - mixtures of various wastes - were established by SVZ which would represent the feedstocks to be gasified at Schwarze Pumpe in the future (table 1). To demonstrate their suitability for fixed bed gasification, large scale tests were carried out in 1996. These tests have confirmed that this type of solid wastes can be converted in one step and under elevated pressure - 24 bar - into a crude fuel/synthesis gas.

As mentioned earlier fixed bed gasifiers require feedstock in lump form. Coal and coke are normally available in coarse form, e.g. 20 - 80 mm. Waste materials such as sewage sludge, plastics, household wastes of fine consistency or in rags, however will require agglomeration prior to being fed to the gasifier.

Mannesmann Seiffert/Berlin in 1996 received the order for a plant for pellet production out of household waste, shredder light fraction, plastics and contaminated wood. After removal of iron and other metals, and drying, the 'fluff thus produced is being pelletized on an annular pelletizing press using a binder, such as bituminous coal, molasses or some other suitable material. A binder is required to achieve the necessary strength for the transportation of the pellets and for the necessary thermal stability when the pellets are being exposed to the conditions of the carbonization/gasification zone in the fixed bed gasifier.

In cases where the waste's volatile matter and ash contents are high, i.e. its fixed carbon content is low - as is the case especially for plastics - addition of extra fixed carbon, i.e. coal or coke, is required. Two methods of coal admixture have been tested (table 1): Pelletization with RDF (menu 6) and separate (menus 7 and 8).

3.3 ENVIRONMENTAL ASPECTS

Extensive measurements have been made in the product gas, of the ash and of the process water.

The product gas meets all requirements of the very strict German regulation called
17. BImSch V. For dioxin e.g. values below 0.01 ng/m_n^3 were measured, less than 1/10 th of the
17. BImSchV requirement (0.1 ng/m_n^3).

Likewise, the produced slag meets the very stringent requirements for its disposal according to the German TA-Siedlungsabfall', disposal Class -1 with regard to its leaching properties.
(Hg in leachate: < 0.01 mg/l; Cd < 0.0005 mg/l; Cr < 0.006 mg/l for example).

The process water is suitable for treatment for in the existing SVZ plants and then for release into the local river Spree.

4 COMMITTED EXTENSIONS OF THE SVZ PLANT.

Since reunification, the Schwarze Pumpe complex has undergone a number of major changes. Some plants have already been pulled down, e.g. the coke ovens; others will be replaced by modem ones, such as the 1000 MW lignite fired power plant presently under construction and new plants will be added to serve SVZ's new purpose as centre for the use of secondary wastes.

Those plants include (Fig. 4):

Waste receiving and pelletizing

BGL gasification of various wastes; gas liquor separation

100.000 tpy methanol systhesis

60 MWe gas/steam-turbine combined cycle power plant.

Since methanol synthesis and combined cycle power plant represent well-known, staodard technology, no further description is provided here.

4.1 WASTE RECEIVING AND PELLETIZING (Fig. 4)

Household and similar industrial wastes as well as plastic waste, shredder light fraction and contaminated wood received by rail or road are first crushed in a shredder to pieces below 80 mm. Iron and metals are removed magnetically and electrically, respectively. If required, organic and inorganic residues are removed either for disposal or separate pelletizing, which provides pellets of better quality. The wastes are then dried from about 30 % to below 10 % moisture, mixed with a binder - e.g. bituminous coal or molasses - and fed to pelletizing presses. In the presses wastes and binder are. pressed through a rotating die ring having a large number of 10-20 mm holes, by inner rollers. In this way pellets of 20 - 70 mm length are being produced. Through internal friction pellets reach a temperature of around 100 'C. They are subsequently cooled in a down-stream air cooler to around 30 *C.

The plant will have a capacity of 120.000 tpy of household waste in the first extension step. Start-up is scheduled for 1998. The plant is designed and will be erected by Mannesmann-Seiffert, Berlin (see also 7).

4.2 BGL GASIFICATION (Fig. 2 and 4)

The pellets and other wastes are gasified at a pressure of 27 bar in a BGL fixed bed gasifier, using steam and oxygen as gasification agent. The BGL gasifier has been developed by British Gas, London and Lurgi for the gasification of coals and cokes. It is the only gasifier in which gasification of large particles and vitrification of inorganic matter can be carried out under pressure in the same reactor.

Steam and oxygen are introduced at the bottom through so called tuyeres. Slag is withdrawn automatically - after quenching in the water filled quench vessel - from the slag lock hopper as a

frit of about 3 mm. Crude gas leaves the gasifier at its top with a temperature of between 400 - 500 °C. It's then scrubbed with recycled gas liquor. CO-shifting of the gas going to the methanol systhesis is required to obtain the necessary H_2/CO ratio.
Gas liquor from gas cooling contains tars and oils from the carbonization zone in the gasifiers. After separation by gravity tars and oils are recycled to the gasifier.

In the first step of extension one BGL gasifier with a maximum capacity of 30 t/hr will be installed. The gasifier capacity will vary with feedstock (i.e. waste) properties and composition.
Start-up is scheduled for 1998. The BGL gasifier is designed by Lurgi in cooperation with British Gas, London.

INVESTMENT COST

Currently about DM 320 m (approx USD 200 m) are being invested at Schwarze Pumpe for the preparation and storage of various wastes, the combined-cycle power plant with steam system for other users, process water treatment, methanol synthesis and ancillary units.
Another DM 100 m (USD 62 m) will be required for waste pelletizing and the BGL plant described under 4.1. and 4.2.

At Schwarze Pumpe the situation for the erection of a centre for the use of secondary raw materials is rather advantageous, since a number of plants do already exist. To prove the viability of the BGL gasification concept it is however necessary to look at a 'grass roots' plant.
This question will be addressed in the next chapter.

ECONOMICS

A 'grass-roots' comparison was made on the following basis:

Plant capacity:	400.000 tpy
Coal requirement:	60.000 tpy
Product:	Syngas, HV = 12,000 KJ/m_n^3
Gas Sales Price:	DM 0.025/kW_{lh} (USD 0.016/kW_{th})

Such a plant will be economically viable assuming a gate fee for the wastes of better than DM 200 (USD 124) on average - depending on plant size - can be collected.

The SVZ plant will be the first of its kind for waste utilization by BGL gasification. It is expected that significant savings can be achieved from operation of this first plant mainly in the following areas ('learning curve'):

Simplification of waste preparation/pelletizing

Higher throughputs in BGL gasification

Overall optimization when using only new, purpose designed plants

Further integration with other plants

After realization of the above it can be realistically expected that even plants of smaller capacities will then require gate fees no higher than stated above.

7 COOPERATION SVZ/MANNESMANN-SEIFFERT/LURGI

Based on prior investigations on the concept of waste preparation/BGL gasification the three companies SVZ Schwarze Pumpe GmbH, Mannesmann-Seiffert (MSE), Berlin und Lurgi, Frankfurt, in 1996 signed a co-operation agreement with the following main points:

- Development of the concept, incl. large scale tests
- Realization of a demonstration project in phases at Schwarze Pumpe
- Joint exploitation of the concept by SVZ, MSE und Lurgi

The consortium Lurgi/MSE - with Lurgi as leader - received the order for the erection of a 120.000 tpy waste preparation/pelletizing plant and one BGL gasification plant in 1996. The next phase - another waste preparation/pelletizing and BGL train - will be authorized by SVZ upon agreement on further waste contracts.

8 OUTLOOK

The BGL concept of pressure gasification of wastes provides following advantages:

- High efficiency;
- Disposal of wastes in environmentally friendly manner;
- Production of a high value gas for systheses or combined cycle power generation;
- Robustness and flexibility with regard to waste properties;
- Export opportunities.

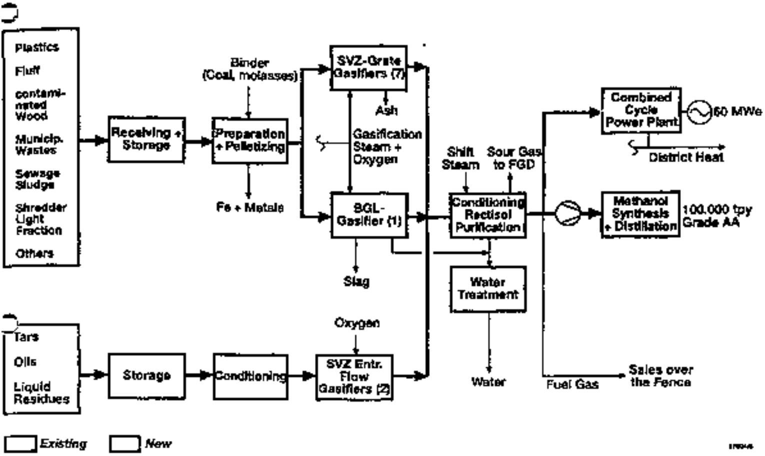
9 REFERENCES

1. Utilization of Secondary Raw Materials by Gasification at Schwarze Pumpe, J. Schneider et al., EPRI-Conference, USA, 1996;
2. Aufbereitung von Hausmull, Dr. L. Plass et al.; Schwarze Pumpe Symposium 1996

Schwarze Pumpe

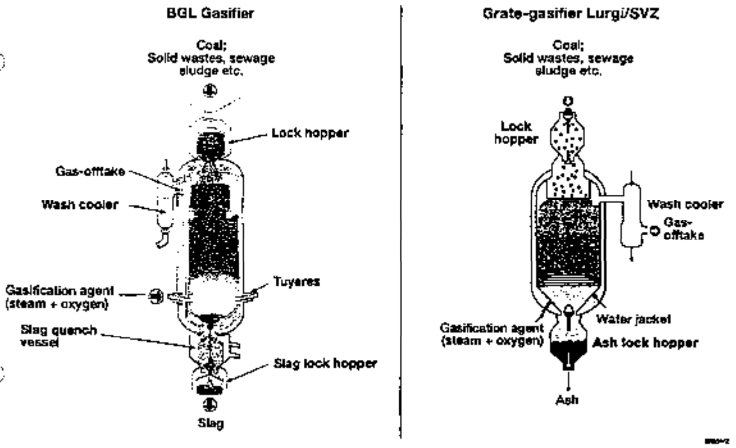
Present Status and Future Expansion (1. Stage)

Fig. 1



Fixed-bed Gasification Technologies

Fig. 2


BGL Gasifier

Grate-gasifier Lurgi/SVZ

Schwarze Pumpe, June 1992-December 1996
Gasification of Solid and Liquid Secondary Raw Materials

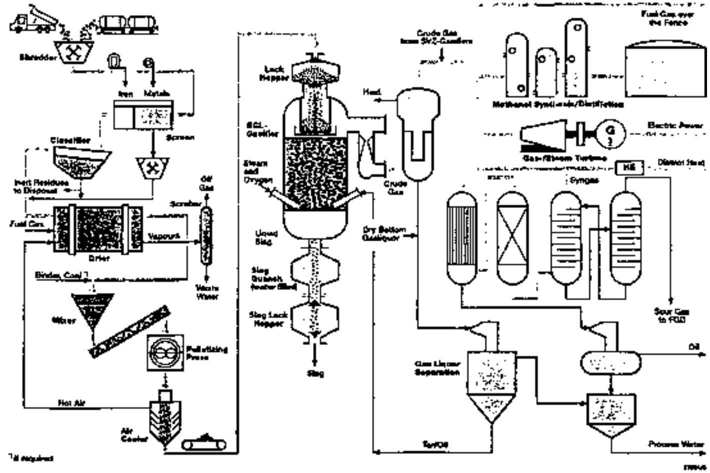
Fig. 3

Feedstocks/mixtures	Quantity gasified mt	Max. Percentage Waste in Feedstock 1) (% wt)
Total waste plastics	148.882	40
thereof agglomerated	141.344	40
Total sewage sludge	35.370	30
thereof dried	16.655	50
Granulated (Swiss Combi)	307	5
Contaminated wood	13.567	50
Oily wastes	2.012	15
Rubber wastes	1.904	30
Dye sludges	186	15
Fluff	3.793	30
Electronic scrap	556	10
Contaminated soil, rubble	17.683	20
Tar residues	4.320	20
Filter residues	487	20
RDF (Refuse Derived Fuel)	2.141	75
Spent activated carbon	68	5
Soot cake	35	5
Other wastes	42	5
Mixtures of wastes		75
Total solid secondary raw materials	**231.026**	
thereof in mixture	ca. 205.700	75
Contaminated oils (PCB, Cl, heavy metals) Oil-/water mixtures Solvent mixtures	**339.735**	100
Total residues / wastes	**570.761**	

1) Permit requires a minimum of 50% lignite in feedstock

SVZ Schwarze Pumpe-
Waste Pelletizing and BGL Gasification

Fig. 4

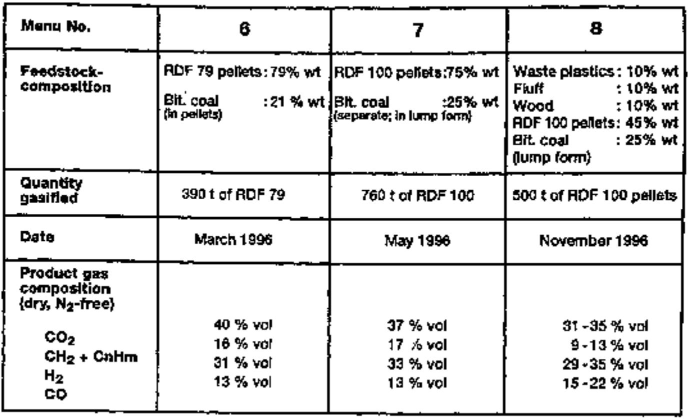


1) if required

Large Scale Gasification Tests at Schwarze Pumpe

Table 1

Gasifier: SVZ-Grate Gasifier Operating Pressure: 24bar



Menu No.	6	7	8
Feedstock-composition	RDF 79 pellets : 79% wt Bit. coal : 21 % wt (in pellets)	RDF 100 pellets : 75% wt Bit. coal : 25% wt (separate; in lump form)	Waste plastics : 10% wt Fluff : 10% wt Wood : 10% wt RDF 100 pellets : 45% wt Bit. coal : 25% wt (lump form)
Quantity gasified	390 t of RDF 79	760 t of RDF 100	500 t of RDF 100 pellets
Date	March 1996	May 1996	November 1996
Product gas composition (dry, N_2-free)			
CO_2	40 % vol	37 % vol	31 - 35 % vol
$CH_2 + C_nH_m$	16 % vol	17 % vol	9 - 13 % vol
H_2	31 % vol	33 % vol	29 - 35 % vol
CO	13 % vol	13 % vol	15 - 22 % vol

9

THE BOTTOMLESS REFINERY:

Gasification Integration as the Optimum Solution to Maximum Refinery Efficiency and Profitability

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9 GASIFICATION TECHNOLOGY IN PRACTICE

FEBRUARY 26-27, 1997

MILAN, ITALY

I. INTRODUCTION

Long-term forecasts for future crude oil supply indicate a gradual decline of crude quality (lower API gravity and higher sulfur). Simultaneously, environmental restrictions are expected to become increasingly stringent in terms of SO_x and NO_x emissions which will impact fuel oil product quality (lower sulfur in fuel oils used for combustion). The foregoing will put increased pressure on most refiners to find outlets for their heavy, sulfur laden residues. Blending these materials into the bunkers market may provide a short term solution for some refiners, particularly those near the coast. However, by the year 2005, a large surplus of high sulfur residue is predicted[1] which can not be absorbed by the expected bunkers market.

One obvious solution is to modify the refinery process configuration and move toward zero high sulfur residue production (i.e., "bottomless refinery"). However, based on current crude and product prices, investments required for installing new conversion facilities are, for the most part, not economically justified. This is increasing the pressure to shut down some refinery capacity. An alternative solution, at a first glance, may not be obvious but has been available for nearly fifty years. Texaco's Gasification technology, which historically has been widely practiced by the chemical industry, is enabling refiners to convert high sulfur residues into higher value products such as hydrogen, power, and steam in a cost-effective and environmentally superior manner. Texaco's Integrated Gasification Combined Cycle ("IGCC") expertise, merges over 50 years of gasification experience with established power generation techniques and, is well positioned to meet these refinery challenges.

II. COMMERCIAL HISTORY

The Texaco Gasification Process was developed in the late 1940s. Early gasification research efforts focused on manufacturing synthesis gas from natural gas to produce liquid hydrocarbons via the Fischer-Tropsch technology. At that time, Texaco was looking to capitalize on its inexpensive and readily available natural gas due to concerns about the lack of long-term availability of liquid hydrocarbons in the U.S. following World War II. Texaco's research efforts resulted in the first commercial scale gasification application in 1950. (Figure 1). Although the synthetic hydrocarbon industry did not materialize, Texaco was able to leverage this successful demonstration of its technology into the manufacture of ammonia for the rapidly growing fertilizer industry. Development of coal and liquids based gasification technologies continued during this period. The first commercial scale use of oil feedstocks occurred in 1956 and early coal work began at about the same time. With the energy crunch of the mid 1970's, Texaco's focus shifted to coal gasification, once again, and culminated with the first commercial coal gasification facility in 1983, at Eastman Chemical's Kingsport, Tennessee Plant.

[1] Oil & Gas Journal - December 9, 1996 - Mr. P. Hunt

This success was followed by the award winning Cool Water IGCC project in California which began operation in 1984. This 1000 TPD coal gasification to power project was the first commercial scale facility to demonstrate the successful integration of gasification technology with state-of-the-art combined cycle technology. A truly remarkable feature was that this first of a kind facility required less than four weeks from mechanical completion to achieve full production. The plant produced up to 125 megawatts (MW) of power from a variety of coals via a combustion turbine operating in a combined cycle mode. The facility met or exceeded all objectives during its limited demonstration life of five years.

Presently, Tampa Electric Company's Polk Power Station in Tampa Florida, utilizes Texaco's Gasification technology to gasify 2000 TPD of coal, producing approximately 280 MW of power. This plant was declared in commercial operation on September 30, 1996 less than three months from initial startup. This plant solidifies Texaco's position as the world's leader in IGCC applications.

Recent project development has been geared to utilizing low valued and waste feeds such as those generated in refineries and chemical plants. The ability of the Texaco Gasification Process to handle a wide range of feedstocks without the need for additional equipment or changes in operating procedures ensures both fuel supply and choice in selecting the least expensive fuel alternative while maintaining a high availability and high reliability.

III.. REFINERY APPLICATIONS

Texaco gasifiers will soon be producing 4.6 million normal cubic meters of syngas per hour in forty-four installations currently operating across the world. An additional eighteen facilities are in various stages of engineering and construction. Today, much of this capacity is used for chemical production, such as hydrogen, ammonia, methanol and oxochemicals. But by the end of this millennium, at least 45% of the syngas generated by Texaco gasifiers, will be used for power generation (Figure 2). In fact, 1996 represented a significant milestone for Texaco gasification technology and for refineries worldwide as three Texaco Gasification IGCC projects successfully achieved financial closure on non-recourse project financing basis. The acceptance by the international financial community of Texaco's Gasification Technology enables refiners, with limited discretionary funds to access investment funds for gasification projects utilizing a limited recourse financial structure. Additionally, Texaco's 40 MW El Dorado, Kansas gasification project was financed via an off-balance sheet operating lease. This is another example of potential funding mechanisms for cash limited refiners seeking to address their bottom of the barrel problems. Table 1 provides an update on these and other current gasification projects.

The API project is now under development by a joint venture between API and Asea Brown Boveri (ABB) called API Energia. This facility will gasify visbreaker residue from the API refinery located in Falconara, Italy. The project will gasify approximately 60 MT/Hr of high sulfur (5-7%) residue in quench gasifiers to produce steam for the API refinery and synthesis gas which will fuel an ABB combined cycle unit to generate approximately 280 MW of power. Construction is scheduled to commence in mid 1997 with startup scheduled for late 1999.

The ISAB project is under development by a joint venture between Erg Petroli and Mission Energy called ISAB Energy. This facility will gasify asphalt from the Erg Petroli refinery located in Priolo Gargallo, Sicily. Approximately 120 MT/Hr of high sulfur (5-7%), asphalt will be gasified in quench gasifiers to produce steam and hydrogen for the ISAB refinery and syngas to fuel Siemens combustion turbines in a combined cycle unit to generate approximately 520 MW of power. Site preparation will begin in 1997 with startup targeted for late in 1999.

The SARAS project is under development by a joint venture between SARAS and Enron called Sarlux. This facility will gasify visbreaker residue from the Saras refinery located in Sardinia Italy. Approximately 145 MT/Hr of high sulfur (5-7%) visbreaker residue will be gasified in quench gasifiers to produce steam and hydrogen for the Saras refinery and synthesis gas to fuel General Electric combustion turbines in a combined cycle unit to generate approximately 560 MW of power. They plan to begin site preparation in late 1997 and startup in the first quarter of 2000.

The API Energia, ISAB Energy, and the Sarlux joint ventures are the first integrated gasification power projects to attain non-recourse project financing. This was no easy task for they represent the first time non-recourse project financing was used on IGCC technology. Moreover, the financings were also first of a kind for Italy which had not previously employed the non-recourse approach to financing. Texaco acknowledges the significant time, expense and personal commitment expended by all the parties involved in these financings and offers sincere congratulations to these pioneering projects for a job well done.

As non-recourse type of financing has traditionally been an engine for independent power production (IPP) project development, this is a significant breakthrough for the gasification to power business. Many of the aspects of the financial arrangements for these projects will undoubtedly serve as a model for future Texaco based IGCC projects.

IV. ECONOMICS

The costs for producing power from refinery residues can be very attractive. Table 2 lists the estimated cost of producing power utilizing a typical Texaco IGCC system based on a heavy oil residue feedstock. These estimates were based on cost information derived from several recent IGCC projects. The capital costs are presented on an instantaneous basis (January 1997) and do not include inflation or interest during construction. The estimated cost of electricity of 4.3c/KW compares favorably with most new power projects excluding natural gas based projects. Moreover, internal rates of return (IRR), on a 75/25 levered basis will yield commercial returns. When reviewing the data in Table 2, we believe that the feedstock and investment costs, the two largest components, require further discussion.

With respect to gasification feedstock cost, Table 3 provides a simplistic derivation for calculating the feedstock costs. A 40$/ton price appears to be a reasonable reflection of expected values, based on current published blend values for this quality feedstock. It should be pointed out that this simplistic model considers only sulfur specifications and does not consider other fuel specifications such as viscosity and particulates which would reduce the blend value of the gasifier feedstock. In addition, most cost projections for gas oil and bunker fuel indicate that high sulfur refinery bottoms are expected to decrease in value. As the IGCC project would enter into a feedstock purchase agreement, the project participants will ultimately decide the appropriateness of increasing the project return (i.e., lowering the feedstock cost) or increasing the refinery return (i.e., increasing the feedstock cost).

The investment cost listed in Table 3 is realistic and is based on current work by several major contractors. The key to achieving the above investment costs include the effective management of project development costs, early decisions on and proper control of the plant design, and effective risk management. Failure to effectively control these activities could easily result in an increase in capital investment as high as $500/KW or a lc/KW increase. Texaco stands ready to assist its customers during project development to maximize the benefits and synergies to the customer. It should also be pointed out that our analysis did not utilize the new class of advanced combustion turbines which have been recently announced by others (i.e., "F", "G", & "H" gas turbines). Utilization of these turbines will further improve the financial performance of refinery IGCC projects. Moreover, Section V lists potential refinery benefits which would further increase the economics of the project and/or profitability of the refinery.

V. REFINERY BENEFITS

In addition to securing an attractive return on its IGCC investment, a refiner will frequently capture a number of additional benefits through the incorporation and integration of the gasification facility into the refinery configuration:

Crude Flexibility

Table 4 lists the feedstocks which have been successfully processed in commercial Texaco Gasification plants during five decades. Since refinery processing schemes have changed over the years and since crude sources can change overnight, many Texaco Gasification units have been designed to handle a wide range of feedstocks without the need for additional equipment or changes in operational procedures. The process feedstock flexibility of the Texaco gasification unit can minimize fuel selectivity and ensure the least expensive fuel for power generation. Therefore, the gasification units can lift the bottoms limitations on a refinery and allow the refiner to have wider flexibility with respect to crude selection and/or process configuration which could be crucial to the long-term economic viability of the refinery.

Minimization of Waste and Disposal Liabilities

Feedstock supplied to a Texaco gasifier need not be limited to just bottom of the barrel materials. The flexibility of a quench gasifier is such that refinery waste materials (i.e., phenolic resins, tank bottoms, etc.) can be co-processed in the gasifier to enhance the investment return by reducing the expense and long term risks associated with on-site disposal or processing by third parties.

Process Flexibility

Texaco Gasification units are used commercially to produce syngas for hydrogen, ammonia, carbon monoxide, methanol, oxo-alcohol, reducing gas and town gas. The key difference among these applications is the hydrogen/carbon monoxide ratio of the clean syngas produced. This ratio can be adjusted by the simple addition of other processing units which can easily be added at a later date to produce these high value syngas derivatives. Because syngas volumes would be obtained with incremental capacity from the existing gasification unit, the economics of any of these applications can be highly attractive.

Environmental Aspects

Texaco Gasification Technology has exceptional environmental performance. Market forces and the worldwide emphasis on reducing emissions are pushing refineries, utilities and chemical companies to employ cleaner fuels and technologies. What is acceptable today may not be acceptable in the 21st century. Therefore, when making technology selections, one must look not only at the current set of environmental standards, but should recognized that future standards will, most likely, be even more restrictive.

Figure 3 shows how a Texaco gasifier incorporated into a refinery can significantly reduce the total refinery emission under a bubble concept. Equipped with a flexible Texaco quench gasifier, the exceptional environmental performance of this particular refinery easily exceeds the new EEC refinery environmental limits.

Minimization of Operating Expense

The addition of a gasification facility which incorporates an air separation unit can enable significant volumes of nitrogen (for purging or blanketing) and oxygen (for increasing the capacity of FCC or Claus sulfur units) to be available at a low incremental cost. Steam, power and other utilities benefit from much higher availability and reliability. This translates into improved refinery operations, lower refinery utility expenses, improved yields and higher on stream performance.

VI. GASIFICATION PROCESS DESCRIPTION

A simplified block flow diagram of a "typical" Texaco IGCC plant, is shown in Figure 4 attached. This plant integrates a heavy oil (i.e. visbreaker tar, asphalt etc.) gasification unit with a combined cycle unit in such a way that optimal overall plant efficiency is achieved, while ensuring excellent plant availability and operational flexibility.

The heart of the IGCC plant is the Texaco gasifier. The gasifier is a refractory-lined pressure vessel in which a carbonaceous feedstock is reacted at high temperature (typically 1300°C) with an oxidant stream. Because the primary reactions are exothermic, the use of a temperature moderator such as water or steam is also required. A typical quench gasifier vessel, capable of processing all hydrogen feeds from natural gas to petroleum coke, is shown in Figure 5.

The product of the reaction is a gaseous stream primarily comprised of hydrogen and carbon monoxide (synthesis gas). Lesser amounts of carbon dioxide, hydrogen sulfide and carbonyl sulfide are also formed. Since this is partial oxidation (i.e. incomplete combustion) and the gasifier is a reducing (not oxidizing) environment, no NOx or SOx are formed.

Ash in the feed oil is non-combustible and remains in a solid form. Due to both thermodynamic reaction limitations and minimal efficiencies of combustion, some small amount of carbon in the oil is not reacted and remain as solid particles combined with the ash. The gasifier pressure is generally chosen to optimize the efficiency of syngas delivery to its end use. Operating pressures in current commercial applications range from 24 to 83 Bar. For IGCC applications, the optimal pressure depends on the pressure requirements of the power generation equipment and the possibility of additional energy recovery by pressure letdown.

The hot synthesis gas and solids (unconverted carbon and ash) leaving the combustion chamber are quenched by contact with water in the lower quench portion of the gasifier.

The majority of the solids in the synthesis gas are washed from the synthesis gas. Any unconverted carbon is recovered in the carbon extraction system and recycled to the gasifier for maximum carbon utilization. The design of the carbon extraction and water treatment sections enables ash to be separated from the recycled carbon, removed in solid form, and processed by others for metal recovery.

The particulate free syngas is then sent to the Low Temperature Gas Cooling Section, which consists of a series of heat exchangers designed to cool the synthesis gas, in stages, while efficiently recovering the heat as medium pressure steam.

The cooled syngas is then sent to the Acid Gas Removal (AGR) Section for removal of sulfur compounds which minimizes the sulfur emission (in the form of SOx) from the combustion turbine. For power applications, the design of the AGR is to selectively remove the sulfur gases (H2S and COS) while leaving the carbon dioxide in the synthesis gas to enhance the gas turbine efficiency - carbon dioxide provides additional mass flow in the combustion gas turbine which translates directly to additional electrical production. There are a wide variety of commercial technologies for selective acid gas removal. Preferred choices are based on economic considerations.

The sulfur off-gas from the Acid Gas Removal Section is sent to a sulfur plant to convert the sulfur off-gas to elemental sulfur, suitable for by-product commercial sale. A preferred technology choice for TGPS applications is the processing sequence of Claus/SCOT Tail Gas Treating, because of its extensive experience in refinery applications.

From the Acid Gas Removal Section, the sulfur-free (clean) syngas is sent to the combustion turbines for generation of electric power. Nitrogen oxides (Nox) emissions may be controlled by diluting the syngas with water or nitrogen upstream of the combustion turbine. Hot exhaust gas from the combustion turbines enter a heat recovery steam generator (HRSG) to produce steam at various pressure levels. The medium pressure steam produced in the Low Temperature Gas Cooling Section is also superheated in the HRSG. The steam is then expanded in the steam turbines to generate additional power.

VII. CONCLUSION

Texaco's Gasification IGCC Plants offer a unique and commercial solution to the challenges currently faced by many refineries. Having been selected in several refinery based IGCC projects, Texaco is well positioned to optimize the application of gasification into any refinery, and apply the expertise and the lessons learned from the projects to enhance future IGCC applications. Texaco is aggressively moving to expand the use of its technology toward refinery based IGCC projects and is looking forward to expanding its traditional licensing role by entering into joint ventures for development and/or ownership of future IGCC projects.

REFERENCES

1. "Comparison of the Economics of Integrated Gasification Combined Cycle Power Generation and Other Clean Coal Power Options," APEC Clean Coal Technology Seminar, Mr. R. Weissman - Texaco.

2. "A European Perspective on the Incentive to Upgrade the Bottom of the Barrel", Mr. P. Hunt - Chem Systems Ltd.

3. "The Gasification Process Associated to Electrical Production: A New Alternative to Refinery Deep Conversion" - Messrs. R. DelBravo, G. Pignateli.

4. "Cogeneration refinery for Total Residue Destruction", Hydrocarbon Processing, November 1993 Messrs. G. Farina, M. Fontana - Foster Wheeler Italiana.

5. 1996 Feasibility Study commissioned by the Ministry for the Environment and the Ministry of Economic Affairs, The Netherlands.

TEXACO GASIFICATION MILESTONES

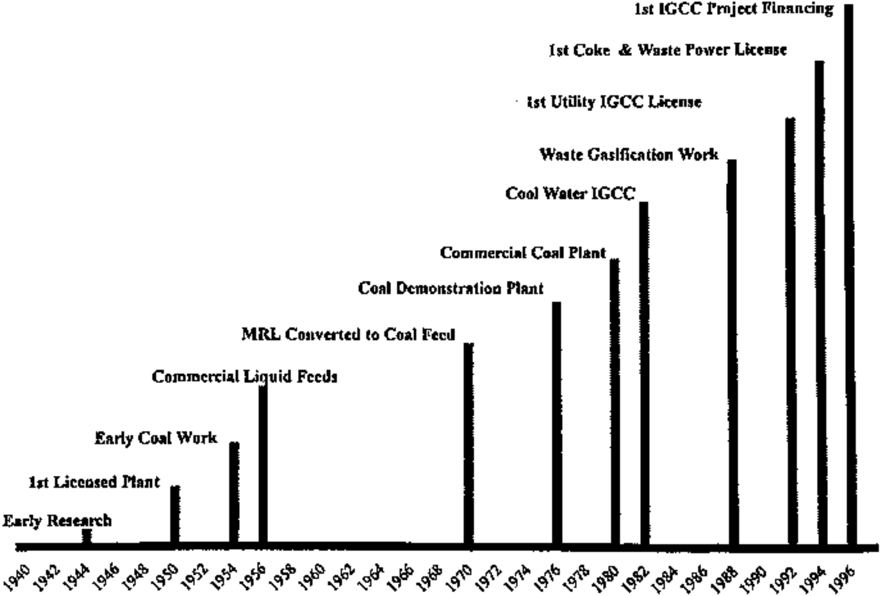


FIG. 1 FIGURE1.PRS

LICENSED GASIFICATION CAPACITY

Growth of Power in 1990's

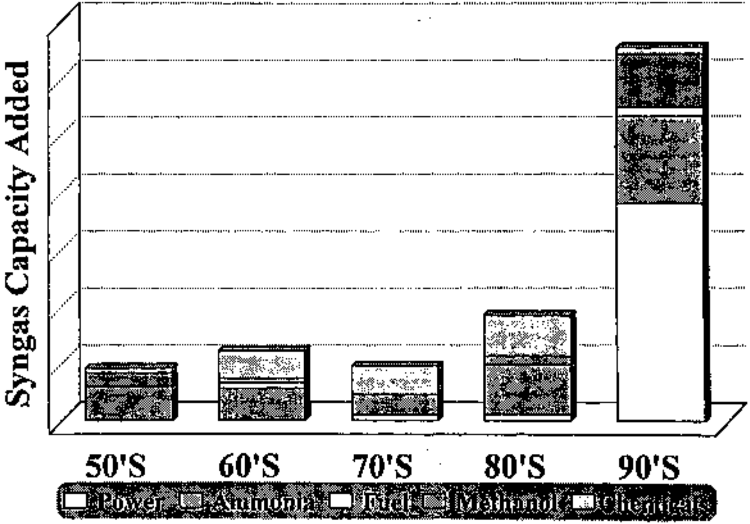


FIG. 2 FIGURE2a.PRS

EXAMPLE REFINERY ENVIRONMENTAL PERFORMANCE

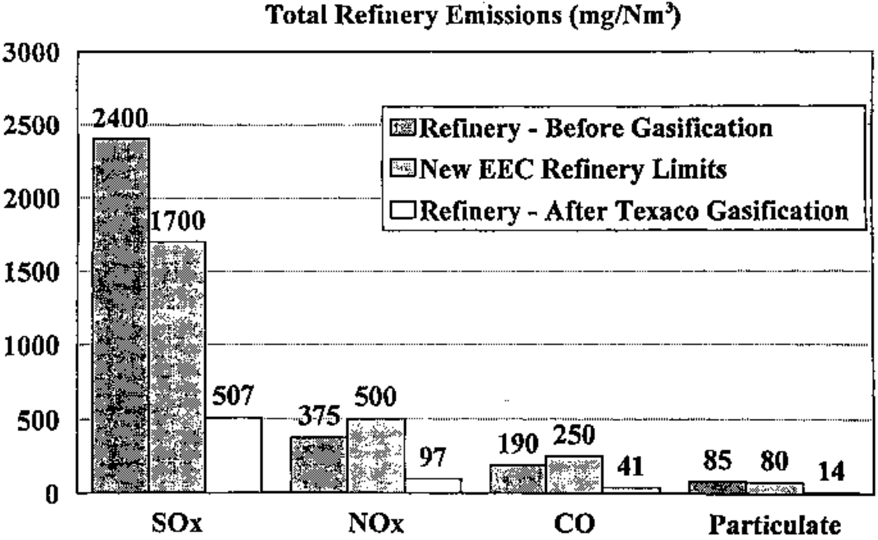


FIG. 3

FIGURE3.FRS

TEXACO IGCC PLANT

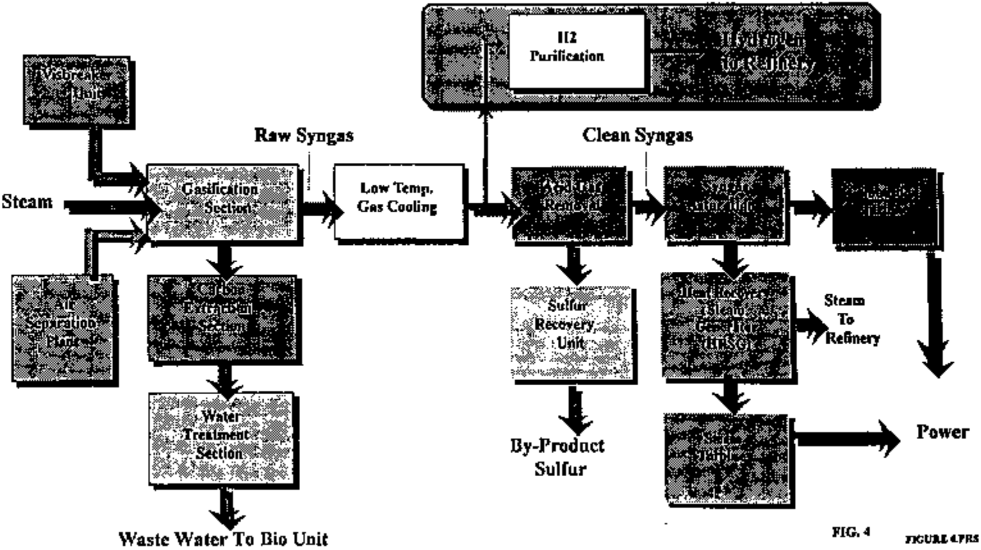
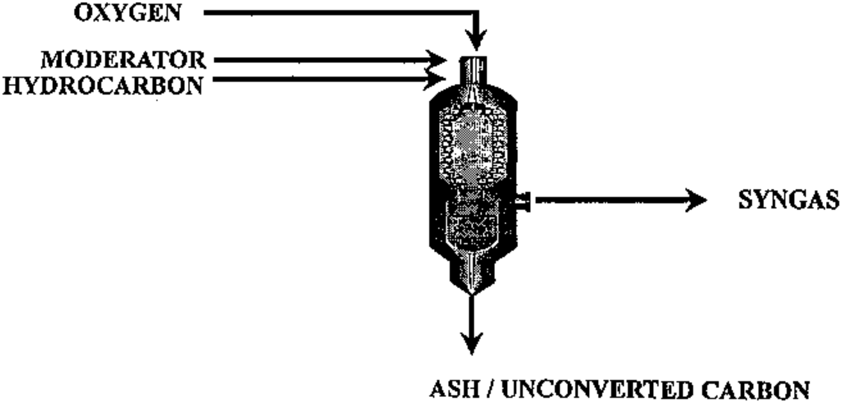


FIG. 4

FIGURE 4.PRS

TEXACO QUENCH GASIFIER



OXYGEN

MODERATOR

HYDROCARBON

SYNGAS

ASH / UNCONVERTED CARBON

FIG. 5

TEXACO GASIFICATION

IGCC PROJECTS FOR THE 1990'S

PROJECT	OWNER	SIZE - MW	LOCATION	FEEDSTOCK	OPERATIONAL
EI Dorado	**Texaco, U.S.A.**	**40**	**El Dorado, KS**	**Waste/Pet. Coke**	**1996**
Polk Power Station	**Tampa Electric**	**260**	**Polk County, FL**	**Coal**	**1996**
API Energia	**API & ABB**	**280**	**Falconara, Italy**	**Visbreaker Residue**	**1999**
ISAB Energy	**Erg Petroli & Mission Energy**	**520**	**Priolo Gargallo, Italy**	**Asphalt**	**1999**
SARLUX	**SARAS & Enron**	**560**	**Sarroch, Italy**	**Visbreaker Residue**	**2000**

TABLE 1

TEXACO FEEDSTOCK DIVERSIFICATION

YEAR	FEEDSTOCK
1950	NATURAL GAS
1956	HEAVY FUEL OIL
1957	WHOLE CRUDE
1961	NAPHTHA
1962	VACUUM RESID
1972	WASTE OILS
1983	ASPHALT
1983	COAL
1984	H-OIL BOTTOMS
1986	PETROLEUM COKE

Shell gasifiers in operation

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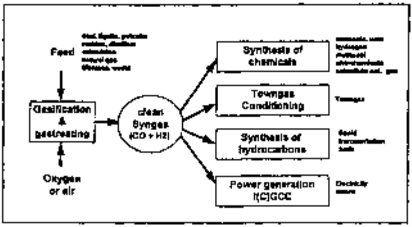
1. INTRODUCTION:

Gasification is a very versatile process to convert a variety of hydrocarbon feed stocks like coal, lignite, oil distillates, residues, and natural gas into synthesis gas ("syngas"). Syngas, essentially a mixture of carbon monoxide and hydrogen, can be converted in subsequent process units into products like ammonia, urea, methanol, oxo-chemicals, town gas and hydrogen.

These applications are found on a world-wide scale. The most application, except towngas, is the utilisation of syngas for its chemical composition. Syngas in integrated gasification combined cycle is a relatively new application and is utilising the clean syngas for its combustion value. Utilisation of low cost feed stock, feed flexibility and superior environmental performance are important drivers to support further introduction of this technology for power generation.



Gasification applications

Shell developed and successfully applies two dedicated gasification technologies, the Shell Gasification Process (SGP) for gaseous and liquid feedstocks and the Shell Coal Gasification Process (SCGP) for coal, lignite and petroleum coke.

Major recent projects based on the SGP-technology are the PER+ project at Shell Pernis refinery and the Shell Middle Distillate Synthesis Process (SMDS). The SCGP technology has recently been applied in the 250 MWe Integrated Coal Gasification Combined Cycle at Demkolec, Buggenum.

Refineries world wide are subject to increasing environmental legislation and gasification of low value heavy refinery residues is one of me options to realise low emissions. The economics of refinery gasification projects may be enhanced by considering co-production schemes. In the Netherlands, construction is nearly completed for the PER+ Shell Refinery project near Rotterdam, scheduled for start up mid 1997. This project, based on the SGP process, is designed to co-produce hydrogen for internal refinery use and steam and electricity in an Integrated Gasification Combined Cycle unit (IGCC), surplus electricity will be exported to the grid.

Syngas derived from gasification of natural gas is typically used for chemical purposes, producing ammonia, urea and methanol. A recent project, taking remote natural gas as feed to SGP units, is the Shell Middle Distillate Synthesis plant in Malaysia which converts the syngas into kerosene, gasoil and wax via Fischer-Tropsch synthesis.

Fuel diversification, longer term supply considerations, increasing costs of conventional clean fuels like low sulphur fuel oil and gas, and competitively priced internationally traded coal has stimulated the development of coal gasification in Integrated Coal Gasification Combined Cycle units (ICGCC). In the Netherlands, a 250 MWe coal gasification combined cycle plant based on Shell technology has been built by Demkolec, a development partnership of the Dutch Electricity Generating Board (N.V. Sep). The construction of the unit was completed end 1993 and after start-up in 1994 the unit is now in its demonstration period, after that the plant will become part of the Dutch electricity generating system.

2. THE SHELL GASIFICATION PROCESS (SGP)

2.1 SGP - FEEDSTOCKS

Historically the SGP process, initially developed in the 1950's, used fuel oil and bunker C-oil as feedstock. By the 1970's vacuum (short) residue had become the standard feed In the eighties vacuum residues were even further concentrated by visbreaking and/or C4/C5 deasphalting.

Over time the feeds became heavier, viscosities and levels of sulphur and heavy metals became higher. Typical properties of present SGP feedstocks are:

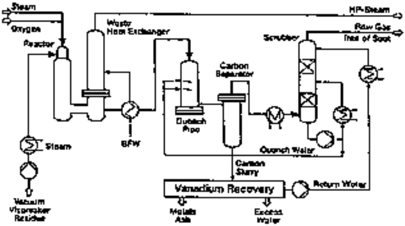
Feedstock type:		Visbreaker Residue	Butane asphalt
Elementary analysis	Units		
C	%wt	85.27	84.37
H	%wt	10.08	9.67
S	%wt	4.00	5.01
N	%wt	0.30	0.52
O	%wt	0.20	0.35
Ash	%wt	0.15	0.08
Total	%wt	100.00	100.00
Vanadium	ppm wt	270 - 700	300
Nickel	ppm wt	120	75
Sodium	ppm wt	30	30
Viscosity (100 oC)	cSt	10,000	60,000
Density (15 oC)	kg/m3	1100	1070

In individual cases considerably higher concentrations than those in the table have been experienced. Other impurities - mostly metals - have also been encountered.

2.2 SGP - DESCRIPTION

Gasification

The main features of a gasification system are the gasification, in which the feedstock is reacted with oxygen to raw synthesis gas (carbon monoxide = CO + hydrogen = H_2), the synthesis gas cooling and the carbon handling system. The gasification is autothermic with as main reaction $2CH_n + O_2 \rightarrow 2CO + nH_2$. Depending on the composition of the feedstock and the oxidant, and the actual gasification temperature (1250 - 1400 °C) the raw syngas contains quantities of H_2O, CO_2, CH_4, H_2S, N_2, and Ar. The gasification pressure is normally between 30 (preferred for IGCC) and 60 bar (preferred for H_2 production).



Simplified Process Flow Diagram of a Residual Oil-Based SGP Unit

The non-catalytic partial oxidation of hydrocarbons by SGP takes place in the gasifier equipped with a specially designed bumer. This design provides for more efficient gas/liquid mixing and **a** better flame temperature control (confined flame).

The oxidant is preheated to minimise oxygen consumption and mixed with steam as moderator prior to feeding to the burner. The burner and reactor are tuned such mat this mixture is intimately mixed with the preheated feedstock within the reactor confinement. The reactor space is optimally used for the gasification to completion. The viscosity range of the feedstock has been widely expanded by replacing the pressure atomising burner by a blast atomising burner. Steam shielding to ensure long burner life and integrated start-up via the main burner are applied. The bumer management system includes a sophisticated safeguarding system as well as a sequence logic block allowing die start-up to be fully automated.

Syngas Cooling

The product of the partial oxidation reaction is a raw synthesis gas at a temperature of about 1300 °C which contains particles of soot and ash. The recovery of the sensible heat in this gas is an integral feature of the SGP process.

Primary heat recovery takes place in a Waste Heat Exchanger (WHE) generating high pressure (e.g. 100 bar) saturated steam in which the reactor effluent is cooled to about 340 °C. The design of the WHE has been developed specifically for these operating conditions and is already used in some 135 installations world-wide. Residues originating from all crudes known to Shell can be gasified in SGP without resulting in serious fouling of the coils. Part of the steam generated in the WHE is used for feedstock and oxidant preheating; the remainder is superheated for use in steam turbine drives. Secondary heat recovery takes place in a boiler feed water economiser immediately downstream of the WHE.

Soot Removal

In the partial oxidation of hydrocarbons the product gas contains a certain amount of free carbon (soot). The plant is normally designed for a soot content in the gas equivalent to about l%wt of the hydrocarbon reactor feedstock. The soot particles are removed from the gas together with the ash in a two stage water wash. This consists of a quench pipe and a soot separator followed by a packed tower, the soot scrubber. In the quench pipe about 95 % of the soot is removed by a direct water spray. In the scrubber me gas is washed in counter current flow in two packed beds. A circulation system is employed over the lower bed using a circulating pump. The upper bed is washed with return water from the soot recovery section. After leaving the scrubber at a temperature of about 40 °C the gas has a residual soot content of less than 1 mg/m3 and is suitable for feeding to the desulphurization unit.

The soot formed in the partial oxidation reaction is removed from the system with the process condensate as a soot slurry and is routed to the soot recovery unit. After soot removal the main part of this water is recycled as return water to the top of the scrubber, excess of water is routed to the waste water treatment section.

Soot Recovery Unit

The traditional approach to handling the soot slurry is to contact it with a hydrocarbon, thus to pelletise the soot, separate the soot from the water and recycle it back to the reactor. The viability of this approach deteriorated with heavier more viscous gasification feedstocks containing amongst others a higher content of metals/ash. This has led to the development of an alternative once-through approach: the Soot Ash Removal Unit, SARU, incorporated in the SGP in the Per+ project.

Waste Water Treatment

The surplus of water ex SARU (i.e. the overall net water produced in the gasification step) is routed to a Sour Water Stripper to remove traces of ammonia, NH_3, hydrogen cyanide, HCN and hydrogen sulphide. The offgas is normally routed to a Claus/Scott unit.

After this pre-treatment the water still contains about 15 mg/1 HCN, 1 mg/1 H_2S and 20 mg/1 NH_3. The final water clean-up takes place in a flocculation-sedimentation system for trace metal (ash) removal and a biological waste water treatment unit. The quality of the treated water fulfils die most stringent (German) standards: Ni < 0.5 mg/1, V < 2 mg/1, BOD5 < 25 mg/1.

2 J SGP HISTORICAL HIGHLIGHTS, THE PRESENT AND THE FUTURE

* 1950: In 1950 Shell envisaged a surplus of fuel oil and initiated R&D on gasification.

* 1950-1973: Between 1950 and 1973 the feedstock was mainly bunker fuel oil. Scale-up took place from type 200 to type 1000 gasifier (type 1000 corresponds to 1000 kNm3 syngas/d or about 350 t/d residue). The pressure increased from 20 to 65 bara.

* 1973-1985: Even heavier feedstocks were processed like vacuum flashed cracked residue during this period. The soot produced is captured by the naphtha soot recovery unit and recycled completely to the reactor. A higher reliability of the waste heat exchanger was realised by applying a double tube sheet and natural circulation. The longer residence time in the reactor, introduced in this period, resulted in a lower soot production, lower oxygen consumption, lower CO_2 content in the syngas and a longer life of die refractory. The first type 1200 reactor is applied.

* 1985-1996: The co-annular burner was introduced for the type 1500 gasifier making it possible to process even heavier feedstocks like asphalts. As alternative for the naphtha soot recovery unit the soot ash removal unit has been developed which introduces the once-through approach for feedstocks with high contents of metals/ash.

Over the years the focus of research has been on the following aspects.

- Burner development resulted in integrated heat-up and an increased burner life (more men 8000 hrs)
- Scale-up from type 200 to type 1500 while a design has been made for type 3000 gasifier.
- The reactor pressure has been increased from 20 to 65 bara.

- Reliability of the waste heat exchanger has significantly been improved by applying a flat front and improved understanding of fouling, erosion and corrosion mechanisms.
- The work-up of soot and ash has improved with the development of the soot ash removal unit.

* 1997: Presently 82 SGP reactors arc producing about 62 million Nm3 syngas/day in 26 plants world-wide. This is equivalent to 23 ktons residue/day or 7.7 million ton residue/year. The availability of a single SGP string has increased to 98 % over the last ten years. Unplanned shut down is typically 2 days/year.

Vacuum flashed cracked residue and asphalts originating from almost all crudes known to Shell can be gasified in SGP with a waste heat exchanger. Due to the development of both burner technology and the soot ash removal unit very heavy feedstocks can be processed.

At present the main feedstock is vacuum flashed cracked residue and the main outlet is still ammonia. However, this may change. Refineries world-wide are being subjected to increasing pressures, both legislative and economic. Both in the United States of America and in Western Europe stringent environmental legislation affects both the refinery operation and the product quality. It is becoming more and more unacceptable to burn inland Heavy Fuel Oil (HFO) without expensive flue gas treatment to meet environmental standards.

In the past refinery schemes were developed to optimise the lighter part of the product slate - gasoil, kerosene and gasoline. The residue was "discarded" into the fuel oil pool, often back-blended with lighter components to meet the HFO quality specifications. The increased demand for distillates (i.e. transportation fuels) and the reduced demand for HFO results in an unbalance between demand and supply. This is forcing refineries to look for alternative conversion of the heavy residues.

Technologies for residue processing as deasphalting, carbon rejection (visbreaking, thermal and catalytic cracking, Flexi-Coking) and hydrogen addition (catalytic hydrotreating, hydro-desulphurization and hydrocracking).Some of the processes from the first group produce very heavy residual products. Schemes for further working-up of these residues include delayed coking or residue/asphalt gasification. Traditional residue and coke usage is becoming more difficult and one of the feasible alternatives is gasification producing high value clean products from "dirty" low value feedstocks.

The main advantages of integrating gasification in a refinery are:

- the capability to process high sulphur crudes because of the almost complete removal of sulphur compounds in the syngas treating unit;
- the capability of processing low quality, very viscous and heavy feedstocks;
- the high growth in demand for gasoil. Fuel oil is converted into mainly high quality gasoil (high cetane) in a hydrocracker. The required hydrogen can be produced via gasification;
- the many outlets for the synthesis gas e.g.: hydrogen for hydrocracking and hydrotreating, electricity and steam production in IGCC, chemical applications e.g. ammonia, methanol, acetic acid, oxo-alcohols etc. and syndietic fuels via e.g. the Shell Middle Distillate Synthesis

According to the International Energy Agency (IEA) gasification of residues could represent the ultimate synergy between power production and refining and the cleanest disposal route for poor quality fuels.

3. THE SHELL COAL GASIFICATION PROCESS (SCGP)

3.1 SCGP - EXPERIENCE

The application of gas-fired combustion turbine-combined cycle systems has grown rapidly in the industry due to the lower cost, higher efficiency and demonstrated reliability of gas turbine equipment. There are already gas fired combined cycle power plants with overall efficiencies around 55% LHV. Developments in gas turbines and steam cycles point in the direction of 55 to 60% LHV efficiency for such plants in the near future.

Modern coal gasification technologies present a unique opportunity to combine the advantages of high efficiency combined cycle power generation with an environmentally friendly coal based process. The Shell Coal Gasification Process (SCGP) is especially well-suited to produce clean turbine fuel gas efficiently and it can be coupled with a combined cycle system in either an integrated or a non-integrated arrangement. This allows considerable flexibility in configuring the gasification plant.

Recent studies show that optimal integration of the three main building blocks of an ICGCC plant (air separation unit, SCGP and the combined cycle) could give an overall efficiency (LHV) of 46-48% based on commercially available gasturbines.

In such cases the air separation unit receives part of its air from a dedicated air compressor, with the remainder coming from die air compressor of the gas turbine. The nitrogen not required in the SCGP is used for dilution of the clean syngas fired in the gas turbine to reduce NO_x emissions and increase electrical output.

Shell's operational experience with coal gasification started with a 6 t/d pilot plant in Amsterdam, followed by a 150 t/d unit in Harburg, Germany and a third unit in Houston with a capacity of 250-400 t/d. This experience has been the basis for the successful design of the 2000 t/d coal gasification unit of the Demkolec plant.

The basic concepts selected for the Shell Coal Gasification Process are:

- Pressurised: compact equipment;
- Entrained flow: compact gasifier;
- Oxygen blown: compact equipment, high gasification efficiency;
- Membrane wall, slagging gasifier: robustness, high temperature, insulation by slag layer;
- Opposed burners: good mixing, high conversion, scale-up possibility;
- Dry feed of pulverised coal: high gasification efficiency, feed flexibility.

The process can handle a wide variety of coals, ranging from bituminous to lignite, in an environmentally acceptable way and produces a high purity, medium-Btu gas as a fuel for power generation, as a chemical feedstock or as a source of hydrogen.

3.2 SCGP - DESCRIPTION

Raw coal is crushed and fed to a conventional bowl mill, similar to those used in a pulverised coal boiler. This mill grinds the coal to a size range suitable for efficient gasification (90% wt less man 100 microns). As the coal is being ground, it is simultaneously dried utilising a heated inert gas stream that carries the evaporated water from the system as it sweeps the pulverised coal through an internal classifier to collection in a baghouse.

The oxygen required in the SCGP gasification step (95% pure for IGCC) is supplied by an air separation plant. Other oxygen purities can be used, depending on project premises. Nitrogen from the air separation unit is compressed to provide low-pressure and high-pressure nitrogen for use in the gasification plant, e.g. for transporting coal in the feed system.

Milled and dried coal from the coal milling and drying area is pneumatically transported to the coal pressurisation and feeding system. Pressurised coal, oxygen and, if necessary, steam enter the gasifier through pairs of opposed burners.

The gasifier operates in the range 20 to 40 bar. The gasifier consists of a pressure vessel with a gasification chamber inside. The inner gasifier wall temperature is controlled by circulating water through the membrane wall to generate saturated steam. The membrane wall encloses the gasification zone from which two outlets are provided.

One opening at the bottom of the gasifier is used for the removal of slag. The other outlet allows hot raw gas and fly slag to exit from the top of the gasifier.

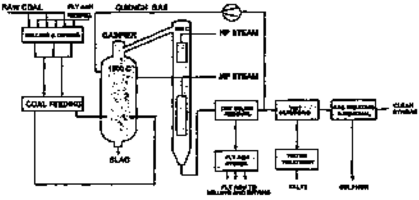
Most of the mineral content of the feed coal leaves the gasification zone in the form of molten slag. The high gasifier temperature (over 1500°C) ensures that the molten slag flows freely down the membrane wall into a water-filled compartment at the bottom of the gasifier. High carbon conversions (above 99%) are obtained, and the high temperature ensures that no organic components heavier than methane are in the raw syngas. The insulation provided by the slag layer in the gasifier minimises heat losses, such that cold gas efficiencies are high and CO_2 levels in the syngas are low. The recycle of fly slag enhances gasification efficiency.

Flux may be added to the coal feed to promote the appropriate slag flow from the gasifier at the preferred operating temperature. As the molten slag contacts the water bath, the slag solidifies into dense, glassy granules. The slag is washed, depressurised and then fed to intermediate storage.

The hot raw product gas leaving the gasification zone is quenched with cooled, recycled product gas to convert any entrained molten slag to a hardened solid material prior to entering the syngas cooler. The syngas cooler recovers high-level heat from the quenched raw gas by generating high-pressure steam.

The bulk of the fly slag contained in the raw gas leaving the syngas cooler is removed from the gas using commercially demonstrated equipment such as filters or cyclones. The recovered fly slag can be recycled back to the gasifier via the coal feeding system. The syngas then goes to a scrubbing system, where the remaining traces of solids and water soluble contaminants are removed, and thereafter to an acid gas removal system, where an amine-based solvent, such as Sulfinol, removes typically 99% of the sulphur species.

A bleed from the scrubbing system is sent to a sour slurry stripper. The water is then clarified and can be recycled to minimise the volume of effluent to be bio-treated and discharged or evaporated. The acid gas from the acid gas removal system and from the sour slurry stripper is fed to a Claus plant, where saleable elemental sulphur is produced. For maximum sulphur recovery and minima! sulphur emissions, the Shell Claus Offgas Treating process (SCOT) is used.



THE SHELL COAL GASIFICATION PROCESS
GASIFICATION & GASTREATING

Of the energy in the coal being fed to the gasifier, approximately 80-83% is converted into raw synthesis gas. Over 99% of the carbon in the coal is converted. With a further 14% to 15% of the heating value of the coal being recovered as steam, the overall thermal efficiency of the Shell Coal Gasification Process is very high. The composition of the syngas is similar for most coals processed, with CO and H_2 accounting for over 90% of the syngas on a molar basis.

Sixteen different coals, varying from bituminous to lignite, as well as petroleum coke were successfully run at SCGP-1 in Houston during some 15000 operating hours.

Coals can be switched during operation, with gasifier operating conditions being adjusted as the new coal "breaks through". The use of automatic process control makes it possible to maintain optimal operating conditions even with variable coal quality.

To accommodate changing power demand from the grid, load-following was thoroughly tested at SCGP-1, including steady-state operation at 50% plant capacity. In response to demand changes, the control system is able to accommodate ramp changes in syngas demand at a rate of 3% per minute between 100% and 50% plant capacity.

3.3 SCGP - ENVIRONMENTAL

Integrated gasification combined cycle power generation based on SCGP has an excellent environmental performance. A priority throughout the development and commercialisation of me process has been the establishment of an environmental data base and also a number of design options for treating and handling effluents.

Gaseous Effluents

A key advantage of using coal gasification for power generation is that the gasification step converts the sulphur in the coal into hydrogen sulphide and some carbonyl sulphide, both of which can be removed down to extremely low levels. The resulting sulphur rich acid gas is normally sent to a Claus unit for conversion into sulphur, a readily saleable product

The nitrogen in me coal is converted to molecular nitrogen, except for small amounts of ammonia and hydrogen cyanide which are completely removed in the syngas clean-up sections.

Particulate removal is done in a dry solids removal section comprising cyclones and filters to produce a syngas that meets particulate specifications of typically 1-5 ppmwt. This is further reduced in subsequent scrubbing and treating steps.

Aqueous Effluent

Process water from SCGP has no detectable amounts of volatile or semi-volatile organics. Biological treatment of the stripped and clarified process water provides oxidation for the small amounts of inorganic nitrogen and sulphur species that remain. Biotreated effluent contains fully oxidised products and very low concentrations of trace metals. Stripped sour water and treated scrubber water can be recycled.

Toxicity tests have been carried out on samples of biotreated effluent. The results showed no acute toxicity in the undiluted SCGP effluent and no chronic toxicity effects at a modest 3 to 1 dilution of the effluent. A "zero water discharge" configuration is also an option.

Solid By-Products

Depending on the coal and the gasifier operating mode, more than 90% of the ash in the coal ultimately leaves the process as bottom slag. Both bottom slag and fly slag are non-hazardous according to RCRA (Resource Conservation and Recovery Act, USA) requirements.

As part of a solids utilisation program, SCGP slag has been used as a major component in concrete mixtures to make roads, pads and storage bins. Other applications of SCGP slag and fly slag that have been demonstrated are asphalt aggregate, Portland cement kiln feed and light weight aggregate.

4. RECENT MAJOR PROJECTS BASED ON SHELL GASIFIERS

4.1 SGP AND THE PER+ PROJECT

Background

Shell's Rotterdam refinery is processing oil since 1936, the intake capacity is some 20 million tonnes of crude per annum. As from 1949 a petrochemical complex has been added, currently supplying 3 million tonnes of chemical products per annum.

The refinery is presently being upgraded to meet product volume and quality demands for lighter and environmentally cleaner transportation fuels well into the next century, whilst reducing emissions to the environment.

The main driving forces for the realisation of this project are the following. The ever tightening restrictions on the emissions of sulphurdioxide, nitrogenoxydes and particulates are reducing the scope for firing residual fuel in furnaces and boilers. Increasing product quality specifications, especially on sulphur content, require the application of more sophisticated production techniques. The decreasing potential for exporting heavy fuel oils, in combination with the desire to process heavy crudes, creates the necessity to apply more and deeper conversion processes.

Extensive studies have been carried out to determine the optimum concept fulfilling above needs in the most cost effective manner. As a result of these studies the PER+ project did materialise, which compromises amongst others:

- a single string hydrocracker unit with a design capacity of 8000 mt/sd waxy distillate intake;
- a three string residue gasification unit, which will process 1650 mt/sd vacuum flashed cracked residue, according to the Shell Gasification Process (SGP);
- associated gas treating facilities to remove contaminants and to produce 285 mt/sd pure hydrogen for the hydrocracker and some 1600 mt/sd clean synthesis gas to serve as gasturbine fuel;
- a co-generation power plant with an installed capacity of 130 MWe and 400 mt/hr process steam, comprising two gasturbine generators with supplementary fired heat recovery steam generators and two steam turbine generators;
- a water demineralisation and condensate treatment plant with a total capacity of 1600 mt/hr;
- over 3000 tie-ins to the existing refinery installations.

The total project will come on stream in 1997 and represents an investment of over 3 billion Dutch guilders, including improvements to existing facilities.

Process selection

Crude supply, product quality, and demand trends played an important role in the selection of the new processing facilities. The new hydrocracking unit (HCU) will replace the oldest of the two fluid catalytic cracking (FCC) units.

The HCU was selected in preference to a major upgrade of an existing unit or a new catalytic cracker because of its better middle distillate selectivity and superior product quality (very low sulphur, high cetane number). By this choice, investments in cat-feed hydro-treaters or FCC product desulphurizers are obviated.

Hydrogen required for the HCU, approximately 250 t/sd, will be produced from syngas originating from gasification of heavy, vacuum-flashed, visbroken residue.

Steam methane reforming (SMR) was considered as an alternative to residue gasification. Unlike gasification, SMR would not have reduced fuel oil make. This in turn would have led to reduced crude supply flexibility, both in sulphur and API gravity, and thus to **a** more expensive crude diet. In addition, expensive natural gas would have been required as SMR feed.

Hydrocracking Unit

The new HCU will be **a** high-conversion, once-through, single-reactor, series flow hydrocracker. The unit will be fed with heavy straight-run VGO (370 - 550 °C), flashed distillates originating from residue hydroconversion (Hycon, Shell's proprietary process) and visbreaking unit and lube oil extracts.

About 80% of the feed will be converted to high-quality 370°C minus products (mainly tops, naphtha, kerosine, and gas oil). The other 20% is low sulphur, low-cokeforming tendency feed for the remaining FCC unit.

The very large reactor vessel weighs around 1200 t. The reactor contains two different catalysts in multiple, stacked-bed configuration. A single reactor design was selected to extract maximum benefits from the economy of scale.

Gasification Unit

The three-train gasifier will have a capacity of 1650 t/sd (3 type 1500 gasifiers) and will be fed with heavy, vacuum-flashed, visbroken residue. Approximately 1600 t/sd of oxygen will come from future third-party sources nearby. The main reason three gasification trains were selected is that, in case of outage of one gasifier, hydrogen requirement for the HCU can be delivered from two gasifier trains.

Under normal three-train operation, the syngas in excess of the requirements for hydrogen production will be used as gas turbine fuel.

The combined process is called the SGHP, Shell gasification hydrogen process. The hydrogen plant consists of a two-stage CO shift (high temperature/low temperature), carbon dioxide removal, and methanation. Lurgi's Rectisol process was selected for H2S removal from the syngas, as well as for C02 removal downstream of the low-temperature CO shift. Both Rectisol treating steps are highly integrated.

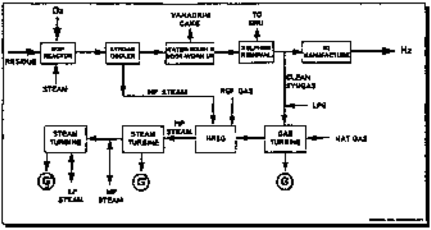
The soot and ash-containing scrubber water from the three trains is filtered and returned to the scrubber; excess water is exported to water treatment facilities. The filter cake will be worked up for metal recovery.

New Combined Cycle Unit

The basis of design and many key equipment choices for the combined cycle choices were governed by the refinery's existing utility infrastructure. Two new gas turbines, General Electric MS 654IB (ISO rating 43 MWe) form the heart of a new 130 MWe cogeneration plant. Saturated steam from the waste-heat exchangers of the SGHP will be superheated in the fired waste-heat boilers of the gas turbines. Part of the high pressure steam will be used as process steam in the gasifiers and the hydrogen plant; the remainder will be sent to steam turbines for generation of lower-pressure steam and electricity. An important issue in the project design has been the gasturbine fuelgas system to meet the complex requirements of flexibility and reliability for the Pernis refinery. The gasturbines therefore have to be able to burn different gases and gas mixtures and meet NOx emission specifications under all conditions. Syngas is the prime gasturbine fuel, natural gas is the start-up and back-up fuel. LPG surplus can be routed to the gasturbine till a maximum of 50 %.

On the steam end the combined cycle is designed to act as a balancing consumer or producer for the MP- and LP-refinery steam system. The low pressure refinery gas system is used for continuous supplementary firing of the waste heat boilers of the gasturbines, required to condition the high pressure steam output.

With the existing powergenerating facilities Pemis remains a net exporter of electricity to the public grid and a large share of the electricity generated in the new combined cycle unit will be exported.



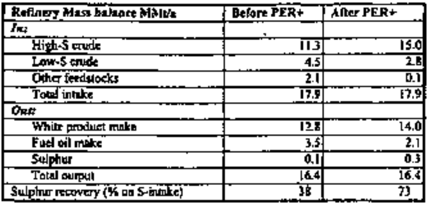
PER+ Scheme, Gasification, gas treating, hydrogen manufacturing and combined cycle

Construction

A major challenge of this project, besides its scale, has been to construct the units in a fully operational refinery. About 3000 tie-ins with, or modifications to existing process, utility, and movements facilities will have to be made. All new facilities will be operated from a single control room and maximum use will be made of advanced process control. The project is co-ordinated through an engineering, procurement, and construction contractor (Fluor) supported by specialised engineering contractors (Fluor, Lummus and Comprimo).

Environmental

The positive environmental effects of the PER+ investments are summarised in the following table:



Refinery Mass balance MMt/a	Before PER+	After PER+
In:		
High-S crude	11.3	15.0
Low-S crude	4.5	2.8
Other feedstocks	2.1	0.1
Total intake	17.9	17.9
Out:		
White product make	12.8	14.0
Fuel oil make	3.5	2.1
Sulphur	0.1	0.3
Total output	16.4	16.4
Sulphur recovery (% on S-intake)	38	73

4.2 SGP AND SMDS

For several decades Shell's research has been involved in synthesis gas chemistry, with special focus on routes to convert natural gas into easily transportable liquid hydrocarbons. This effort has resulted in the development of the Shell Middle Distillate Synthesis (SMDS) process, for which the first commercial plant is in operation in Sarawak, Malaysia.

The SMDS technology is essential a three stage process.

In the first stage of the SMDS process, synthesis gas is obtained by partially oxidising some 2200 t/d natural gas with pure oxygen in SGP gasifiers. The oxygen is produced by an air separation unit, a 2500 t/d oxygen extraction plant.

In the second stage of the process, the Heavy Paraffin Synthesis, the synthesis gas is passed through synthesis reactors, where the Fischer-Tropsch reaction takes place. The reaction takes place over a very active and selective Shell proprietary catalyst. The reactor conditions have been chosen so that the formation of long chained liquid paraffinic molecules (wax) is favoured whilst that of gaseous compounds such as butanes and even lighter components is minimised. This approach maximises the desired products as the liquid paraffinic molecules can be converted almost entirely into middle distillates in the third stage. This second stage employs four reactors, and is me heart of the SMDS process.

In the third and final stage, the wax molecules are converted into middle distillates in the Heavy Paraffm Conversion unit by mild hydrocracking. The middle distillate stream is then fractionated to produce kerosene and gasoil, and some naphtha. The total output of the SMDS complex is some 12500 barrels/day.

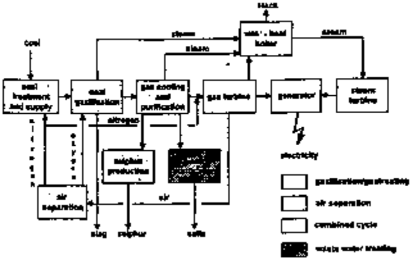
The SGP unit has six parallel reactor trains. Each train consists of a type 1200 gasifier.

Problems experienced in the project were related to metal dusting. These problems are presently being addressed. Although metal dusting has negatively influenced the on-stream factor per reactor it has not influenced the overall on-stream time of the unit. The concept of 6 parallel train has proven its added value while, in addition, the robustness of the gasifiers made it possible to operate each reactor significantly above design value.

4.3 SCGP AND DEMKOLEC

The fuel diversification policy of the Dutch Electricity Generating Board, N.V. Sep, aims at using a balanced portfolio of fuel, including coal, for power generation. At present some 40 % of Dutch electricity is produced in modern, conventional fired pulverised coal boilers. The fuel diversification policy together with stringent environmental requirements were main drivers to select coal gasification for the so called Demkolec ICGCC unit, officially named "Willem Alexander Centrale". Demkolec B.V., a subsidiary of N.V. Sep, is responsible for construction and operational testing during the demonstration period. Thereafter the plant will be used as a commercial production unit.

The main elements of the 253 MWe Demkolec unit are shown in the blockscheme below. The plant is highly integrated aiming at a high efficiency, amongst others through extraction of the total feed to the air separation unit from the gasturbine air compressor. Furthermore the steam systems of the gasification, gas cooling and gastrearing sections are fully integrated with the steam systems of the combined cycle unit and the auxiliary boiler. The gasturbine is connected to the natural gas grid for start up but also to satisfy the obligation for a two fuel supply system.



DEMKOLEC BLOCK DIAGRAM

The following table summarises some performance parameters for Demkolec at full load:

Coal input	585 MW
Coal gas	460 MW
Heating value gas in GT	4300 kJ/kg
Gas Turbine output	156 MW
Steam Turbine output	128 MW
Own use	31 MW
Net power output	253 MW
Efficiency	43 %

In 1988, after a pre-selection of gasification processes, a comparative study was made in which three coal gasification technologies were evaluates in depth. In April 1989 the Shell process was selected and the basic design was started. Construction started in the second part of 1990 followed by commissioning which was completed as scheduled end 1993. After start up early 1994 the plant is in its demonstration phase to assess the various aspects of ICGCC at commercial scale like availability, reliability, operability, environmental performance, load following and economic aspects.

During the demonstration phase it became clear that the gas-turbine had high frequency vibrations (humming) when operating on syngas. Initial solutions to the humming were achieved early 1996 while the final solution was implemented in the third quarter of 1996 after a dedicated test program.
In the mean time the ash and slag handling problems experienced initially in the gasification section were addressed. Foregoing activities significantly improved plant performance resulting in a plant availability of the complete ICGCC system of 66 % over the last couple of months in '96. The total amount of on-stream time of the gasifier added up to some 5000 hrs.

The following list summarises the operational experience until end 1996:
- 12 types of coals successfully gasified (including 6 blends);
- Overall efficiency: according to design (Carbon conversion higher than 99 %);
- Different systems tested between 40 and 100 % load;
- Longest continuous operation of the gasifier some 600 hours;
- Environmental performance is equal or better than design and within the permit levels, i.e. minimum circa 98% desulphurization (S02), overall maximum 75 gr/GJ NOx, zero discharge of all water streams, noise emission below 54 dB.

The GT13E2 Medium Btu Gas Turbine

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Gasification Technology in Practice
26-27 February 1997, Milan, Italy

Abstract

ABB's GT13E2 gas turbine is utilised in a refinery gasification process where the engine is modified to operate on Medium Btu gas (GT13E2/MBtu). The standard GT13E2 is now a well proven machine with a performance of 165 MW and a simple cycle efficiency of 35.7%, reaching 55.5% (gross) in combined cycle performance. ABB advanced combustion systems routinely operate at NO_x values of 25 ppm (15% O_2) with a turbine inlet temperature (TIT_{Iso}) of 1100 °C.

The GT13E2 now has over 100,000 hours fired operation with 28 units in the field. Significant experience has been gained by ABB with the operation of these GT's and the GT13E2/MBtu design has benefited from the implementation of measures which will ensure that the GT13E2/MBtu is reliable when it is entered into service.

The modification of the standard GT13E2 to operate with Medium Btu (MBtu) has resulted in a unit where the main performance parameters of the GT13E2/MBtu are improved to exceed 185 MW and 37% (simple cycle) at ISO conditions with a turbine inlet temperature of 1080 °C (TIT_{IS0}).

The compressor module of the unit is upgraded to incorporate an extra end compressor stage to boost the pressure ratio to 17:1 and improve performance. The unit has an inherently high surge margin and can be used in the gasification process without the requirement for air extraction. Therefore, no delivery of compressor discharge air to an air separation unit (ASU) is foreseen. Instead the ASU is supplied with a separate compressor, designed for the optimum gasification conditions. This power plant concept gives a high flexibility for start up, shut down and load shedding operation.

In the present paper the features of the GT13E2 will be explained and the conversion of the gas turbine to MBtu syngas firing for gasification projects will be described in detail, highlighting the engineering and research work carried out to enable the GT13E2/MBtu to operate reliably within gasification process.

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Introduction

Low emissions and high efficiency are the major requirements for the current power generation market. With the 165 MW-class turbine GT13E2 both needs can be satisfied for a wide variety of power plant applications. Due to the significant increase in gas turbine and combined cycle efficiency and the very low emissions of gas turbines burning clean fuels (i.e. no fuel bound nitrogen or sulphur components) complex gasification combined cycle processes are becoming economically feasible. The hydrocarbons used for the gasification can be residual oils, coal or industrial waste. Oxygen-blown residual oil gasification is a well established technology for hydrogen production in refineries.

To achieve low emissions with these fuels, modern high temperature gas turbines designed for natural gas and oil No. 2 have to be modified, especially with respect to the combustion technique used. The ABB combustion technique for fuels from gasification processes is premix combustion with moderate nitrogen dilution for NO_x control. With this technique, no air extraction is required from ABB's 13E2 gas turbine due to a sufficient surge margin. Major modifications of the 13E2 gas turbine are restricted to the premix EV burners and the fuel distribution system. Compared to diffusion burner techniques, much smaller volume flows have to be controlled for the operation of the gas turbine and no water is used for NO_x control.

The GT13E2 gas turbine

The GT13E2 follows in direct line from the successful aerodynamic and mechanical design features of the GT13 product family, with a single shaft concept with two bearing sections, a welded monolithic rotor, a subsonic compressor, a highly efficient turbine, efficient cooling systems for turbine rotor, vane carrier and front stages and one combustion chamber. With more than 130 machines sold from the GT13 product family these design features have demonstrated their reliability in more than 2.8 million operating hours and are, therefore, the key for power plants with high reliability and availability (Viereck , 1992). The GT13E2 is the latest member of the GT13E product family with a turbine inlet temperature of 1100°C (acc. to ISO definition) and a pressure ratio of 15:1, which gives a power output of 164.3 MW and an efficiency of 35.7% in single cycle application (ISO at base load conditions with gas fuel).

Since launching the GT13E2 , 28 are operating in the field, with over 100,000 fired operating hours. The GT13E2 has a proven reliability record with a significant amount of effort going into testing the units at ABB test site in Sodegura, Japan and also with extensive field trials at several customer sites.

Fig. 1 shows the turbo-generator group of the GT13E2 and GT13E2/MBtu. As can be seen, the GT13E2 inlet casing surrounds the journal and the thrust bearing on the compressor side of the rotor. Inspection work on both bearings can be carried out without removing the inlet casing. The compressor of the GT13E2 is nearly the identical 21 stage subsonic design of the direct predecessor GT13E. It features a single stage of variable inlet guide vanes for high combined cycle part load efficiency (Viereck, 1992b).

The compressor diffuser, which was designed using inverse fluid dynamics, guides the main air flow from the last stage of the compressor to the plenum of the gas turbine casing, where a small amount of the air is used for cooling the first two vane rows, while the main air enters the combustor via the EV burners.

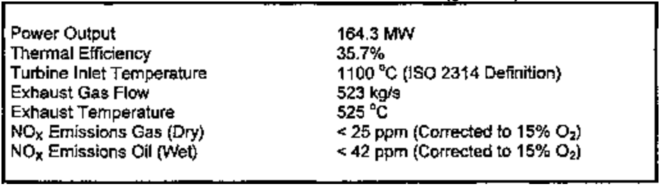
The combustion system of the GT13E2 is a single annular combustor design (no cans) with 72 lean premix EV burners, arranged symmetrically in four rings around the turbine. This symmetrical arrangement gives a homogeneous mixture of hot gas and, therefore, an excellent temperature pattern factor in front of the first stage vane (see Senior et al, 1993). The annular combustor was designed with respect to an optimised combustion with low NO_x emissions. Using the EV lean premix burners, NO_x levels of less than 25 ppm (15% O_2) for gas fuel without water or steam injection and less than 42 ppm (15% O_2) for fuel oil (wet control) are guaranteed.

The turbine of the GT13E2 is a highly efficient 5 stage design with identical blade and vane profiles as the GT13E. Inspection of each turbine stage is possible either in situ (1 stage via annular combustor, 5th stage via exhaust diffuser) or via horoscope technique without dismantling the gas turbine.

The rotor of the GT13E2 is an ABB traditional monolithic welded rotor design which is maintenance free. The rotor has three balancing planes which enables re-balancing without opening the casing.

Table 1 summarises the performance data of the GT13E2 at ISO conditions in base load operation.

Table 1: Performance data of the GT13E2 (gas fuel)



Power Output	164.3 MW
Thermal Efficiency	35.7%
Turbine Inlet Temperature	1100 °C (ISO 2314 Definition)
Exhaust Gas Flow	523 kg/s
Exhaust Temperature	525 °C
NO_x Emissions Gas (Dry)	< 25 ppm (Corrected to 15% O_2)
NO_x Emissions Oil (Wet)	< 42 ppm (Corrected to 15% O_2)

Specification of syngas fuel properties

Oxygen-blown gasification delivers syngases with a heating value of the order of 10 to 16 MJ/kg compared to a heating value of less than 5 MJ/kg for air blown gasification. The higher heating value facilitates the syngas desulphurisation and HCN removal processes and also allows the use of ABB gas turbines designed for natural gas without exceeding the compressor surge margin. Different feedstocks for the gasifier lead to different syngas compositions and heating values.

Residual oil gasification syngas is the most challenging syngas composition as far as premix burner technology is concerned. -Residual oil gasification fuel from oxygen-blown processes can be represented by a typical volumetric composition of 45% H_2 , 48% CO and 7% N_2 and a lower heating value of 15 MJ/kg. The H_2/CO ratio is about 1. The first application of a GT13E2 to hydrogen-based syngas fuel (api Energia, Falconara, Italy) will be for residual oil gasification syngas with a H_2/CO ratio of 0.96 and a lower heating value of 13.9 MJ/kg.

Coal gasification syngas has a typical volumetric composition of 30% H_2, 60% CO and 10% N_2 and a lower heating value of approx. 12 MJ/kg. The H_2/CO ratio is only 0.5. Since the hydrogen content is much lower than residual oil gasification syngas, the flame velocity and the peak flame temperature is considerable lower.

Reliable premixed combustion systems for gas turbines have only been developed for natural gas to date. The main reason for this situation are the unique properties of natural gas under lean premixed conditions. Long ignition delay times and high self ignition temperatures make it possible to control pre-ignition and the aerodynamic and thermoacoustic behaviour of such combustors with well designed burners. Application to other, more reactive fuels is much more critical, since combustion chemistry introduces additional constraints. The problem of ignition of oil no. 2 during the evaporation phase is well known. To highlight the difficulty of applying premixed combustion for MBtu fuels, their basic combustion properties are discussed in detail below.

The laminar flame speed, adiabatic flame temperature and chemical reaction time of syngases and of natural gas are plotted in figure 2. Properties of residual oil syngas with a 55%(Vol.) dilution of N_2 are given by the dotted lines in the left column in figure 2. All values have been calculated with a one dimensional laminar flame code (Kee et al. 1992) for a pressure of 14.5 bar and a preheat temperature of 300°C. The chemical kinetics data base was that of Miller and Bowmann (1989) with 52 species and 251 elementary reaction equations.

The peak laminar flame velocity of the syngases is about an order of magnitude higher than the laminar flame velocity of methane. Since intense turbulence will increase the flame speeds significantly above the respective laminar values shown in figure 2 (Liu et al. 1989), it is obvious that the effective flame speed reaches the order of the flow speed in the burner. As a consequence, it is difficult to prevent flashback into the premixing section. Wall boundary layers, wakes or local zones of low velocity (e.g. downstream of fuel jets mixing with air) are particularly critical. The highest flame velocity of the syngas/air mixture occurs at fuel rich conditions (fuel equivalence ratio approx. 2). Hence during the mixing of a fuel jet with the air, fuel rich zones in the jet mixing layer can act as flame holders and prevent the fuel from fully mixing with the air prior to ignition.

The maximum flame temperature for both oil and coal syngas is 2600 K, which is about 200 K higher than that of methane. This is not a problem if full premixing is achieved, since the (mixed) flame temperature can be selected via the overall equivalence ratio. However, if flame stabilisation occurs in regions where the mixing is not yet perfect, the stoichiometry in the flame front will vary. The fuel which burns richer than average will produce very high peak temperatures with NO_x formation rates far higher than in the natural gas case.

The time scale of the chemical reaction in a premixed laminar flame is plotted in the lower row of figure 2. This time scale is the integrated time of a fluid element passing the flame's reaction zone. The reaction times for both syngases are one fifth of that of natural gas. The time scale of the premixing process must be faster than the chemical time scale. The constraints resulting from Damkohler number considerations are very similar to those mentioned above: Flashback can only be avoided if no low speed flow regions exist within the mixing zone.

The effect of diluting syngases with an inert gas (nitrogen in this case) is seen in the left column of figure 2 (dotted lines). With only 55% (Vol.) N_2 dilution, the flame speed can be approximately halved, while the maximum flame temperature drops to values comparable to natural gas. This effect can be exploited to delay ignition until further downstream. In practice, the effect of moderate dilution is very strong, since rich combusting zones, which are most critical for flame stabilisation and NO_x production, are no longer present if the shift of the ignition point downstream is sufficiently large.

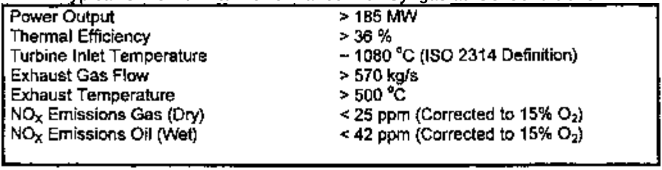
In comparing residual oil and coal syngases, the flame speed is considerably lower for the latter. It is expected from the calculations that premixing can be achieved for coal derived gas with much less fuel dilution. In contrast, the maximum flame temperature is only slightly lower for coal gasification syngas.

Conversion of the GT13E2 to syngas operation

Performance Data

The operation with syngas without any air extraction after the compressor leads to an increased compressor pressure ratio and an increased mass flow rate through the turbine. Increased power and efficiency are the consequence (Scherer, 1994). To minimise the load of the turbine, the turbine inlet temperature has been reduced to 1080 °C (ISO) compared to the standard natural gas fired GT13E2 (1100°C). Typical performance data for a syngas with a heating value of 7.5 MJ/kg are:

Typical GT13E2/MBtu Performance with Syngas at ISO Conditions



Power Output	> 185 MW
Thermal Efficiency	> 36 %
Turbine Inlet Temperature	~ 1080 °C (ISO 2314 Definition)
Exhaust Gas Flow	> 570 kg/s
Exhaust Temperature	> 500 °C
NO_X Emissions Gas (Dry)	< 25 ppm (Corrected to 15% O_2)
NO_X Emissions Oil (Wet)	< 42 ppm (Corrected to 15% O_2)

The basic design of the GT13E2 compressor with its high surge margin allows the operation with the increased pressure ratio without air extraction. Therefore, no delivery of compressor discharge air to an air separation unit is foreseen. Instead the ASU is supplied with a separate compressor, designed for the optimal gasification conditions (non-integrated GCC). Figure 3 shows this basic power plant concept, which gives a high flexibility for start up, shut down and load shedding operation.

Field operation of the standard GT13E2 machines have shown that the basic machine is extremely reliable and delivers the expected performance. In order to incorporate many of the lessons learned in the field the compressor diffuser for the GT13E2/MBtu has been improved in terms of operational reliability and serviceability by the design of a new unit as shown on Figure 10, which indicates the mechanical integrity of the new diffuser. Additionally the compressor has a 22nd stage installed which raises the pressure ratio to 17:1, from the standard machines 15:1. The 22nd stage is a repeating stage derived from the existing 21st stage but with a shorter span.

As already mentioned the main changes in the gas turbine configuration is restricted to the combustion system. Therefore, the main interest of this paper is focused on the modifications of the burner. A slightly modified version of the ABB EV burner fulfils the needs of environmentally friendly syngas combustion.

Adaptation of the burner system

Operation principle of the EV burner

ABB's EV burner is also known as the Double Cone Burner because it consists of two half cones shifted perpendicular to their centrelines thus forming two inlet slots of constant slot width (Figure 4). Air entering through these slots is mixed with gaseous fuel emerging from a large number of holes along each of the slots. Since the slot width is constant and the diameter of the burner cross section increases from the cone tip (upstream end of the burner) to the end of the cone (downstream end of the burner) the swirl number (defined by the ratio of circumferential to axial momentum flux) of the air entering the burner increases continuously. It is well known that if the swirl number exceeds a certain threshold, vortex breakdown occurs on the axis of the swirling flow. With a suitably selected ratio of slot width to burner length this central recirculation zone is formed on the centreline at the end of the burner and serves as an aerodynamic flame holder. Due to the central recirculation zone, stable combustion is possible, even at conditions close to extinction with flame temperatures well below 1500°C, without the need for piloting flames (see Sattelmayer et al. (1990) and Aigner et al. (1990)). This guarantees a minimum of NOx emissions. The burner is inherently safe against flashback since the fuel is injected and mixed in the inlet slots where the flow velocities are high and no fuel is present upstream of the burner.

A large amount of operating experience exists for ABB's Double Cone Burner operating with natural gas. NO_x emissions considerably lower than 15 ppm (15% O_2) have been measured for ABB's gas turbines. ABB's GT8, GT9, GT10, GT11 and GT13 type machines using the same burner in multi-burner assemblies either in silo combustors or in annular combustors. Details on the operation experience with silo EV combustors are given by Aigner and Muller (1992), annular combustor operation experience is described by Strand (1993).

Modification of the EV burner for MBtu fuels

Due to the high flame velocities of hydrogen-containing fuels and the higher volume flux, injection of the fuel along the inlet slots does not lead to a reliable system. If injected along the slots, the large volume and momentum flux of the fuel distorts the incoming air flow profile and at very high hydrogen fuel content, small flames stabilise at the gas injection holes. To overcome these difficulties a simple and effective injection design for the MBtu fuel has been developed. Instead of the injection along the slots, a number of plain holes close to the burner end injecting the fuel radially inward enable inherently safe operation of the burner, even with high hydrogen content fuels. Due to the high velocity of the gas injection, the flame stabilises downstream of the burner (figure 5). The air leaving the burner is entrained into the fuel jets thus premixing the fuel and air. The standard EV burner has been shortened to increase the air velocity at the burner exit where the syngas injector holes are located. This acts to enhance mixing thereby lowering NOx emissions. This system uses the standard EV burner with only minor modifications and is still operable with natural gas or oil No. 2. In the case of natural gas as backup fuel, additional fuel channels to supply the MBtu injection holes at the end of the burner are provided.

In contrast to the standard EV burner design the fuel distribution channels are now located near to the hot end of the burner. To prevent build up of thermal stresses caused by temperature gradients between the cold fuel channels and the burner shells, the two have been mechanically and thermally isolated in the current design.

Burner Experimental results

Two full scale test rigs were used for a systematic testing program of the EV burners, one operating at ambient pressure, the other at gas turbine pressure. Burner tests have been carried out with a fuel composition of the oil gasification syngas (45% H_2, 48% CO and 7% N_2). Oil gasification syngas is more difficult to burn in a premix flame (higher flame velocities) and will also give higher NO_x emissions (higher maximum flame temperature) compared to coal gasification syngas. NO_x emission values measured with this syngas composition in the full scale high pressure experiments can therefore be regarded as a conservative estimate for coal gasification fuel in the gas turbine combustor.

Ambient pressure tests

A full range of fluid supply services are metered, automatically monitored and logged by a microcomputer. These include full temperature, non-vitiated combustion air up to 3000 kg/h and two separate combustor cooling air supplies. CO, H_2 and N_2 are available from pressurised bottles and mixed directly in front of the burner. The single burner test rig is constructed to permit the use of EV burners identical in scale to those found in ABB's gas turbines.

Screening of a large number of different injection geometries was done in the atmospheric tests. Data values marked with triangles in Figure 6 are the NO_x emissions at atmospheric conditions for the injection geometry for which lowest emissions at high pressure were obtained, plotted as a function of the nitrogen dilution. Starting from approximately 20 ppm (15% O_2) for the undiluted syngas, emissions decrease to less than 2 ppm if the syngas is diluted with nitrogen to achieve a lower heating value of 7.5 MJ/kg. Measured CO emissions are less than 8 ppm over the whole range of heating values. No burner overheat or flashback was observed. The air preheat and the combustor outlet temperature was set to the design values of the GT13E2 and kept constant during the tests.

Tests under elevated pressure

Figure 7 shows the high pressure test rig. An axial compressor, a two-stage radial compressor with intercooler and a high pressure air preheater provide up to 5.5 kg/s non-vitiated combustion air at 16 bar and 500°C. The test rig consists of a plenum chamber upstream of the burner, a segmented water cooled tubular pressure and the rectangular chamber liner. The hot exhaust gases are quenched upstream of the back pressure valve and discharged to the chimney.

The liner is cooled convectively in order to avoid film cooling which might cause undesired secondary effects on the emissions measurements. The plenum upstream of the burner is equipped with several windows enabling observation of the mixing zone through the burner slots. Four water cooled suction probes are mounted at different axial stations. Pressure fluctuations as well as static pressures are monitored at different positions in the flame tube. Temperature is measured at 40 different locations in the burner and the liner walls. CO and H_2 are supplied in pressurised bottles, liquid nitrogen is evaporated from a supply tank and subsequently mixed with the CO and H2.

Data values marked with open circles in Figure 6 show the results of the high pressure tests at 16 bar. Again the air preheat and the combustor outlet temperature were set in order to match the GT13E2 operation conditions. It is seen that the NO_x emissions are in the order of 350 ppm (15% O_2) for undiluted syngas and decrease to values between 20 and 35 ppm (15% O_2) when the syngas is diluted to 7.5 MJ/kg. At a heating value comparable to coal gasification syngas (12 MJ/kg), the NO_x emissions are 120 ppm (15% O_2) but it should be noted that oil gasification syngas diluted to 12 MJ/kg has a higher hydrogen content than coal gasification syngas.

During the tests burner temperatures were always at or below design values even if undiluted syngas was used. This ensures that safe operation is possible if the nitrogen dilution supply fails or if large heating value fluctuations occur during load changes of the gas turbine or the gasifier/ASU plant. The MBtu EV burner is of the same inherent safe design as the standard EV burner for natural gas operation. Burner noise (pressure pulsations) was observed to be even lower than for natural gas operation. Due to the very compact reaction zone and the rapid mixing, complete CO burnout was achieved. The strong decrease in NO_x emissions with lower heating values is caused by a more complete premixing of fuel prior to the initiation of combustion. At lower heating values, the flame stabilises further downstream of the fuel injection nozzles, whereas at high heating values only partial premixing is possible before reaction is ignited.

The NO_x emissions scale with pressure to the power of approx. 1.0 which is considerably higher than values obtained for natural gas with the EV burner. It should be noted that other injection geometries gave lower emissions at atmospheric conditions but had a pressure scaling with exponents of more than 1.2, i.e. emissions at high pressure were 40 times higher compared to ambient pressure results. It must be concluded that low emissions at ambient pressure tests are a necessary but not sufficient condition for low emissions at high pressure. It was also found that geometrically similar burners of smaller size have considerably lower emissions compared to full scale burners. The effect of burner size is much stronger than in the case of natural gas. In summary, it was found that only full-scale full-pressure burner experiments with the exact fuel composition give reliable data concerning safe burner operation and emission values.

The high pressure test results have shown, that NO_x emissions of approx. 30 ppm (15% O_2) can be achieved with a modification of the EV burner using a moderate amount of nitrogen dilution. The combustion technique is safe against flashback and tested under full engine conditions.

Part Load

The part load performance was investigated in ambient and high pressure tests with the MBtu EV Burner. Figure 8 shows the NO_x emissions for burner thermal loads from 70 to 100% at both 1 atm and 15 bar. As expected, NO_x emissions decrease with lower load (higher air equivalence ratio). In this load range, CO emissions were well below 8 ppmvd (at 15% O_2) even at atmospheric conditions, which usually produce CO emissions substantially higher thrn high pressure tests. Furthermore, if the compressor mass flow can be reduced by approx. 30% with a variable inlet guide vane system, these very low emissions can be sustained down to less than 50% gas turbine power output without the need for fuel staging. The high hydrogen content in the moderately diluted fuel and the rapid mixing in the near field of the burner are very advantageous in this context.

Burner Durability and Reliability

The GT13E2/MBtu burner system has been analysed with full three-dimensional finite element techniques in order to check its operational reliability. The finite element predictions indicate (Figure 11) that the stress levels will be well below the materials inherent strength and yield the required lifetime in terms of creep and low cycle fatigue. In addition the burner has been instrumented with thermocouples during full load testing and the results indicate that burner material temperatures will be below 850°C which corroborate the finite element stress analysis computations.

Advantages of Nitrogen Dilution

It has been shown that with ABB's lean premix burner technology it is possible to achieve low emissions and high total cycle efficiency with moderate nitrogen dilution for NOx control. Compared to water saturation or steam injection nitrogen dilution offers the following advantages:

- **no** water costs
- better cycle efficiency
- higher power output
- no requirement of high fuel temperature to prevent water condensation within the GT fuel delivery system
- lower dew point of exhaust gases allowing the use of a more efficient steam cycle with lower stack temperatures
- greater potential for even lower NOx emissions without the problems of higher CO emissions and incomplete burnout normally found in steam-injection diffusion burner technology.

In the residual oil application at api Energia the syngas will be diluted to approx. 37 volume % with Nitrogen to obtain a lower heating value of 7 MJ/kg.

The **GT13E2** within the **GCC** process

Fig. 3 shows a simplified flow chart for a non-integrated gasification process. Non-integrated means that an independent ASU with a separate compressor is applied to produce the O_2 for the gasifier. The N_2 from the ASU is used for moderate syngas dilution (7.5 MJ/kg) to assure NO_x emissions below 25 ppm (15% $_{O2}$).

Figure 8 shows the flow diagrams for the combined cycle part of the power plant. A triple pressure combined cycle is depicted in Figure 8 as an example. The lay-out is typical for residual oil gasification as well as for coal gasification. The plant is designed for a net. power output of approx. 270 MW and a net. efficiency of > 44%. Dobbeling et al. /8/ demonstrated that the power output and the efficiency of a non integrated GCC based on EV burner premix combustion with moderate N2 dilution is superior compared to a fully integrated GCC.

Conclusions

High pressure full-scale single burner experiments have shown that premix combustion of hydrogen containing syngases with moderate dilution is possible. By changing the fuel injection system of ABB's EV burner, the 13E2 gas turbine can be operated with MBtu syngas. Equipped with this premix burner technique, NO_x emissions lower than 25 ppm (15% O2) are achieved. The high surge margin of the GT3E2 compressor allows the operation without compressor air extraction. The ASU layout can be made independently from the gas turbine leading to highly efficient and reliable GCC power plants.

Acknowledgements

The authors would like to thank ABB for permission to publish this paper. Special thanks are also due to Roland Muecke, Wulf Raether and others who formed part of the team at ABB to progress the research, development and manufacture of the GT13E2/MBtu.

References

M. Aigner, A. Mayer, P. Schiessel, W. Strittmatter, 1990, "Second Generation Low-Emission Combustors for ABB Gas Turbines: Tests Under Full-Engine Conditions", **Paper No.90-GT-308,** *Gas Turbine and Aeroengine Congress and Exposition,* Brussels, Belgium.

M. Aigner, G. Muller, 1992, "Second Generation Low-Emission Combustors for ABB Gas Turbines: Field Measurements with GT11N-EV", **Paper No. 92-GT-322,** *Gas Turbine and Aeroengine Congress and Exposition*, Cologne, Germany.

K. Dobbeling, H. P. Knopfel, T. Sattelmayer, P. Muller, K. Reyser, 1993, "ABB's Medium Btu EV Burner for Syngas Applications - Test Results and GCC Integration Concepts", EPRI Conference on Clean Coal Combustion Technology, San Francisco.

K. Dobbeling, H. P. Knopfel, W. Polifke, D. Winkler, C. Steinbach, and T. Sattelmayer, 1994, "Low Nox Premixed Combustion of MBTU Fuels Using the ABB Double Cone Burner (EV Burner)," **Paper 94-GT-394,** *Int. Gas Turbine and Aeroengine Exposition,* The Hague, Netherlands.

R. J. Kee, J. F. Grcar, M. D. Smooke, and J. A. Miller, 1992, "A Fortran Program for Modelling Steady, Laminar, One-Dimensional Premixed Flames," *Sandia Report,* **Sand 85-8240,** UC-401.

Y. Liu, B. Lenze and W. Leuckel, 1989, "Investigation on the Laminar and Turbulent Burning Velocity of Premixed Lean and Rich Flames of CH_4-H_2 Mixtures," Paper presented at the 12th International Coll. on the Dynamics of Explosions and Reactive Systems (ICDERS), Ann Arbor, MI, USA.

J. A. Miller and C. T. Bowman, 1989, *Prog. Energy and Comb. Sci.,* 15, 287.

Th. Sattelmayer, M.P. Felchlin, J. Haumann, J. Hellat, D. Styner, 1990, "Second Generation Low-Emission Combustors for ABB Gas Turbines: Burner Development and Tests at Atmospheric Pressure", **Paper No. 90-GT-162,** *Gas Turbine and Aeroengine Congress and Exposition,* Brussels, Belgium.

V. Scherer, K. Doebelling, K. Reyser, D. Viereck, 1994, " The ABB type GT13E2 Gas Turbine and its conversion to Mbtu Syngas Firing for Gasification Projects", 1 st

International conference on combined cycle power generation, Calcutta, India, January 6-8, 1994.

P. Senior, E. Lutum, W. Polifke, T. Sattelmayer, 1993," Combustion Technology of the ABB GT13E2 Annular Combustor", Paper No. G22, *20th International Congress on Combustion Engines (CIMAC),* London.

T. Strand, 1993, "Retrofitting a Low Emission Combustor to a Twin Shaft Gas Turbine", Paper No. G13, *20th International Congress on Combustion* Engines (CIMAC), London.

D. Viereck, U. Weicht, H. Seelinger, H. Wettstein, 1992a, "Combined Cycle Plants Based on GT13E2 Gas Turbines", *9th CEPSI Conference,* Honkong, pp. 264-273

D. Viersck, H. Wettstein, M. Aigner, H.J. Kiesow, 1992b, "GT13E2, the Cleanest Gas Turbine for Combined Cycle Cogeneration Application", IGTI-Vol. 7, *ASME Cogen-Turbo,* Houston, pp. 231-237

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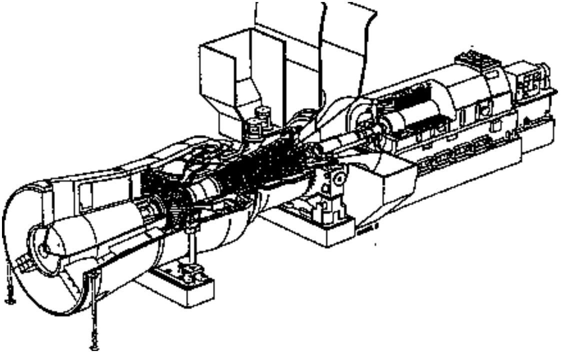

Figure 1. ABB GT13E2 gas turbine cross section.

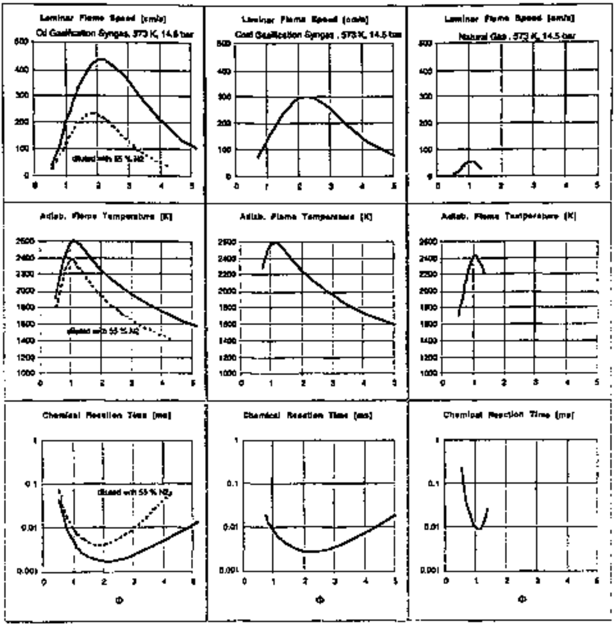

Figure 2. Flame velocity, flame temperature and chemical reaction time for different syngas compositions and for natural gas.

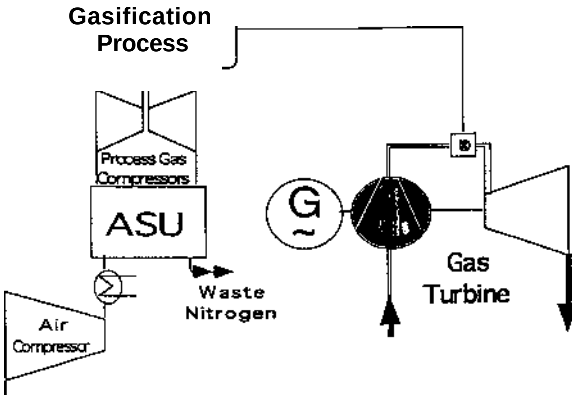

Figure 3. Simplified flow chart of non-integrated GCC.

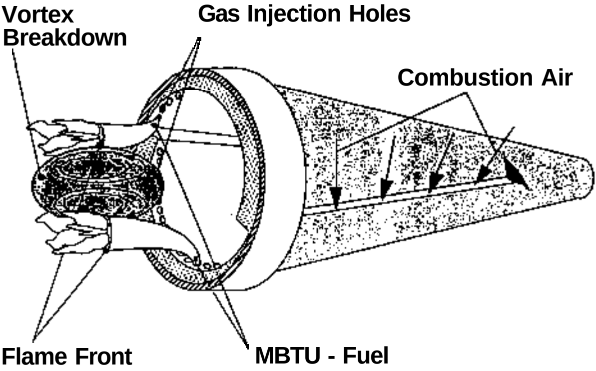

Figure 5. Mixing pattern in the Mbtu EV burner.

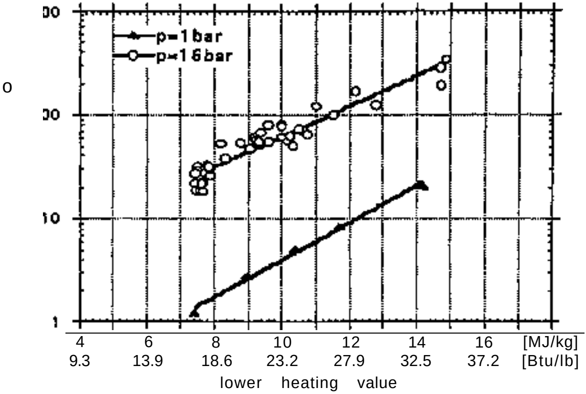

Figure 6. NOx emissions as a function of the lower heating value for MBtu fuels.

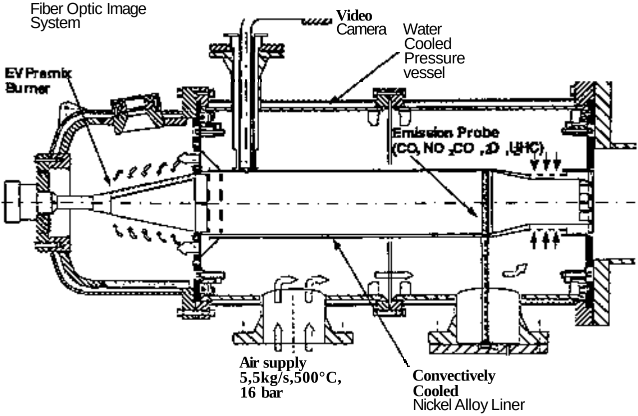

Figure 7. High pressure burner test facility.

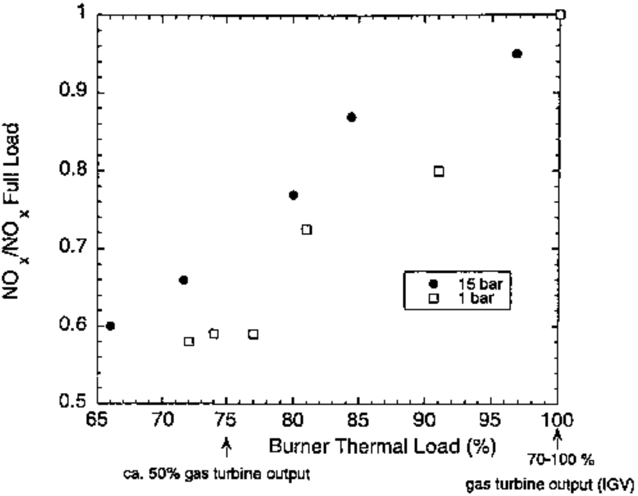

Figure 8. Part load NOx emissions performance of the MBtu EV burner.

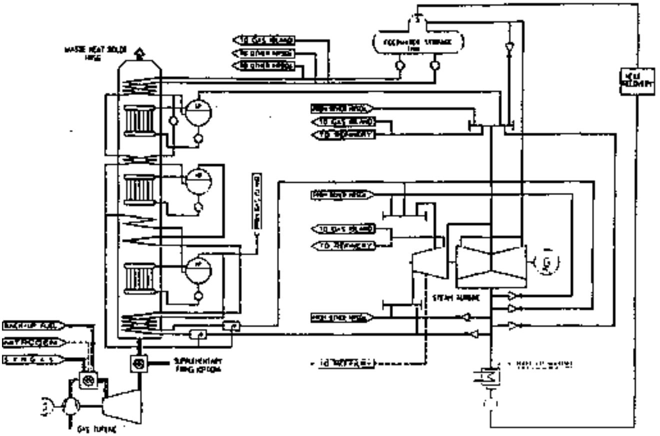

Figure 9. Principle diagram of combined cycle.

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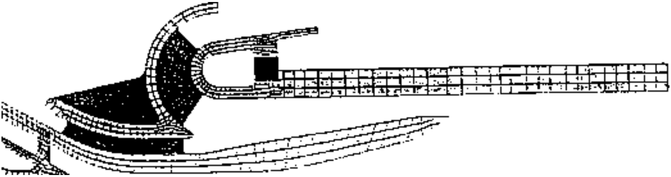

Figure 10. 2-D finite element analysis mesh for diffuser.

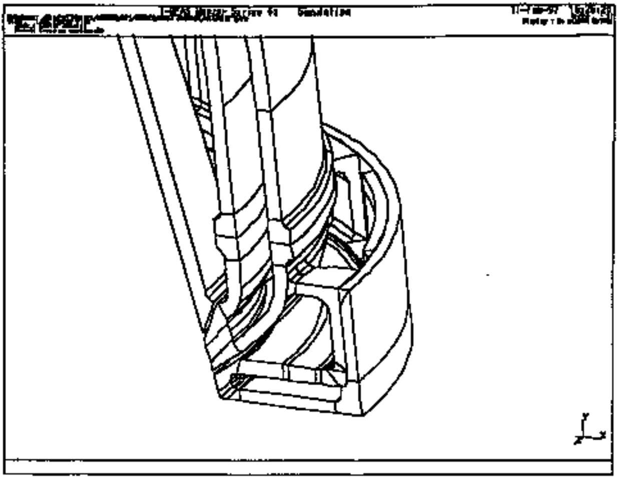

Figure 11. Burner design delivered to finite element code. (Shown is the gas delivery channel at the burner exit.)

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IGCC POWER BLOCK-LESSONS LEARNED

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INTRODUCTION

As IGCC Power Plants become a commercial reality with a number of plants already commissioned, it is important to focus on the early operation to best direct the design of the next generation plants toward any gains that can be accomplished and to steer away from the difficult areas. Approximately 20 plants are now committed with 9 in operation and more than 50 on the screen. Al! of these upcoming projects should be able to benefit from interpretation of the lessons learned from these first 9 plants.

Figure 1 shows 15 plants which are committed and an overview of their applications. Note that eight different gasifiers are used due to the unique circumstances of the various projects. Plant size varies from 40 MWs to 550 MWs. The first four projects were heavily subsidized to support first of a kind costs but the others have not had subsidies for the plant cost. Seven plants use coal fuel, five are for cogeneration, two are for repowering of existing plants and four coproduce power and chemicals. This is a very good spread of potential market applications. Eight of these plants are in commercial operation or startup providing some indications of the potential for the technology.

Since IGCC technology is used for many fuels and applications, there are some generic lessons learned and some specific issues that must be considered on a case by case basis. The generic issues can be categorized by areas of concern where the design varies from established Combined Cycle practice as well as the economic features needed to make these plants profitable for the owners. Figure 2 illustrates some of these areas of concern.

Early operation results will also be discussed. The technical parameters for the Combined Cycle such as; parts life, operability, safety and reliability/availability are most important at this early stage. Current data shows good results for the power blocks. In addition, the economic features developed to enhance the cost effectiveness appear to be proving out in a careful step by step introduction.

DEVELOPMENT PROGRAM

A combined cycle technology development program for IGCC applications, established in 1978, produced the successful Cool Water 120 MW Plant in 1984. In 1990, the program was reestablished to capture the gains from the modern gas turbine development (Figure 3).

System development with various gasifier suppliers has advanced to some simplified systems for air blown and oxygen blown gasifiers. These systems have now been applied successfully in a 4 stage step by step introduction, from steam side only integration to partial air side and finally full air side integration. For oxygen blown gasifiers, the system can be arranged for non integrated air side - 0% air extraction with or without 100% nitrogen return. Air extraction with a high pressure Air Separation Unit (ASU) can be accommodated with partial extraction-up to 50% preferred-or for some cases full air side integration with 100 % of the air needed for the ASU extracted. This amount of air is usually less than 20% of the gas turbine air flow but it must be extracted without affecting the cooling air for the turbine components.

Combustion development jointly funded by users and suppliers has been completed to establish compatible hardware to match the system development. It is important to consider the successes and failures case by case. Field measurements are confirming the full scale laboratory test program.

During development, GE discovered that IGCC systems lead to combined cycle enhancements that improve overall economics. The most notable enhancement ia a 20% higher rating for gas turbines (GTs) with syngas combustors than for conventional fueled GT's. The enhanced ratings derive from the increased flow of the low caloric value syngas, and are limited by the specific turbine design criteria such as pressure ratios and torque ratings. The limits are different for each machine. Several models with the enhanced ratings are now in successful operation, proving the designs and allowing the developer to choose the most economical arrangement.

Next generation GTs will also benefit from the lessons learned on these early IGCC applications. For GE's H machines we expect to obtain another breakthrough in single train size with its large effect on efficiency and plant cost. Gains in efficiencies to 50% along with plant cost reductions of 200 $/KW can be projected.

OPERATING EXPERIENCE

Until 1995, the most important IGCC operating experience was established at Southern California Edison's Cool Water 120 MW IGCC plant (Figure 4). This plant used a 2000F/1100C class gas turbine and operated for 27000 hours at availability's of 80% in the last two years setting the bench mark for IGCC. Power availability was 95%. Today's "F class" gas turbines with higher firing temperatures, have shown more economic efficiencies using the same IGCC configuration.

A comparison of syngas constituents for the various projects gives some perspective to the variety of applications (Figure 5). Columns 1,2,5,6,7 and 8 are coal fueled. Columns 5,6, and 7 are air blown gasification while the others are oxygen blown. This chart shows plants like El Dorado where nitrogen integration is used, the gas turbine combustor does not know whether it is air or oxygen blown.

The first rule of IGCC is that "SYNGAS WILL FIND ITS WAY TO AIR" with the potential to auto ignite. In 1995, the PSI-Wabash River Repowering Project came on line demonstrating the first use of a model FA - 2300F/1260C class gas turbine in IGCC application (Figure 6). The rating was confirmed at 192 MWs - enhanced by more than 20%. In addition to moisturized syngas, steam injection controls No_x below 20 ppmvd at 15% Oxygen. There is no air side integration. Successful transfers from distillate to syngas and vice versa are accomplished on an automatic basis. Full load trips have been accommodated. The major lesson learned here was that more careful attention needs to be placed on nitrogen purge control and fuel piping arrangements. At 260 MWs size, the fuel piping and control valves become very large even without nitrogen integration.

Tampa Electric's Polk IGCC station started the gasifier on July 17th, 1996 with a 20 hour run (Figure 7). The CC plant had been operating since spring and on Sept. 12th the first successful gas turbine operation on syngas was demonstrated. The design does not use extraction air but incorporates nitrogen return for No_x control. More importantly for a warm climate, the performance has established the capability to extend the enhanced rating to 90F/ 32C. This plant also incorporates an experimental Hot Gas Clean Up (HGCU) slip stream of about 35 MWs which will be demonstrated in 1997. The gas turbine fuel control system and materials must be compatible with the high temperature fuel mixtures.

Sierra Pacific's Pinon Pine IGCC plant uses the first 6FA machine combined with an air blown fluidized bed gasifier (Figure 8).This plant also produces syngas at high temperature -1000F/ 538C. 100% air is extracted to feed the gasifier. This feature has been demonstrated while on natural gas fuel without effecting the GT cooling flows. While the CC is operating commercially and has confirmed the combustion development program design for the back up fuel, syngas operation must wait for the gasifier start up scheduled for early 1997.

Texaco's El Dorado IGCC plant uses Petroleum Coke and waste oils to produce syngas for a 6B gas turbine cogen unit (Figure 9). The power plant maintains electrical load for the refinery and its reliability is critical to refinery operations. The CC was commissioned on natural gas in March of 1996. The first transfer to syngas operation occurred on Sept. 12th and was successful on the first try. Co-firing of syngas/natural gas has been very successful. This unit has demonstrated full air side integration with 100% air extraction to the gasifier and full nitrogen return for No_x control. Air extraction and nitrogen injection can also be used when the unit is operating on natural gas to ease the IGCC start up and fuel switching. Again, the fuel and nitrogen delivery system and its valving provided many of the lessons learned.

Schwarze Pumpe's IGCC plant in Germany is the first operating Coproduction plant as was reported in San Francisco October 1996 (Figure 10). This feature can be very important to power producers as it allows electricity supply to be varied significantly while maintaining full write off on the gasification plant. The plant feeds Lignite, Waste oils and Plastics to many gasifiers which provides a wide variety of syngas compositions. Lessons learned at the previously discussed plants were used to modify this gas turbine configuration prior to start up allowing a safe and easy switch over from the back up distillate fuel to syngas operation. Special precautions are needed to protect the distillate fuel nozzles from coking up. Since this unit must operate on a great variety of syngas compositions due to different types of gasifiers, we can anticipate more lessons learned during the first full year of operation.

Ilva's Taranto plant in Italy, while not literally an IGCC, utilizes IGCC technology within the power block (Figure 11). The same configuration with full flow, full pressure and full temperature combustion testing produced a design for this steel mill application. Operation of these 3 units is quite successful. In addition this plant incorporates single shaft fuel gas compressor on the GT shaft. The start up of the plant has again confirmed the need for attention to the details in considering the "first rule of IGCC." Designs must take into account that valves do teak so adequate purge techniques must be met.

UPCOMING IGCC PROJECTS

Now that start-up has occurred for many IGCCs and the early experience can be factored into the next plants, it will be important to follow the first few years operations to determine where improvements can be made. These plants and the other important ones in Italy as well as Portugal will provide a steady source of knowledge to IGCC designers (Figure 12). It should be readily apparent where the successes are and how to avoid problem areas in the future. Each of the next IGCCs will demonstrate specific features leading to the large size plants.

Shell International's Pernis IGCC concentrates on production of Hydrogen with coproduction of power and steam (Figure 13).

IBIL's plant in India will be the first power only IGCC plant using Lignite (Figure14). The air blown fluid bed gasifier was developed in Finland. An extensive Combustion development program has been completed and the gas turbine has been shipped. It will operate on Naphtha until the gasifer is ready for operation.

The Fife project has been planned for operation at the original site of the BGL gasification pilot plant in Scotland (Figure 15). It is planned for cofiring of Nat. gas with syngas derived from sludge.

Star's refinery IGCC plant in Delaware is developing final costs for the investment decision in 1997 (Figure 16). This plant can benefit directly from the lessons learned at El Dorado.

Sarlux IGCC, sponsored by Saras and Enron has been financed on a non recourse basis and is well under way toward finalizing the detailed design (Figure 17). Project financing dictated the scaleup allowed for the gasification system and the CC followed with matching trains. This forced some extra costs but the reported EPC plant cost numbers indicate viable economics depending on the cost of fuel.

The GSK plant is the first commercial IGCC in Japan and because it uses only 2 trains for the same size as Sarlux, it has even lower costs (Figure 18). It was chosen in the first round of IPP bidding in Japan because it was the lowest cost of electricity proposed. This plant faces the challenge of designing for load cycles consistent with economic dispatch for Japan where electricity value varies more than 2-1 between daytime and nighttime.

CONCLUSION

There has been a recent flurry of IGCC activity after ten years of low interest due to poor early economics. Gas turbine combined cycle advancements, as well as more innovative applications has opened the door to a new round of commercialization. Recent experience in most plants indicates modern combined cycle systems can be modified to operate well when properly integrated with gasification plants. IGCC should be able to address the growing market place demand for clean and economical power from coal and heavy oil by incorporating the lessons learned and establishing reliability records.

Customer	C.O. Date	MW	Application	Gasifier	Power Block
SCE Cool Water - USA	1984	120	Power/Coal	Texaco - O_2	107E
PSI/Destec - USA	1995	260	Repower/Coal	Destec - O_2	7FA
Tampa Electric - USA	1996	260	Power/Coal	Texaco - O_2	107FA
Sierra Pacific - USA	1996	100	Power/Coal	KRW - Air	106FA
Texaco El Dorado - USA	1996	40	Cogen/Pet Coke	Texaco - O_2	6B
SUV/EGT - Czech.	1996	350	Cogen/Coal	ZUV - O_2	209E
Schwarze Pumpe - Germany	1996	40	Power/Methanol/Lignite	Noell - O_2	6B
ILVA - Italy	1996	450	Power/BFG	Steel	3x9E
Shell Pernis - Netherlands	1997	120	Cogen/H_2/Oil	Shell/Lurgi - O_2	206B
IBIL/Sanghi - India	1998	60	Power/Lignite	Tampella - Air	106B
FIFE - Scotland	1998	80	Power/Sludge	BGL	6FA
Sarlux/Enron - Italy	1998	550	Cogen/H_2/Oil	Texaco - O_2	3x109E
EXXON - USA	1999	40	Power/H_2/CO/Pet Coke	Texaco	6B
STAR - Delaware	1999	240	Repower/Pet Coke	Texaco	2x6FA
GSK - Japan	2001	550	Power/Oil	Texaco	209EC
		3280			

Figure 1. Current IGCC Projects

- Technical Parameters Affecting Parts Life
- Operability - Start/Stop
- Safety
- Reliability/Availability
- Environmental Performance

- Economic Features/Cost
  - Integration Technique
  - Enhanced Ratings
  - Multi-Fuel Capability
  - Coproduction
  - Repowering
  - Differential Fuel Cost
  - Bankability

Figure 2. IGCC Power Block Design Areas of Concern

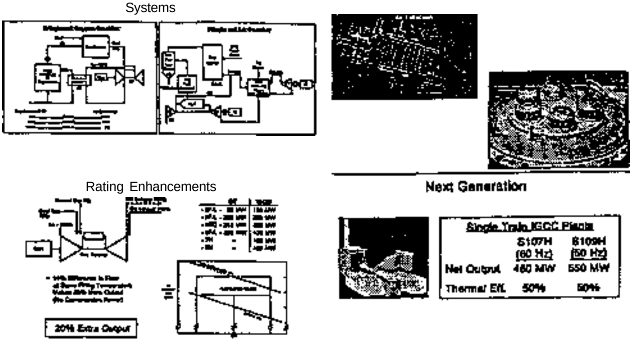


Figure 3. IGCC 10 Year Development Program

- 120 MW
- 27,000 Hours, 5 Years
- Emission: 1/10 of Regulations
- Ash Safe and Saleable

7-E Technology		7-F Technology
32% Efficiency	B K 1m^	42% Efficiency
120 MW	^	265 MW

Figure 4. Cool Water IGCC

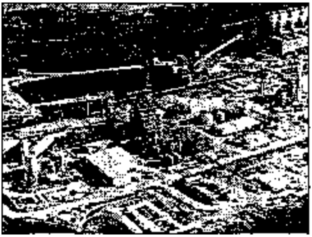
Syngas	PSI	Tampa	El Dorado	Pernis	Sierra Pacific	ILVA	IBIL	Schwarze Pumpe
H_2	24.8	27.0	35.4	34.4	14.5	8.6	12.7	61.9
CO	39.5	35.6	45.0	35.1	23.6	26.2	15.3	26.2
CH_4	1.5	0.1	0.0	0.3	1.3	8.2	3.4	6.9
CO_2	9.3	12.6	17.1	30.0	5.6	14.0	11.1	2.8
N_2 + AR	2.3	6.8	2.1	0.2	49.3	42.5	46.0	1.8
H_2O	22.7	18.7	0.4	–	5.7	–	11.5	–
LHV, - Btu/ft³	212	202	242	209	127	193	115	318
- kJ/m³	8350	7960	9535	8235	5000	7600	4530	12,520
T_{fuel}, °F/°C	570/300	700/371	250/121	200/98	1000/538	400/204	1020/548	100/38
H_2/CO Ratio	.63	.75	.79	.98	.62	.33	.83	2.36
Diluent	Steam	N_2/H_2O	N_2/Steam	Steam	Steam	–	–	Steam
Equivalent LHV								
- Btu/ft³	150	118	113*	198	110**	–	–	200
- kJ/m³	5910	4650	4450	7800	4334	–	–	7880

* Always co-fired with 50% natural gas
** Minimum range

GT26217

Figure 5. Syngas Comparison

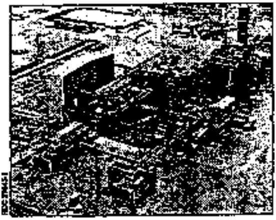
- 192MW7FA
- Repowering - 262 MW
- Dow Gasifier 2 x 100%
- 1995 Operation
- Coal Fuel
- 1400 $/kW-1995$



Confirmed 2300 F/1260 C Class GT With IGCC Enhanced Rating

Figure 6. Wabash River Repowering Project-PSI Energy

- **260 MW IGCC**
- **192** MW7FA Flat **Rating**
- Texaco Gasifier
- Nitrogen Injection
- HGCU Demonstration
- First GT Power From Syngas - 9/12/96



Confirmed Nitrogen Injection for NO_x Control
and Enhanced Flat Rating to 90 F/32 C

Figure 7. **Tampa** Electric-Polk **IGCC Project**

- 100 MW - 6FA IGCC
- Coal Fuel
- KRW Gasifier
- Fluid Bed/HGCU
- Full Air Extraction
- Unit Shipped February 1996
- Syngas Operation 1997

CC Operation Has Confirmed Air Extraction
Capability Without Effect on Cooling Flows

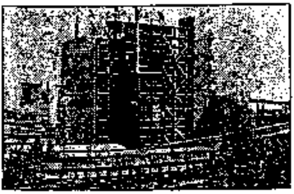
Figure 8. Pinon Pine Project-Sierra Pacific

- Texaco Refinery - El Dorado, Kansas
- 1xMS6001B
- Texaco Quench Gasifier
  - Pet Coke/Waste Oil
- Multi Fuel With N_2 Return and Air Extraction
- First GT Power From Syngas-9/12/96

Confirmed Full Air Side Integration With Nitrogen Injection

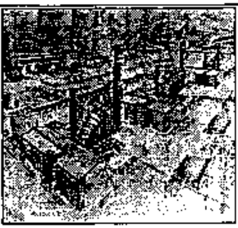
Figure 9. **El** Dorado IGCC Project

- 40 MW 6B Gas Turbine
  - Syngas
  - Methanol Purge Gas
  - Natural Gas/Distillate
- Power and Methanol
- Noell Gasifier Combined With Fixed Bed Gasifier
- Lignite/Oil Slurry With Waste Plastic & Waste Oil
- First GT Power On Syngas Sept. 1996



Confirmed 6B Development for Syngas/Distillate Combustors in Coproduction Plant

Figure 10. Schwarze Pumpe–Germany



- 450MW 3 x 9E
- Blast Furnace/Coke Oven Gas
- Syngas Operation Oct 96
- Integrated Single Shaft Gas Turbine/Fuel Gas Compressor

Confirmed 9E Development Program for BFG

Figure 11. ILVA–Taranto

Shell Pernis	- 115MW	97	6B	Demonstrate Compatability with Hydrogen Coproduction
IBIL	- 60MW	98	6B	Demonstrate Lignite-Based Air-Blown System
FIFE	- 82MW	98	6FA	Demonstrate Compatability with Sludge Feedstock
STAR	- 240MW	99	6FA	Demonstrate Multiple Train and Parallel Train Operation
Sarlux	- 550MW	99	9E	Demonstrate Large 3-Train Coproduction Plant Under Project Financed Conditions
GSK	- 550MW	00	9EC	Demomonstrate Large 2-Train Plant

Lessons Learned to Date Will Make These Projects
- Shorter Cycle
- Lower Cost
- More Reliable

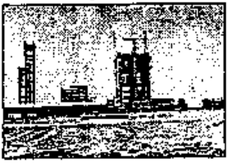
Figure 12. Upcoming IGCC Projects

- 1650 t/d Vacuum Residue
- 2 x 6B Gas Turbines
- Shell/Lurgi Gasifier
- 255 t/d Hydrogen
- 115 MW Power
- Steam to Refinery
- Operation Early 97

Figure 13. Shell Pernis Coproduction Plant

Gasifier

Sanghi Cement



- 60 MW - 106B CC
- Power Only
- Carbona (Tampella) Gasification

- Lignite/Naphtha
- Phased Operation 97/98
- Desal

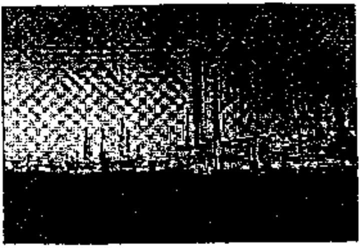
Figure 14. IBIL IGCC Project–India

- 82MW - 6FA
- Sludge/Pet Coke/ Natural Gas
- BGL Gasifier
- 1998

Based on Westfield & Great Plains Experience

Figure 15. FIFE–Scotland

- 180MW2X6FA
- Repowering
- 1999 Operation
- Petroleum Coke
- Texaco Gasifier



GT22911

Figure 16. Star-Delaware IGCC

- 550MW-3xS109E
- Power/Steam/Hydrogen
- Texaco Gasification
- Refinery Residues
- Turnkey-1999
- Project Financed
- Sponsors - Saras/Enron

Figure 17. Sarlux IGCC Project-Sardinia, Italy

- 550MW 209EC IGCC
- Vacuum Resid Feedstock
- 2-Train - Texaco Quench Gasifier
- Sponsors - GSK/Exxon/Mitsui
- Contractors - Chiyoda/Toshiba
- First IPP/IGCC in Japan
- Start Up - 2000

Figure 18. General Sekiyu-Kawasaki Power Generation Project

GAS TURBINE EXPERIENCE ON AND DESIGN FOR SYNGAS OPERATION

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1. Introduction

IGCC is a very promising power generation technology with a wide fuel flexibility. It can be applied to solid or liquid feedstocks such as coal, petroleum coke, heavy oil residues, Orimulsion and biomass. Since the gas produced by gasification (syngas) can be cleaned very effectively of all relevant contaminants, it can be used as a fuel for gas turbines.

Since 1994 a Siemens gas turbine model V 94.2 in the IGCC power plant Buggenum, Netherlands, (figure 1) has been fired with coal gas with a combustion system modified to burn both, natural gas and coal gas and has so far accumulated several thousand hours of coal gas operation. After having generated a total of 70 Gwh in 1994 in about 1000 hours of coal gas operation, 255 MW (gross) at 29°C $T_{am}b$ (design base load 286 MW at 15°C $T_{am}b$) could be reached on April 24th 1995 /1/. The design data referring to efficiency and emissions could be confirmed *121.*

This paper gives an overview of IGCC power plants with Siemens gas turbines. Especially the Siemens syngas burner concept and its application in the Buggenum IGCC plant is discussed. The further use of this combustion concept for the ISAB residual oil gasification project is explained.

2. Syngas Projects with Siemens Gas Turbines

Currently two coal-based IGCC power plants in Europe are equipped with Siemens gas turbines. In the Buggenum coal gasification power plant (Netherlands) (figures 1,3) a V94.2 (T-ISO=1050°C) is applied in conjunction with the SHELL coal gasification process. In the Puertollano plant (Spain) syngas from a PRENFLO coal gasifier is applied as fuel for a V94.3 (figures 2,3) with higher turbine inlet temperatures and higher thermal efficiency (T-ISO=1120°C). The plant is in commercial operation with natural gas. Coal Gas operation is expected for 1997.

An IGCC plant based on TEXACO residual oil gasification is in the design phase for ISAB in Sicily. The plant will be equipped with two V94.2 gas turbines delivered by Siemens/Ansaldo.

For future syngas projects Siemens intends to introduce a new annular combustor machine derived from the recently developped Vxx.3A family (table 2) with even higher turbine inlet temperatures and higher thermal efficiency.

3. Concepts for Syngas Power Plants

An important issue for the performance of an IGCC power plant and for the gas turbine compressor design is the degree of air side integration of the compressor and the Air Separation Unit (cf. figure 4). Another aspect to be considered for the compressor design - mainly for air side none-integrated plant concepts - is the NO_x control philosophy, i.e. fuel gas dilution either with nitrogen and/or via fuel gas saturation (figures 4,5).

The Buggenum and the Puertollano IGCC plant concepts are based on 100% air and nitrogen integration. In this concept all the air for the Air Separation Unit is extracted from GT compressor. The nitrogen from the air separation unit is reintroduced into the gas turbine by compressing and mixing with the undiluted syngas. The turbine mass flow is about the same as for natural gas or fuel oil operation. As a consequence the same compressor as for standard fuels (natural gas or fuel oil) without any modification can be used for the syngas machine .

In a non integrated concept all the air for the air separation unit has to be compressed by an additional compressor. The introduction of the syngas with high flow rates through the burners without air extraction upstream of the burners causes an unusual high difference between the compressor and turbine mass flow. This leads to an increased pressure ratio in comparison to standard fuels and measures have to be taken to keep a sufficient safety margin to the compressor surge limit. For the oil gasification plant in Sicily (ISAB), which is non integrated, Siemens is designing currently a modifk J compressor for the V94.2 gas turbine for optimum performance.

4. Fuel and Combustion system design

4.1 Fuel system

In table 1 an overview for the syngas composition and the mass flow rates for both of the two V94.2 syngas applications - the Buggenum IGCC and the ISAB project - is given. The fuel system for a syngas machine has to supply much higher mass flow rates to the burners than for a standard machine. In the Buggenum case this is about ten times higher than the corresponding natural gas flow (about 9 kg/s natural gas at Base Load), for ISAB five times. The fuel system of the Buggenum gas turbine (cf. figure 6) consist therefore of a pipe work with extremely large diameters (DN600) including the use of flaps (butterfly valves) as control and stop valves.

Syngases are highly toxic and explosive. Correspondingly, the fuel system has to meet very high safety standards. It includes extremely tight piping, evacuation systems and gas detectors. Also inside the piping, the high reactivity of syngases causes risks of explosion when exposed to mixing with air. Therefore a complex purging procedure with steam and nitrogen is carried out before and after syngas operation. In addition all vents are connected to the central flare system. In several thousand hours of operation the gas turbine syngas fuel system has proven to be reliable and safe.

The syngas system of ISAB plant will have reduced diameters because of the lower maximum syngas flow rates, but apart from this it will be similar to Buggenum.

4.2 Burner Design

The Buggenum plant concept includes the use of natural gas as a back up fuel. Consequently a syngas burner with full capability of burning natural gas is used. The syngas burner, which is shown on the right side of figure 7, contains all parts of the standard Siemens hybrid burner (left side of figure 7) which are necessary for the combustion of natural gas or fuel oil. Since natural gas is only a back up fuel, the syngas burner contains no elements for lean premix combustion and steam is used as a diluent. The coal gas channel has an annular cross section and contains a swirler. Because of the extraordenary high mass flow rates of the low calorific syngas the syngas passage has a much larger cross section than a corresponding nozzle for natural gas. Eight syngas burners are situated in each of the two silo combustion chambers. Its outer dimensions allow the syngas burners to be installed into completely unmodified standard flame tubes.

For the ISAB plant oil is used as back up fuel. The syngas burner is combined for this purpose with a standard fuel oil iance.

4.3 Emission Control

An important feature of the burner design is its low NO_x capability. Two basic concepts are used in the design of low NCvsystems for gas turbine combustion: lean premix combustion or lowering the flame temperature of a diffusion flame by dilution of the fuel with inert gases like nitrogen or steam. The Siemens syngas burner uses the second concept. This is possible because of another important feature of syngas: its much higher flame speed in comparison with Natural Gas. The high flame speed offers the possibility to use highly diluted diffusion flames with comparably low flame temperatures but nevertheless good flame stability for syngas combustion /3,4,6/. The use of diluted diffusion flames is favoured by the large amount of dilution nitrogen available in the fully integrated IGCC (Buggenum and Puertollano). For the ISAB project steam is used as diluent. Because of the higher heat capacity of water in comparison with N_2 the amount of diluent added is lower.

Meanwhile, the low emission potential of this diffusion burner concept has been demonstrated by the Buggenum gas turbine for more than two years. Figure 8 shows measured values of NO_x and CO in a broad range of heating values (different degrees of dilution).

4.4 Problems with syngas combustion

While all conventional aspects of combustion, e.g. emissions and flame stabiity, turned out to be as positive as predicted by test rig results and calculations /3,4/, the coal gas operation was not entirely free of flame induced pressure oscillations, which can lead to unacceptable high combustion noise levels. These oscillations occured during the first coal gas base load experience in 1995 at high loads. Consequently Siemens started a burner optimization program during which the problem was solved. In September 1996 gas turbine base load operation with 291 MW at 12°C ambient temperature without flame induced pressure oscillations could be demonstrated. In november 1996 the plant was operated continuously for about

4 weeks at high loads including base load without shut down. This was followed by a short combustion chamber inspection which showed no findings.

One factor that favored flame induced pressure oscillations in Buggenum is an extremely low pressure drop across the burner syngas nozzles during operation with undiluted syngas /5.6A For the ISAB project the fuel pressure drop has been increased considerably. So pressure oscillations caused by to low pressure drop will be avoided.

5. Summary

For the use in IGCC plants standard gas turbines need specific modifications, among which the syngas combustion and all operational demands resulting from combination with a gasifier and an air separation unit are very important. Among all IGCC concepts the fully integrated IGCC in Buggenum clearty sets the highest standards with respect to versatility of the combustion system (e.g. wide range of heating values and gas compositions) and gas turbine operational flexibility (e.g. gas turbine has to control sudden trips of gasifier, air separation unit and nitrogen supply). During the demonstration period in Buggenum the capability of the gas turbine to meet all these complex requirements has by now largely been proven.

The concept of diluted diffusion flames for syngas combustion with very low NO_x emissions was applied successfully in the Buggenum plant with V94.2 gas turbine. The same concept is used in the Puertoilano plant with a V94.3. It will be adapted also to other syngases especially from residual oil gasification in the ISAB project in a V94.2. The use of this technology in Annular Combustor machines of the .3A-series is intended.

The Buggenum experience constitutes a valuable basis for the design of Siemens syngas operated gas turbines, which are available for gasification combined cycle plants with full, partial or no integration.

IV S. de Ligt ,"255 MW uit kolengas, Eerste resultaten Demo KV-STEG in Buggenum", INSPANNING, Uitgave van N.V. Sep Jg.5, Nr.2, Juni 1995

121 H.M.J, de Winter, W. Willeboer, "Design, Construction and Startup of the 250 MWE IGCC Plant Buggenum", 13th EPRI Conference on Gasification Power Plants, San Francisco, California, Oct 19-21, 1994

/3/ B. Becker, B. Schetter, H. Termuhlen,"Low-Emission Combustion in Advanced Gas Turbines with Conventional and Coal-Derived Fuel", Tenth Annual Conference on Gasification Power Plants, San Francisco, California, October 17, 1991

IAI P. Berenbrink,"Hochtemperaturbrennkammer fur Kohlegaseinsatz - Abschlu3bericht", BMFT 03E-6232-C, September 1990

/5/ N. Vortmeyer, B. Schetter, B. Becker," Verbrennung von Kohlegas in Siemens-Gasturbinen: Erfahrungen bei der Inbetriebsetzung des Kohlevergasungs/GuD-Kraftwerkes Buggenum", 17. Deutscher Flammentag, VDI Bericht Nr. 1193, VDI-Verlag, Dusseldorf, 1995

/6/ N. Vortmeyer, M. Huth, B. Schetter, B. Becker, J.Karg, W. Emsperger, "Experience in the Design and Operation of Syngas Gas Turbines", EPRI Gasification Technologies Conference, San Francisco, California, Oct 2 -4, 1996

		H2	CO	N2	H20	Heating value	fuel flow rate at Base Loac	Fuel tern-1 perature |
| Buggenum coal gasif.	undiluted	30 Vol%	65 Vol%	3 Vol%	1 Vol%	11-12 MJ/kq		
	diluted	12Vol%	25 Vol%	42 Vol%	19Vol%	4.3 MJ/kg	106 kg/s	310°C i
ISAB 1 oil gasif.	undiluted	42 Vol%	49 Vol%	1 Vol%	0 Vol%	12-13 MJ/kg		
	diluted	27 Vo!%	32 Vol%	1 Vol%	35 Vol%	8,6 MJ/kg	52 kg/s	195°C

Table 1: Composition, properties and mass flow rates of syngases from coal and residual oil gasification. The diluted fuel gases are the gases that are burned in the Siemens syngas Burner.

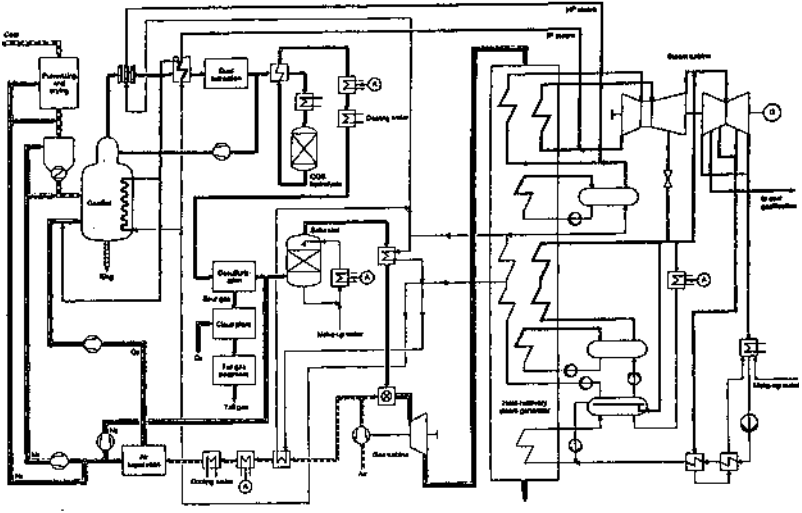
Type	I	[V84.2	V94.2	V64.3	V84.3	V94.3	V64.3A	V84.3A	V94.3A
Frequency	Hz	60	5O	50/60')	60	50	50/60[1]	60	50
ISO-Temperature	C	1060	1060	1130	1160	1160	1190	1190	1190
Compr. press, ratio		11,0	11,1	16,1	16,1	16,1	16,2	16,8	16,2
Exhaust gas flow	kg/s	360	519	192	433	624	192	449	629
Exhaust gas temp.	C	544	540	531	551	550	571	560	571
SC Power output*'	MW	109	159	61,9	153	221	69,3	168	232
SC Efficiency?)	**%**	34,0	34,5	35,0	36,1	36,5	36,5	38,0	37,3

ISO-Conditions; Fuel: Natural gas; Base load

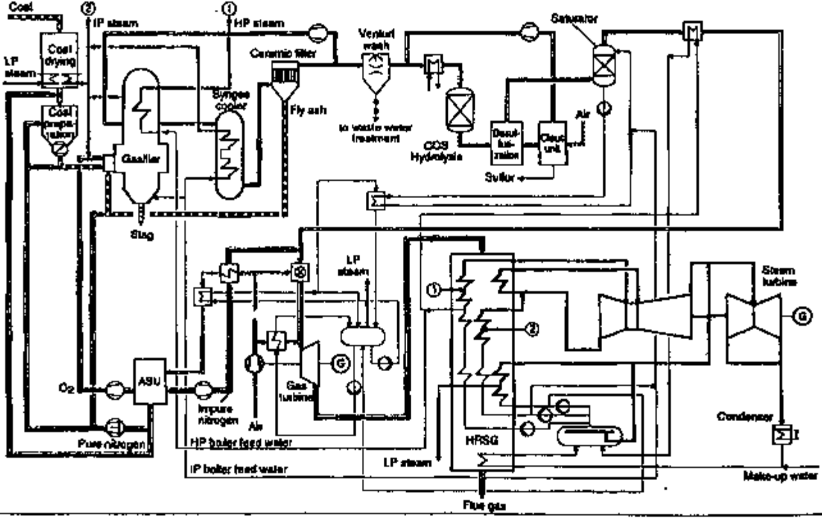
e losses in fron of the compressor and behind the turoine

Table 2: Siemens gas turbines, thermodynamical data (Status February 1996)

SIEMENS


Coal
Pulverizing and drying
Gasifier
Slag
Dust removal
COS hydrolysis
Cooling water
Desulfurization
Sour gas
Claus plant
Tail gas
Make-up water
O2
N2
Air separation
Cooling water
Air
Gas turbine
HP steam
IP steam
Steam turbine
G
to coal gasification
Heat-recovery steam generator
Make-up water

Puertollano IGCC 1.94.3 Block
Featuring Oxygen-Blown PRENFLO Coal Gasification



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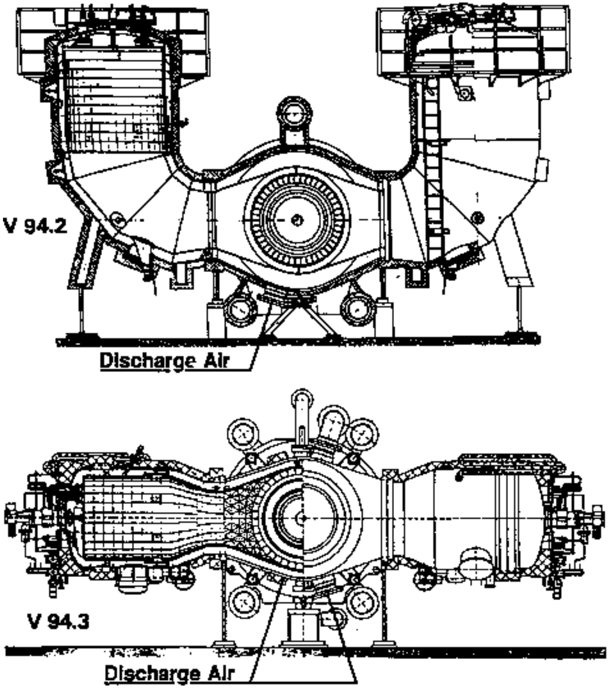
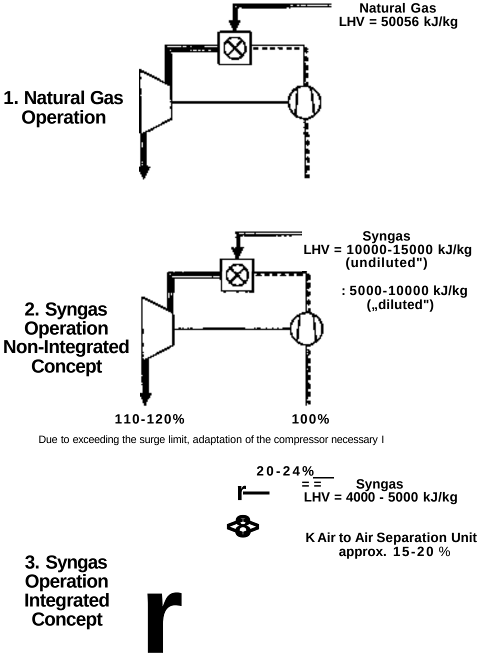


J Figure 3: V 94.2 - V94.3: Cross sections with location of discharge air flanges



1. Natural Gas Operation

2. Syngas Operation Non-Integrated Concept

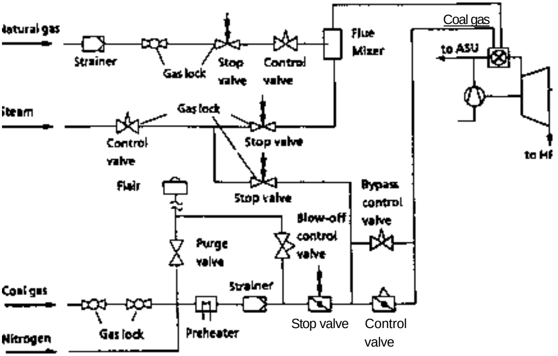
3. Syngas Operation Integrated Concept

Due to exceeding the surge limit, adaptation of the compressor necessary I

Figure 4: Influence of integration concept on gas turbine/ compressor mass flow ratio

	Dimension	DEMKOLEC	ELCOGAS	ISAB Energy
Number of gas turbines	-	1	1	2
Air extraction for ASU from 1 gas turbine				
• related to ASU demand	%	100	100	0
• related to compressor mass flow	%	16	18	0
Nitrogen for syngas dilution (related to ASU production)	%	100	100	0
Fuel gas saturation				
• syngas			+	+
• syngas/nitrogen mixture	-	+		
Fuel gas temperature	°C °F	300 572	302 576	195 383

Figure 5: Comparison of main gasside and integration characteristics



VG8 Kraflwerkslechnik 74 (1994), Heft 6

Figure 6: ICoalGCC Buggenum, fuel supply system for coal gas and natural gas

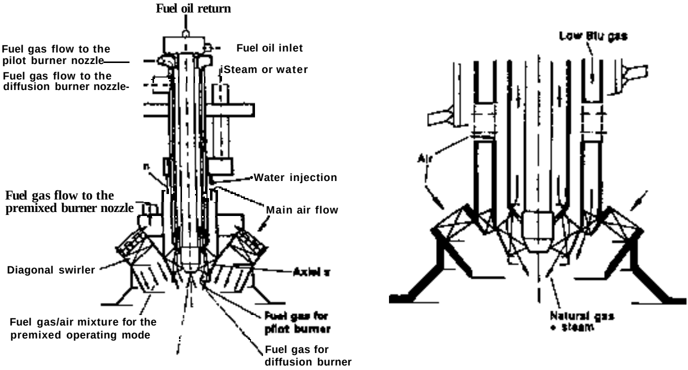


Figure 7: Siemens standard hybrid burner Siemens syngas burner

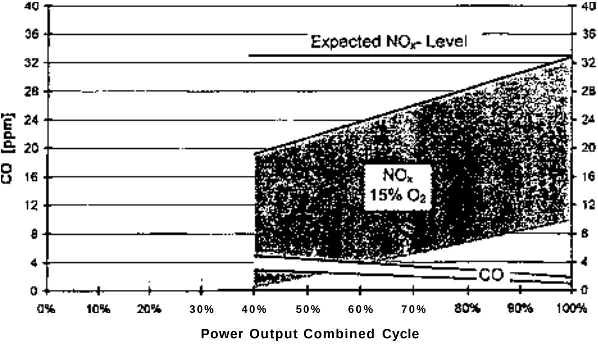


Figure 8: Measured CO- and NOx-emissions, coal gas operation ICGCC Buggenum, Siemens gas turbine model V 94.2

Subject:

Paper for :" Gasification Technology in Practice "

Authors:

M. Baldizzone R. Macchia P. Zanello
(FIAT AVIO " Power plant design and construction - Application Engeneers -")

Title:

ELECTRICITY FROM NON CONVENTIONAL FUELS IN HIGHLY EFFICIENT POWER PLANTS

INTRODUCTION

Gas Turbine is recognised as the best prime mover for modern Power Plant based on Simple Cycle as well as in Combined Cycle configuration according to the features of the specific Project.

Presently Gas Turbines manufacturers have two main objectives:

- Technological improvement in order to get better performance and in this way to reduce fuel consumption involving new materials as well as component design
- Proper engine and system design in order to allow as much as possible the direct use of fuels typically not palatable for gas turbines

Capability of burning a wide range of fuels in a Gas Turbine power Plant has become, in the last years, a very important factor, due to fuel availability, quality and cost.

The direct use of fuel into the Gas Turbine is widely the most economic way to use it unfortunately there are limits in physical and chemical properties that can not be overcome from engine as well as system stand point.

Gasificationis a typical process used to get fuel palatable for the gas turbine even if the resulting gas, because of its low calorific content, requires some engine readjustment in order to allow the increased mass flow through the turbine.

The available experience in both fields of FiatAvio - Mitsubishi H.I. - Westinghouse GT family, direct use of contaminated fuels as well as burning low BTU gas from gasification process, is reported highlighting the most important technological aspects associated to the engine and the system in both cases.

CONTAMINATED FUELS

Fuel availability and costs considerations are the key parameters for selection of residual oil as primary fuel. Its price is generally variable, but as a general rule a difference between twenty to one hundred percent compared with light oil price can be assumed in economical evaluations. This evaluation has to be based on many variable effects such as electric power selling price, power plant running time, fuel costs, fuel treatment and additivation costs, additional maintenance. An accurate analysis and its direct validation through actual plant operation experience is necessary for a final answer. On a general basis, considering that treatment costs will never increase fuel cost, by more than 20% economical benefits are typically obtained.

CRITICAL ASPECTS OF BURNING CONTAMINATED FUEL IN GAS TURBINES

These fuels have a high level of contaminants because they contain all these already existing in the crude, plus some introduced in the refining process and transportation. The effect of these compounds can be a corrosive action on the Gas Turbine hot parts.

An additional area of possible problems is related to the low level of Hydrogen and the considerable presence of Asphaltenes. These components are responsible for high flame radiation, causing severe operating conditions for turbine hot parts. These effects can be compensated by an adequate reduction of engine firing temperature.

In addition to that, residual fuel physical characteristics', namely density and viscosity, require care in fuel handling to assure adequate pumping and atomisation capability.

The above mentioned corrosive phenomena, due to Sodium, Vanadium and Potassium, occur at high temperature involving the turbine components working on direct contact to the combustion gases at temperature above 600 °C.

Usually they are due to a reaction between the oxides and melted substances present in the combustion gases.

The corrosion occurs when the protective oxide layer of the metal surfaces is broken. As it is destroyed, non oxidised metal is exposed and attacked with a continuous process. In particular the corrosion propagation due to Sodium attack is inter-granular, and therefore the material matrix is disintegrated with consequent decrease of the heat resistance of the alloys, especially under the action of mechanical and thermal stresses. The improvement of high temperature corrosion resistance of Nickel and Cobalt alloys has therefore become a key point both under a technical and economical viewpoint.

Sodium and Potassium, combined with Vanadium, form eutectic salts whose melting point is lower than 566 °C; when combined with sulphur, they form sulphates which start their corrosive effect at a temperature which is in the operating range of the Gas Turbine.

Na + K level must thus be limited and considering that they are water soluble this is possible by means of fuel washing.

Vanadium is present in the fuel in the form of metal-organic compound (Vanadium - Porphirin) and cannot be removed with chemical or physical treatments; its negative effects are inhibited with an oil soluble Magnesium additive.

The Magnesium will combine with Vanadium to form compounds which have a melting point higher than turbine operating temperature. It is very imponant to assure that all the Vanadium is inhibited by the additive; on this regard a consolidated experience shows an optimum additivation ratio Mg/V=3.

An adverse effect of Magnesium additivation is related to consistent deposits of Magnesium compounds on turbine parts.

These deposits can be responsible for engine performance deterioration, combustion asymmetry with increase of temperature spread and sometimes modification of engine vibration behaviour. The deposits are however water soluble, so that they can be removed by proper turbine water washing which frequency is considerably depending on engine operating mode. Frequent starts and stops are typically beneficial with the deposit spalling off subsequent to the thermal transient phases.

The physical fuel characteristics require special care in plant design. Because of its high viscosity, the residual fuel must be heated for handling purposes and to ensure adequate atomisation in the Gas Turbine injectors.

Moreover the fuel requires a specific operation procedure. It is normal to start and stop the Gas Turbine on distillate oil to avoid plugging of piping, filters and nozzles downstream the heaters; for the same reason fuel pipes must be purged with distillate oil after every emergency shut down or trip.

Another key point is a constant control and monitoring of the fuel quality to tune the treatment system to the actual fuel composition.

FUEL TREATMENT

The proven approach to overcome the above mentioned turbine corrosion problems is based on:

- fuel washing in order to reduce Sodium and Potassium contents;
- fuel additivation to inhibit Vanadium corrosion effect.

Fuel washing

In order to reduce the contents of Sodium and Potassium FiatAvio has adopted in the latest application a two stage electrostatic fuel treatment system.

Nominal performance for these treatment plants, is a Sodium plus Potassium content in the residual oil at treatment system outlet not exceeding 0,5 ppm.

The untreated residual oil is heated first by means of steam to a proper temperature to reach the optimum viscosity for the electrostatic process. Subsequently, fresh water is added to the residual oil and is mixed with it in order to dilute the water phase, that contains all the contaminants. Fresh water i; added to the second stage and then pumped from a re-cycle water pump to the first stage in order to get the maximum efficiency at the final stage with a counter flow effect. In order to minimise the water requirements inside the pressurised vessel, the electric field created by a high voltage transformer (short circuit proof type), increases of thousands of times the phenomenon of water droplet coalescence, thus achieving a separation of the two phases:

- In the bottom of the vessel, water that contains almost all the Sodium plus Potassium previously dispersed in the fuel;
- At the top of the vessel, a water free residual oil.

De-emulsifier is injected before each stage.

Fuel additivation

A Magnesium organometallic oil soluble additive injection system is used to inhibit the Vanadium compound. The system consists of a tank, a circulation pump and dosing pump which inject the additive in the high pressure line downstream the fuel injection pump of the Gas Turbine.

The additive injection point has been selected near to the Gas Turbine fuel nozzles, in the high pressure line downstream the final filter, and upstream the fuel flow divider in order to have the best mixing of additive with fuel and therefore the best inhibition effect.

FUEL SYSTEM

The fuel system of a plant burning residual oil is typically composed by:

- Distillate oil system
- Residual oil system

Distillate oil is adopted to start up and shut down the Gas Turbine and consists of standard system with no specific features.

Residual oil system shall be designed to overcome the above mentioned physical characteristics of residual oil and perform fuel treatment.

From the unloading bays the residual oil is stored in the untreated tanks, than is transferred to the treatment plant to remove water soluble Sodium and Potassium.

The treated residual oil is then forwarded to certification tanks. When one of the two certification tanks is full, the treated oil is certified by chemical analysis. If the contaminants level comply with fuel specification, the fuel is transferred by the pump station to the Treated storage tanks. If the fuel does not comply with specification it is sent back to the untreated tanks to be treated again.

All the tanks of the residual oil handling system are insulated and provided with bottom heating coils capable to maintain the residual oil at the temperature which allows fuel handling (typically 50°C).

Each Turboset is fed by a residual oil forwarding pump. A final heater is installed on each Gas Turbine supply line upstream the fuel change over valve.

An automatic fuel change over valve is used to switch from distillate oil to residual oil after start up and vice versa during shut down. The automatic fuel change over valve, one for each GT., is a three ways valve, electrically operated, installed on the forwarding pump station. One three ways valve, one for each GT., is also installed on the recirculation line to tanks: the valve is necessary to send back residual oil to residual oil tank and distillate oil to distillate oil tank.

LOW BTXJ GAS

The integration of a combined cycle with a gasification plant is an evolution imposed by different reasons:

- availability of consolidated gasification technologies to gasify different low commercial value products or to gasify refinery slag.
- increased quantity of those products caused by ambient protection regulations aggravation
- government aids to auto production
- high commercial value by-products

The integration of a combined cycle gas turbine in such a plant requires a deep analysis; the accuracy of the analysis covering all aspects technical as well as economical has an important impact on the initial investments and on operating results.

IGCC plant is competitive respect other plants like:

- Conventional plants with desulfuration and denitrification
- Plants with partial gasification and conventional combustion

They have increased performances like:

- High combined cycle efficiency (approx. 46% referred to electric power in lieu of 34% of a conventional thermal plant, and 90% referred to the total utilised power (eiectric and heating power)
- Production costs similar to those of conventional thermal plants with lower emission values
- Potential technologies evolution in order to increase efficiency (target 52%) e reduce emissions levels.

Otherwise:

- Higher investments costs
- More interconnected systems
- Insufficient acknowledgement to guarantee application of large power plants.

I IMPACT ON THE PLANT

To operate with low BTU gas the GT compressed air feeding the combustion section must be reduced respect to the design value usually based on natural gas or distillate oil. The reduction changes with the ambient temperature and the load. The limitation could be obtained with compressor bleed or inlet air reduction.

The air required by the Air Separation Unit could be fed by the air bleed from the GT compressor. Otherwise an independent air compressor should be installed.

The independent compressor increases the flexibility of the system, has a rapid response in starting or tripping phases, and at part load.

The compressor air bleed increases the global efficiency of the plant. Anyway a complementary independent compressor has to integrate the insufficient air bleed flow. Control system will be adequate to manage much more parameters.

Two possible solutions are typically examined for emission control:

- Nitrogen injection
- Steam saturation

Different is the impact on the plant; to have the same NOx reduction nitrogen mixed quantity is bigger than corresponding steam. It requests larger size piping.

The percentage of steam can be limited; a catalaiser, dimensioned to reach the imposed value, is installed on the exhaust section.

The particular syngas composition (high hydrogen and Carbon oxide content) imposes classification areas, where gas piping is installed, in Class 1 Division 2 Group B according NFPA 70. The interested areas are:

- Gas turbine enclosure
- Auxiliary systems enclosure
- Gas detection and treatment area

Gas detection and treatment area is open air and standard configuration was adequate.

All components installed on the gas turbine enclosure and auxiliaries enclosure are explosion proofing type. In the standard configuration only gas skid components are explosion proofing type. Also fuel oil skid, lube oil skid, water injection skid are equipped with explosion proofing components.

Gas turbine and auxiliary enclosures standard ventilation systems are both provided of 3 50% capacity extraction fans, two in operation and one in stand-by. For this application the system will be provided with two 100% capacity each, fans one in operation and one in stand-by

Purge time will be increased in order to assure the total volume changing 6 times in lieu of 3 as per standard.

Gas piping vent system has been implemented with a nitrogen purge system to assure, at fuel changeover or at trip the total elimination of syngas. A system with bottles, piping, valves and automation was added.

Gas detection system already existing on standard gas turbine and auxiliary enclosures, was been implemented with CO detectors, to assure personnel safety.

IMPACT ON THE ENGINE

The design of a combustor for a syngas application requires special care to:

- Take into account the higher flame speed of syngas, typically with significant amount of $H_{2)}$ several times greater than of methane. Consequently will have a rather short flame compared to that of methane-based natural gas, impacting on the cooling requirements of the combustor dome and on the primary zone local equivalence ratio.
- The amount of fuel injected through the combustor fuel ports is much greater than the typical amount of natural gas. Such a large amount calls for a careful dimensioning of primary, secondary (if any) and dilution scoops (chutes) together with sizing cooling air devices devoted to keep the combustor wall temperature down.

The above impose design choices which might not be enough proven or are totally new. This usually leads to hot rig tests of the combustor prior of installing it into the engine.

One combustor has been designed referring to a preselected fuel composition. As most of the design choices were based on standard practices it was decided to have an extensive run of rig tests to validate the final design. The tests were carried out with the same syngas composition used for design and rig geometry as closely as possible to engine pressure shell configuration.

Low pressure tests

Tests were carried out from No Load (0%) to Base Load (100%) with 20% power increments.

Ignition was attained at a rather low F/A with respect to nominal conditions, to prevent the possible accumulation of large quantities of unlit fuel at the stack.

Weak extinction limits were investigated at each of the above conditions by reducing fuel flow in small steps (~ 10%o) until the exhaust thermocouples indicated flame extinction. Also a rich point (nominal + 20%) was run to investigate flame stability in the rich zone.

Emissions (NO_{Xj} CO and UHC) were recorded at the T/D exit and at the EPA. Combustion efficiency was in excess of 99.7%. Maximum metal temperature topped at 1123 K in the first cooling ring when running the 100% load simulated condition. Visual investigation after these test showed some distortion on the dome cooling skirt indicating temperatures around and over 1173 K, possibly caused by the very short flame occurring at this low pressure conditions.

The combustor was performing with good ignition characteristics and with a wide stability range; also emissions, especially CO emissions were much lower than expected. (See fig. 1 & 2)

High Pressure Tests

Testing conditions simulated engine running at No Load and Base Load conditions.

The ignition was carried out with no problems. For Base load test it was decided to rise the rig pressure by oil running and performing a fuel switch over to syngas once reached the test conditions.

The fuel transfer was successful with no signs of combustion instabilities during the process. During Base Load with syngas a complete exhaust traverse reading at sixteen equally spaced angular positions at T/D exit was done. A simulation of engine load rejection was attempted by suddenly cutting down the fuel flow to the No Load value; no problems with flame stability were experienced.

Syngas emissions test results (all the NO_x emission values are dry - 15% 02 corrected, all the other emission are on dry basis) are given in Table 1. Combustion efficiency was found to be in excess of 99.9%

The highest recorded metal temperature was about 1150 K in the second cooling ring, this showing that the flame peak has moved downstream from the position it took when running at low pressure, as it was expected because of the higher fuel flow. Combustion-induced pressure fluctuations when running the syngas high pressure tests were rather small if not negligible.

On oil running a complete exhaust traverse reading was carried out at each of the test conditions (i.e. with different water/fuel ratios). Emissions results are given in Table 2.

Basket metal temperature were found to be lower than with syngas with a peak maximum temperature of - 893 K. This was expected because of steam injection and because of a longer more narrow flame shape.

Additional tests were carried out to investigate the syngas combustion properties while changing steam content (25%, 36% to 45%, the case with 35% steam added being already tested).

In the end these tests showed that the syngas combustor gave good performance when handling different syngases. CO emissions, source of primary concern, did not increase when increasing the syngas steam content and seemed to reach a plateau of minimum value at nearly the maximum steam content. Stability at all conditions was satisfactory (pressure fluctuations well below 10 kPa).

The emission results for Syngas Base Load tests, that is, the NO_x level, were transposed to the actual engine conditions to give the following (all values are):

Syngas N°l	from 20.4 to 32.4 ppmvd - 15% 0_2
Oil w/ W/F=0.6	from 59 to 85 ppmvd - 15% 0_2
Oil w/ W/F=078	from 45 to 65 ppmvd - 15% 0_2
Oil w/ W/F=0.96	from 35.6 to 50 ppmvd - 15% 0_2

GAS TURBINE

Both reported studies have been based on use of TG 50D5 gas turbine.

The machine is a result of 45-year history of developing and manufacturing large heavy duty combustion turbines for industrial and utility service. It is the current production version of the large 50-Hz combustion turbine first developed by Westinghouse and its licenses, Fiat Avio and Mitsubishi, in the mid-1970s. This development was derived from the highly successful Westinghouse model W501 series of large 60 Hz units, first introduced in the late 1960s. The combined Westinghouse/Fiat/Mitsubishi fleet of such large combustion turbines totals more than 300 operating units. Collectively they display an excellent record of high reliability, availability and economy both in simple and combined cycle applications.

EXPERIENCE

Within the FiatAvio - Mitsubishi H.I. - Westinghouse family operating experience of TG 50D5 is available on contaminated fuels as well as Low BTU gas

Al Nasseriah and Zayzoon Syrian power plant in operation from 1995 with residual (Viscosity = 7 °E, Na + K = 20 ppm, Va = 60 ppm)

Plaquemine (Dow Chemical) and Kawasaki Steel are most important low BTU gas operation experiences:

A 7 years long experience has been accumulated by Plaquemine Power Plant, burning a 239 BTU/SCF gas from coal gasification process (41.41 % H2, 38.52 % CO, 0.11 % CH4, 18.49 %

C02, 1.48 % N+Ar). Extremely good results have been achieved in terms of plant availability and reliability as well as of emission control.

Kawasaki Steel Power Plant is operative since 1987 and has accumulated more than 50,000 **operating hours, burning a 112 BTU/SCF gas, composed by blast furnace gas mixed with cook oven gas, which are by-products in steel works.**

Additional experience are:

Heavy fuels experience

Florida Power and Light, USA, Putnam Plant (1976)

W 5012 x (2 x 2 x l) Contaminated Heavy Fuel

EBES, Belgium, Mo! (1977)

W 251 1 CiSe Contaminated Heavy Fuel

Kelanitissa power plant, Sri Lanka (1997 under construction)

TG50 D5 1 x Ci.Se Contaminated **diesel oil.**

Low BTU gas experience

NSC, Japan, Hirohata works (1989)

MW 251 1 Cogeneration, LDG

Nisshin Steel Co., Japan, Kure works (1989)

MW 251 l x l x l Blast Furnace gas

REFERENCE

- Sawyers gas turbine engineering handbook - vol. 3
- ASME 87-gt-176 - reliable operation of gas turbines on crude oil By S.Gabella - L.Maderni - Fiat TTG-6/1987
- CIMAC - tl 131 - multifuel gas turbine power plant operation by S.Gabella - FIAT TTG - 6/1985
- ASME publication - operating experience on residual fuel oil with a W251 combustion turbine by J.L.Degreef, P.Maes, K.W.Johnson - 12/1977
- Morrison, Pillsbury, 1989, "Coal Generated Synthetic Gas Operating Experience with two 100 MW- Class Combustion Turbine", ASME paper No. 89-GT-257.
- Takano, Kitauchi, Hiura, 1988, "Design for the 145 MW Blast Furnace Gas Firing Gas Turbine Combined Cycle Plant", ASME paper No. 88-GT-97.
- R. Macchia 1994 " 350 MW gas turbine power plant operated on highly contaminated residual oil ", Power Gen'94 ASIA
- P. Zanello, A. Tasselli 1996 " Gas Turbine firing medium BTU gas from gasification plant "ASME paper N° 96 - GT - 8.

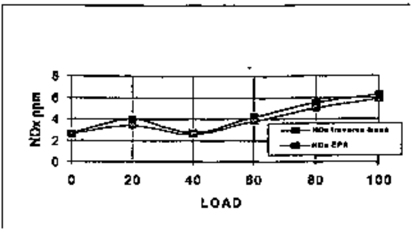


Fig.1 : Low Pressure test NO_x emissions (Syngas)

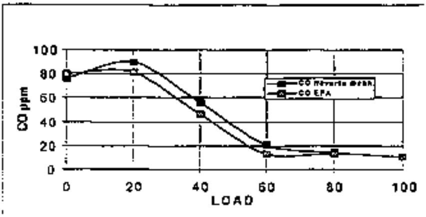


Fig. 2 : Low Pressure test CO emissions (Syngas)

Tab. 1 High Pressure Syngas Emissions

Engine Cond.	Sample Position	No_x [ppmvd]	CO [ppmvd]	UHC [ppmvd]
0%	Traverse Mean	3.0	26.6	0.0
0%	EPA Plane	3.0	24	0.0
100%	Traverse Mean	20.4	0.0	0.0
100%	EPA Plane	22.9	0.0	0.0

1. Table 2 - High Pressure Oil Emissions (with water injection)

W/F	Sample Position	NO_x [ppmvd]	CO [ppmvd]	UHC [ppmvd]
0.6	EPA Plane	59	0.7	1.3
0.78	EPA Plane	45	0	1.4
0.96	EPA Plane	35.6	0	0.9

Stead / CVG
21 Jan 97

Gasification of Refinery Residue

J. Ilg, H. Jungfer, P. Kummann, P. Stead
Linde AG, Germany

ABSTRACT

For the successful implementation of the IGCC concept, the process for the gasification of residual oil must be both efficient and reliable. This paper describes the positive experience which has been gained over the last fifteen years in the construction and operation of plants based on the gasification of residual oil using Texaco quench reactors. The technology is proven, and correct engineering, operation and maintenance leads to high reliability. The construction of IGCC complexes based on this technology can benefit from the experience of the last fifteen years to give a gas generation process of the required reliability for the power industry.

1. REFERENCE PROJECTS

Last year (1996) the gasification complex at Dalian, China, for the production of 1000 MTD ammonia from residual oil was put on stream and handed over to the client. This gasification unit is the most recent of several fertilizer plants based on residual oil which use Texaco quench reactors at high pressure (85 bar).

This plant concept was introduced to the fertilizer industry by Linde for GNFC in Bharuch, Gujarat, India (Figure 1). A detailed process description was presented in paper ref[1].The complex went into production in 1982, and even in the first years, when unreliable power supply caused some production loss, a capacity utilisation of about 80% was achieved. In subsequent years, the complex has been running continuously with record production figures and high reliability. The high pressure gasification process allowed the design of a 1350 MTD ammonia plant based on two gasification reactors. Further, each of the two reactors is rated to produce 80% of the nameplate gas generation capacity of the total plant. This is a key factor for achieving the high reliability and record output demonstrated in the GNFC plant.

2. DETAILS OF THE SELECTED TECHNOLOGY AND ITS PERFORMANCE

Both the Dalian and the GNFC complexes are based on integrated gas generation systems (see Figure 2) which include:
- the Texaco gasification reactors and gas scrubbers for soot removal and the naphtha/soot recycle system for extracting soot from wash water and recycling it to the charge oil feed.
- the Linde air separation unit (ASU) with internal compression for HP supply oxygen to the process.
- the Linde sour gas removal process (Rectisol Wash), giving a purified syngas, and a concentrated sulphur fraction as by-product.

The production record of the high pressure gasification process results from many factors in both conceptual design and detailed engineering. Of particular interest for the gasification sections of the IGCC are the following aspects:-

a) Reliable high pressure oxygen supply.

b) Sufficient reserve gasification capacity to allow routine inspection and regular maintenance of burner, quench ring, and refractory lining in a gasification reactor.

c) Proven control concept to allow smooth shut-down and start-up of a gasification reactor within the operating complex.

d) Burner platform lay-out.

e) Correct materials of construction and corrosion protection measures.

These aspects are discussed in more detail below.

2.1 RELIABILITY AND AVAILABILITY OF OXYGEN SUPPLY

For a reliable high pressure oxygen supply, Linde has developed an advanced version of the "internal oxygen compression" concept, where liquid oxygen is pumped to high pressure within the low temperature air separation unit.

> The critical oxygen turbo compressor is avoided. For gaseous oxygen turbo compressors, the high reactivity of oxygen at ambient temperatures and above requires that stringent safety regulations must be observed. Despite the stringent regulations, fires do occur on oxygen turbo compressors, and there remains a finite risk of damage to equipment and loss of production. Also, the availability of the overall complex is reduced by shut-downs initiated by the automatic safety system of the oxygen turbo compressor.
>
> The liquid oxygen pumps are variable speed triplex piston pumps with low piston velocity, and operating at-180°C, risks of oxygen reactivity are eliminated.
>
> The cold in the high pressure liquid oxygen can conveniently be recovered against a recycle flow of air or nitrogen. The compressor for this duty is not critical.
>
> A reserve liquid oxygen pump is installed, and is normally running, so that each delivers approx. 50% rate. On shut-down or trip of one pump, the second pump takes over the full rate with no disturbance in supply to the gasifiers. Duplicate oxygen pumps give a high security oxygen supply.

In comparison, a reserve oxygen turbo compressor is not always installed, due to the high cost, and would anyway not be able to automatically take up load in case of a trip of the first compressor. When the supply of high pressure oxygen is by turbo compressor, a back-up system is often considered to allow continued operation of the gasification unit if the compressor trips. The back-up system consists of:

a) high pressure storage of oxygen available at all times for immediate supply to the gasifier. The pressure and volume of the storage is enough to ensure oxygen supply for the first few minutes.

b) within a few minutes, the supply changes over to liquid oxygen which is pumped from a storage tank.

With the security of duplicate oxygen pumps, the additional investment in the back-up system is not necessary.

2.2 GASIFICATION CAPACITIES AND ROUTINE INSPECTION

A residual oil gasification plant should have sufficient reserve gasification capacity to allow routine inspection and regular maintenance of burner, quench ring, and refractory lining in a reactor without serious loss of the overall plant output and efficiency (see Figure 3 for general arrangement of a gasification reactor). Some examples of the capacities of gasifiers in Linde residual oil based plants is given in Table 1.

GASIFICATION CAPACITIES OF OPERATIONAL PLANTS BUILT BY LINDE

Plant	No. of gasifiers	Operating pressure bar a	Normal feed rate to one gasifier t/h residual oil	Max. design feed rate to one gasifier t/h residual oil
1	2	85	21	34
2	2	63	16	24
3	2	84	14	20

It has been shown that for heavy feed stocks, regular inspection of the gasifier internals at intervals of 3 to 4 month has benefits in limiting deterioration in the burner, quench ring, and refractory lining, and allows the performance of these items to be monitored.

2.3 GASIFIER CONTROL, START-UP AND SHUT-DOWN

To allow regular inspection and maintenance of the gasifier reactors, and to avoid any impact on overall plant operation, the control concept must allow smooth shut-down and start-up of a gasifier within the operating complex. Systems which are critical include:

oxygen supply, since this is common to more than one gasifier. When a gasifier is taken off line, the oxygen supply is instantaneously shut off, using two valves in series. An additional oxygen blow-off must take the excess oxygen with no disturbance to the running gasifiers. Equally important is the stability of oxygen supply in case of failure of a pump or compressor (see above notes on internal oxygen compression),

burner shut-off block valves. The oxygen block valves are quick-acting, bubble-tight ball valves. Special materials and the correct design of the seats and packing are necessary to achieve the required tightness for the cycles of operation,

purging of the burner inlet lines. When the gasifier is shut down, the burner inlet lines must be purged to avoid reaction of residual oxygen with combustible material within the burner. Nitrogen or steam can be used for purging. The successful implementation of this important safety aspect requires the correct detailed design of the piping system. Also, the duration and rate of the purge are important questions to ensure a satisfactory shut-down procedure.

water supply to the quench ring. When the gasifier pressure reduces following shut-down, the supply of water to the quench ring, and the drainage of excess water from the quench chamber, must be secured.

cooling water supply to the burner. The flow of cooling water to a cooling coil on the burner tip must be maintained during removal and refitting of the burner in the gasifier.

2.4 BURNER PLATFORM LAYOUT

The burner platform layout must consider both correct piping system design and adequate access for operation and maintenance. In close proximity to the burner at the top of the gasification reactor are the following critical piping systems.

high pressure oil supply with shut-off valve, recycle valve, steam injection point, isolation valve, etc.

high pressure oxygen supply with shut-off valves, blow-off valve, steam injection point, isolation valve, etc.

nitrogen purging system

burner cooling water supply and return.

In addition, the layout of the burner platform must allow safe access and working area for a number of operators who are involved in the shut-down, inspection, and start-up procedures. Room must also be allowed on the platform for a spare process burner, a preheat burner, and the services, tools and instruments necessary for inspection of the burner and refractory of a gasification reactor during the shut-down period.

In order to keep the shut-down period as short as possible, and also to ensure complete safety during the intense activities on the burner platform during this period, detailed working procedures must be worked out. The execution of these procedures must be fully considered in the development of the burner platform layout.

2.5 CONSTRUCTION MATERIALS

Correct materials of construction and corrosion protection measures are important for the life of the plant. Material selection must take into account the presence of trace components, such as NH3, H2S, HCN, HCOOH, as well as salts and suspended ash and carbon in the liquid phase. Realistic allowance must be made for changes in feed stock quality, when wide variations in the content of trace components can occur. Whereas carbon steels are generally adequate, there are some critical areas where special materials are required.

During the last decade, the feed stock quality processed by GNFC has changed considerably from the original values, and valuable information has been gained on the interaction of trace components present in the feed stock and their influence on plant operation. As an indication of the variations which have been experienced, the content of some key trace components in the feed stock gasified over the years can be compared (Table 2).

Table 2 VARIATIONS IN FEED STOCK QUALITY AT GNFC

	1983	1986	1989	1992	1995
Feed stock analysis					
Sulphur wt. %	3.0	0.9	1.1	0.5	1.2
Nickel wt. ppm	5	85	300	90	100
Vanadium wt. ppm	55	35	80	45	25
Sodium wt. ppm	15	10	8	55	70
Calcium wt. ppm	-	33	60		
Chloride wt. ppm	28	6	8	10	5

Particularly, there is an interaction of sulphur and nickel contents in the feed stock on the formation of nickel carbonyl in the raw gas. This compound is a vapour at the conditions in a gasification reactor, and can then lead to solid deposits in down stream equipment where conditions cause decomposition. The possibility of nickel carbonyl formation must be considered during the design phase, and suitable measures taken where necessary to protect down stream equipment.

2.6 PERFORMANCE AND AVAILABILITY RECORD

The production record of the GNFC gasification complex, given in Table 3, is demonstration of the correct engineering of the plant, and the high quality of the operating and maintenance personnel on this site.

Table 3 OPERATING RECORD OF GNFC

Year	Capacity Utilisation %	Days of nil production
1986-87	92.6	36
1987-88	92.8	40
1988-89	108.0	9
1989-90	106.8	17
1990-91	94.1	35
1991-92	114.2	10
1992-93	113.5	33
1993-94	106.2	26
1994-95	111.5	29
1995-96	116.9	17

Notes: Capacity Utilisation = Annual Production / (Name Plate Daily Capacity x 330)

The client is justifiably proud of this record, for which he has won wide recognition, including in 1990 the Fertilizer Association of India Award for the best performance of a nitrogen fertilizer complex. This was the first time that a plant based on gasification of residue won such an award, in competition with plants based on natural gas or naphtha using the conventional steam reforming process.

The engineering and construction of the Dalian 1000 MTD ammonia plant in China was performed with extensive co-operation between Linde and several Chinese partners. The establishment of sound working relationships between the partners gave rise to many challenges. The successful execution of the project demonstrates the commitment and ability of all the parties to meet these challenges, and provides a good basis for the performance of future large projects.

The start-up of the Dalian gasification plant was carried out on schedule and led to the performance test run in July 1996. The guaranteed performance and consumptions were met and operation at 70% plant load on one gasification reactor was demonstrated. The complex was accepted and handed over to the client on 4 August 1996.

The long-term operating results from this plant will add to the available experience on refinery residue gasification, and can be presented on a future occasion.

The experience gained in partial oxidation of residual oil in the chemical industry is directly applicable to the design of an IGCC complex. The similarity between the two fields can be appreciated from Table 4, which gives some overall information on the size and operating conditions of three Linde built gasification complexes.

Table 4 OVERALL CAPACITIES OF GASIFICATION PLANTS

Plant	1	2	3	**IGCC Unit 250 MW** Typical figures
Gasification Unit				
Residual oil feed rate	42 t/h	33 t/h	29 t/h	53 t/h
Operating pressure	85 bar	63 bar	64 bar	50 bar
Oxygen consumption	32 000 Nm^3/h	25 000 Nm^3/h	22 000 Nm^3/h	40 000 Nm^3/h
Air Separation Unit				
Oxygen capacity	35 000 Nm^3/h	28 000 Nm^3/h	24 000 Nm^3/h	42 000 Nm^3/h
Supply pressure	102 bar	75 bar	100 bar	62 bar
Nitrogen capacity	39 000 Nm^3/h		30 000 Nm^3/h	64 000 Nm^3/h
Supply pressure	76 bar		75 bar	22 bar
Acid Gas Removal Unit (Rectisol® Wash)				
Rate of Feed Gas	190 000 Nm^3/h	134 000 Nm^3/h	133 000 Nm^3/h	160 000 Nm^3/h
Recovered Sulphur Fraction	2 400 Nm^3/h	2 000 Nm^3/h	1 000 Nm^3/h	4 000 Nm^3/h
with H2S+COS concentration	57 5 mol %	74.4 mol %	12.8 mol %	45 mol %
Turbo Machinery Compressor rated powers and driver type(steam turbine or electric motor)				Gas turbine 170 MW Steam turbine 90 MW
Air compressor	20 MW(turbine)	15 MW(motor)	14 MW(turbine)	
N2 or O2 compressor	15 MW(turbine)	8 MW(motor)	8 MW(turbine)	14 MW / 7 MW (motors)
Syngas compressor	12 MW(turbine)		6 MW(turbine)	
Refrigeration compressor	9 MW(turbine)			

3. DOWNSTREAM TECHNOLOGIES IN GASIFICATION PLANTS

The three gasification complexes listed in Table 4- generate synthesis gas for products in the chemical and fertilizer industries. An IGCC complex can also be designed to produce synthesis gas for chemical by-products.

The proven processing routes and products include:

hydrogen by PSA purification or membranes
cold box separation unit for carbon monoxide
methanol synthesis
ammonia synthesis

For any of these products, optimised processing schemes can be applied, for which proprietary technology is available in the Company.

All the applications are proven in a wide range of capacities.

4. IGCC

The design of an IGCC complex can benefit from the experience gained of partial oxidation of residual oil in the chemical industry. The plant capacities and feed stocks processed in the reference plants listed above are also typical for IGCC.

Although IGCC integration schemes and process units differ from an integrated ammonia plant, the design and operating experience is relevant for both processing routes. In view of the importance of reliable and profitable operation for the successful financing of IGCC projects, it is of great benefit to the buyer that the following experience in the design of the Gas Island can be found in one Company:

design and contracting experience for Texaco Gasification plants for high capacity and a wide range of heavy residual oil feed stocks.

proprietary in-house technologies for related units, Air Separation, Gas Scrubbing, Sulphur Production.

design and operating experience of complete plants based on gasification of refinery residue.

supplementary downstream technologies for production of by-products to create additional financial benefits.

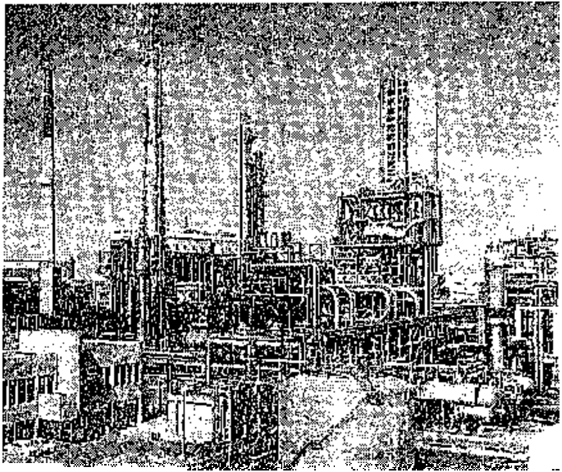
integrated methanol peak-shaving technology, to adjust to varying power output demands.

proven expertise in managing billion DM turn-key projects.

REFERENCES

[1] H. Jungfer "Synthesis Gas from Refiney Residues", Linde Reports on Science and Technology 40, 1985.

Gujarat Narmada Valley Fertilizer Co., India



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Feed gas:

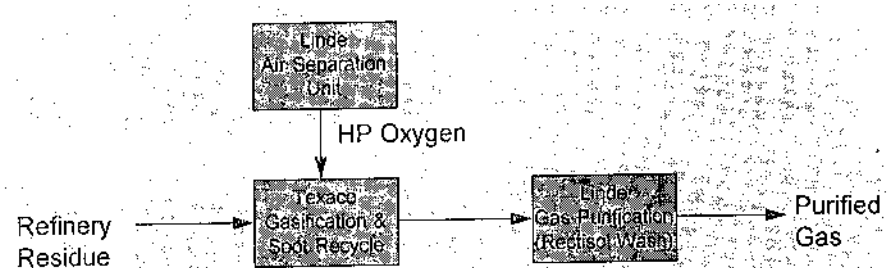
Heavy Oil and
Vacuum Residue

Products:

1.350 MTD Ammonia
1.800 MTD Urea
60 MTD Methanol

Gasification of Refinery Residue

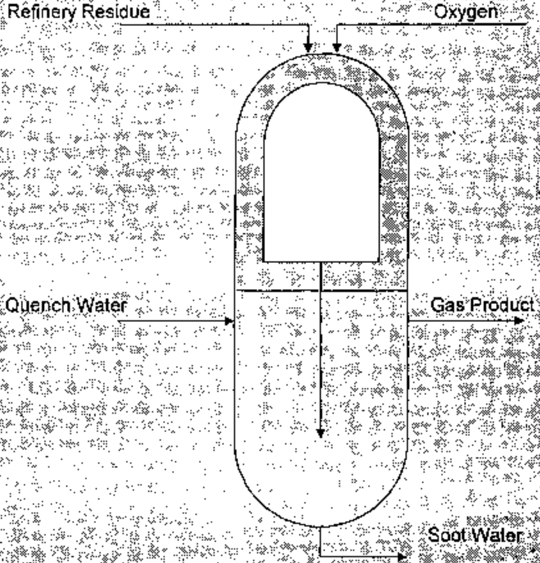
Integrated Gas Generation System



Texaco Gasification Reactor with Quench Cooler

Linde

Figure 3



Refinery Residue

Oxygen

Quench Water

Gas Product

Soot Water

Gasification in practice: the ISAB Energy 512 MW IGCC Complex

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Abstract

New European Union and Italian Environmental regulations which will limit the permissible sulphur content of fuels burned in equipment not fitted with flue gas desulphurisation units, are causing a rapid decline of the high sulphur (2.5% wt - 3% wt) and medium sulphur fuel oil (1% wt - 2% wt) market and the refiners must either make a major capital investment to change their product mix or make a better use of their heavy products.

This is particularly true for the ISAB Refinery which was originally designed to process heavy crude oil (a configuration which yields higher refinery margins) and consequently has a relatively high output of heavy products (28% of total output). Rather than invest in an extensive desulphurisation programme, ISAB has decided to implement a more brilliant and economic alternative of gasification of its oil residues and production of electricity for sale to ENEL.

The Italian Government has created a favourable environment for development of independent environmentally friendly power stations through tariff incentives during the first eight years of production. In 1992 ISAB undertook its first steps towards the development of the Integrated Gasification and Combined Cycle Project (hereinafter IGCC) and engaged Foster Wheeler Italiana to perform the initial conceptual design for the Project.

At the end of 1993 the Basic Design of the overall process and utility units of IGCC Complex was completed; at the same time a joint venture between Erg Petroli S.p.A. (participating at 51%) and Edison Mission Energy (an American Company belonging to Southern California Edison group, the second private electric Company in the U.S.A., participating at 49%), was created.

The IGCC Complex will be built in Priolo (Siracusa), close to ISAB Refinery, by SnamProgetti-FosterWheeler-Energy,under a lump-sum turnkey construction contract and under the formula of "Project Financing non recourse" (six International Banks, Barclay, Societe Generate, City Bank, DKB, IMI and MedioCredito Centrale, are involved). Construction activities have just started and will have a duration of about three years with a manpower involvement of about 2.000 people per day (as peak value); start-up of commercial operations is foreseen by the end of 1999.

IGCC Complex processes the asphalt obtained from ISAB Refinery Visbroken tar, after a vacuum distillation and subsequent treatment in a new ROSE deasphalting plant. Asphalt is converted into a synthesis gas, that after H2S purification, is sent to a Combined Cycle for a net electric power output of 512 MW.

1.0 Introduction

Erg Petroli S.p.A., 80% owned by Erg S.p.A. and 20% by Agip Petroli S.p.A., and Edison Mission Energy, which is a wholly owned subsidiary of Edison International, have formed, at the end of 1993 ISAB Energy S.r.l., for the purpose of developing, owning and operating a 512 MW Integrated Gasification and Combined Cycle (hereinafter IGCC) power plant located in Priolo, 8 Km from Siracusa, Sicily, adiacent to the ISAB Refinery.

Erg Petroli, that is partecipating to this project at 51%, is the largest Italian private oil refinery and marketing group; its core unit, ISAB Refinery commands a 13% share of the Italian oil refining capacity. Edison Mission Energy, at 49%, is one of the world's leading developers of independent power projects and owns interest in about 50 operating power generation facilities with an aggregate generating capacity of 6646 MW.

The project is driven by impending E.U. legislation designed to limit the permissible sulphur content in fuels and emissions. Italy government is therefore encouraging the development of independent, environmentally friendly power stations by subsiding the initial eight years of operation of all non-polluting generating plants.

One such candidate is power generation by gasification of refinery residues from which sulphur is removed in the gasification process, the resultant clean syngas being burnt in gas turbine. This concept qualifies as non-polluting and benefits from a subsidy according to CIP Resolution No. 6 of 29 April 1992.

2.0 General Overview

2.1 Trend in the Refinery Industry

In many countries most of the refineries have been equipped with severe processes, thermal (Coking) and catalytic (Hydrocaking, Catalytic Cracking, Dewaxing), aimed to transform the heavy hydrocarbons. These are processes capable of increasing the share of light products. Neverthless in Italy this renewal has been delayed, being preferred less severe thermal processes, like Visbreaking and thermal cracking.

These processes have in input the residue from the bottom of the Vacuum column and produce as output other distillates plus a residue, precisely named TAR. Its characteristics are the high viscosity, which hinders its displacement, and the content of polluting elements, like sulphur (it ranges between 3 and 6% depending upon the quality of the processed crude) and various metals (mainly nickel, vanadium, iron).

In Italy the choice to limit investments required to face the evolution of the demand and to upgrade the refining plant, was practicable thanks to the availability of the National Electricity Board (ENEL) to buy and to burn the oil with high sulphur content obtained from such residues.
Tar represents a considerable part of the process crude, about 12-18 wt% as average figure, depending on its quality, and has a calorific value around 9400 Kcal/Kg.

When we consider the global capacity of the Italian industry, we can estimate the considerable amount of 7.5 million tonnes of Tar produced per each year.

In the past, and nowadays as well, it was mixed with lighter products with very low sulphur content to generate an output with high sulphur content (1-3%) and good calorific value, which is still burnt in the thermal power stations.
From 1997 the environmental regulation will not allow products with sulphur content over 0.25-0.30% to be burnt in power stations. This fact forces to find out a new way to work off the residue, or better, to contrive an use economically profitable.

The existing refinery plant has substantially three alternatives to conform itself to the new environmental regulations and need of the market demand at the same time:

A. Deep Conversion: it is a deep re-design of the refinery cycle. More severe manufacturing activities based on catalytic processes are installed in order to extract a larger quantity of light products. In this case the whole plant has to be re-designed and the cost is very high;

B. Tar Desulphurisation: the second possibility is to build a desulphurisation plant in order to clean the residuals and make them adapted to the requirements of the electricity industry. This is too a quite costly alternative as these processes involve a complex technology;

C. Tar Gasification: the last available chance consists in the gasification of the residuals. It yields a fuel gas suitable to feed gas turbines and therefore to generate electricity in a combined cycle. Whenever the electric power generated may a have a good economic value, this choice results the most interesting alternative and the investment in the gasification plant can be recovered in a reasonably short time.

2.2 The Italian Electricity Supply Industry

Italy is extremely poor in primary energy sources. A large amount of produced energy comes from imported fuels, our country is experiencing an increasing difficulty to meet the demand for electricity in compliance with the environmental regulation.
The market for high and medium sulphur fuel oil (3.5% - 2% wt) is declining rapidly and refinery must either make a major capital investment to change their product mix or make a better use of their heavy products.
This is particular true for ISAB Refinery which was originally designed to process heavy crude oil. Different alternatives were evaluated: at the end, rather than invest in an extensive desulphurization programme, it was decided to implement a more economic alternative of gasification of its oil residues and production of electricity for sale to ENEL.

The choice of the Italian oil industry to produce electricity to be sold to ENEL has a justification in the law in force since 1991 which allots a price really attractive for the electricity produced with IGCC power stations.
The IGCC technology complies with two fundamental needs of the national energy industry: the disposal of the refinery residual with high sulphur content and the construction of new capacity for the electricity generation.

3.0 Plant Description

Feedstock (asphalt, visbroken tar, heavy fuel oil, etc.) is introduced into the gasifiers which are operated at a temperature above 1400 °C and a pressure of about 70 barg.
Gasification takes place in the presence of oxygen and high pressure steam acting as a temperature moderator. The syngas and solids (consisting of unconverted carbon and ash) are quenched by water sprays before they exit the gasifier. The solids are trapped in the quench water, leaving the syngas clean and relatively cool. Unconverted carbon is recycled back to the gasifier to achieve 100% carbon conversion, using naphtha as the soot carrier. Quench grey water from the gasification is treated and filtered to recover a metal cake which contains large amount of nickel and vanadium. The metal cake is intended to be sold to metal reclaimers. Thereafter the process waste water, after a further pre-treatment for ammonia removal, is sent to a municipal treatment facility.

The particulate free raw syngas is then further cooled through heat exchangers, generating medium pressure steam which is utilized to generate electric power in the steam turbines. The sulphurous compounds in the syngas (acid gas) are removed so that the sulphur emissions in the form of SO, are minimized when the syngas is burned in the combustion gas turbine. From the acid gas removal unit, the sulphur offgas is sent to a Claus sulphur unit to convert 99.8% of sulphur offgas to elemental sulphur suitable for by-product commercial sale. The virtually sulphur-free, high pressure syngas is then sent to the gas turbines, via a gas expander that recovers the high delta-pressure of syngas through 10 MW electric power and a humidifier which saturates the syngas in order to avoid high NO_x formation in the combustion chamber of gas turbines. The exhaust gas from the two gas trrbines is ducted to heat recovery steam generators which produce high pressure steam used: (i) to drive the steam turbine generators to produce additional power; and (ii) as a temperature moderator in the gasifiers. Additional syngas is combusted in supplemental burners in the heat recovery steam generators to produce additional steam for the steam turbines.

4.0 Technologies adopted

4.1 Texaco Gasification Process

The main licensor of the IGCC Complex is Texaco that designed gasification section. ISAB Energy gasification will be the first in the world to operate with asphalt. Texaco has tested heavier feeds in Montebello Gasification Pilot Plant.
Two gasifiers convert asphalt produced from the visbroken tar of refinery, through a Solvent deasphalting plant of Kellogg Technology, into a fuel gas for the Combined Cycle. Syngas produced from gasification reaction, at a temperature above 1400 °C , is cooled down to 250 °C using a water stream injection to control the reaction avoiding the oxidation of CO into $C0_2$.

4.2 Heavy Metal Recovery

The grey water from gasification section is processed to remove sulphides, cyanides, thiocyanates and suspended solids including all the heavy metals. Chemicals injection (caustic soda, polyelectrolite, ferrous sulphate, etc.) will favour the precipitation of dissolved salts of heavy metals and the other compounds mentioned and concetrate the precipitate in a filter press to yield a "cake" containing approximately up to 28% vanadium and up to 9% nickel which are intended to be sold for heavy metals recovery.

4.3 Acid Gas Removal

This is a conventional technology employed in the refining, petrochemical and natural gas industries. The acid gas removal system utilises a solution of Methyl-Di-ethanolamine (MDEA) solvent solution to wash the gas and concetrating H_2S gas routed to Sulphur Recovery system. The process operates at relatively low temperature to improve selectivity and avoid co-absorption of $C0_2$ into R_2S gas.

The washing absorbs virtually all the H_2S and a very low amount of $C0_2$ yielding a purified fuel gas of less than 15 ppm total H_2S + COS and less than 10 ppm total HCN and NH_3 by volume.

4.4 Sulphur Recovery

The sulphur recovery process is based on the Claus reaction in which H_2S and SO_2 react to form elemental sulphur. Most of the Claus plants use air for this process; however, many of the units installed over the last few years have determined it is more economical to use oxygen, especially if an oxygen plant is being installed for another purpose as is the case for ISAB Energy Project.

About 50% of conversion to sulphur takes place in the combustion chamber; the gas leaving the chamber is cooled to condense sulphur vapour. The heat removed from the gas is used to generate medium pressure steam. The remaining conversion of the sulphur gases to elemental sulphur occurs in two stages of catalytic reactors.

The liquid sulphur produced will be sent in an off-site solidification plant to produce about 190 tonnes per day of sulphur in pellets to be sold to external companies.

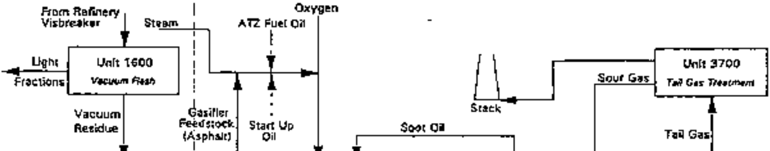
4.5 Combined Cycle

The power plant design comprises two combined cycle gas turbine modules. Each module consists of one Siemens type V94.2 gas turbine modified for use with low calorific syngas from the gasification plant, one supplementary Heat Recovery Steam Generators (HRSG) and one re-heat condensing steam turbine generator unit. The proposed rating for the gas turbine is approximately 161 MW per unit and the proposed rating for the steam turbine unit is approximately 95 MW per unit. An additional 10 MW is generated in the gas expander.

4.6 Auxiliary Systems

The IGCC Complex will be equipped with all the necessary auxiliary systems including cooling water (a closed circuit with about 60.000 m3/h of circulating sea water), demi water, desalinated water (through two MED Desalination Units with a capacity of 300 m3/h each), air, fuel-oil and fuel-gas utilities, flare and blow-down system, electrical distribution, firefighting system, buildings, etc.

Herebelow a schematic block diagram is shown.



ISAB ENERGY IGCC POWER STATION

5.0 Plant Performance

The perfomance of the IGCC Complex in converting heavy oil into electric energy is is dependent on the level of load, the ambient air temperature and, to **a** lesser extent, on the characteristics of the feed. Herebelow the normal operating case performance are reported:

Reference Conditions		
Air Temperature (°C):	18	
Gasification feed:	Asphalt	
Flowrate (t/h):	132	
Electric Energy Production		
N. 2 Gas Turbines	322,4	MW
N.2 Steam Turbines	230,4	MW
Expander	9,8	MW
Gross Power Output	562,6	MW
CCU Electrical Internal Cons.	11,8	MW
Net Power Output	550,8	MW
Process & Utility Units Cons.	30	MW
Net IGCC Power	520,8	MW

6.0 Environmental

6.1 Solid Effluents

6.1.1 Filter cake Production and Management

Filter cake is a by-product of the gasification process produced in the heavy metals recovery unit of the IGCC Complex. The composition of the filter cake, in particular the vanadium and nickel content, makes it a marketable product to the metallurgical industry. According to the current Italian legislation, the filter cake will be classified as a non-toxic hazardous material which will require special handling. The filter cake production rate will be approximately 7.5 tonnes per day (as dry material) for normal operation with asphalt.
ISAB Energy is currently developing a strategy for the management of the filter cake and is finalizing preliminary contracts with different potential reclaimers.

Taking into account the world-wide Vanadium production (about 25000 tonnes per year), ISAB Energy production will represent about 2.5 % wt.

6.1.2 Sulphur Production and Sale

Elemental sulphur is a by-product of the gasification process produced during the syngas clean-up. The sulphur is of a commercial grade suitable for sale to agricultural and chemical industry market. The sulphur production rate is approximately 60.000 tonnes per year.

6.2 Liquid Effluents

The sanitary and pre-treated process water coming from IGCC will be routed to a Municipal Treatment Plant; the blow-down from cooling tower system will be discharged to the sea after inspection in a suitable pit.

6.3 Gaseous Effluents

Due to ecological plants provided in IGCC Complex and to the final treatment of the flue gas in a SCR system (Selective Catalytic Reduction) located in the HRSG of the Combined Cycle the emissions are largely below the current environmental regulations.

7.0 Main Commercial Agreements

7.1 Feedstock Supply Agreement

Erg Petroli and ISAB Energy have entered into a 20 years feedstock supply agreement. Feedstock means principally asphalt but also, in case of unavailability of asphalt, other types of feedstocks either sourced from the Refinery (such as visbroken tar, vacuum residues, fuel oil) or purchased from third parties.

Erg Petroli's primary obligation is to supply asphalt, provided that where asphalt is not available for any reason, Erg Petroli will supply an equivalent alternative feedstock.

7.2 Power Purchase Agreement

The terms of the sale of electrical energy to ENEL are set forth by two agreements with ISAB Energy. These agreements, known as the Convenzione Preliminare ("Preliminary Agreement") and the Accordo Integrativo ("Integrative Agreement"), together with a series of laws and decrees which they reference, provide all the key legal and commercial terms of the agreement.

The price settlement was defined by CIP Resolution No. 6 of 29 April 1992 which sets out the tariff arrangements to be applied to independent producers of electricity.

ISAB Energy will receive a price according to CEP 92 resolution.

7.3 Oxvgen Supply Agreement

The gasification process requires large quantities of oxygen and nitrogen. Ail Liquide will supply on a long term basis "over the fence" to the Project. The oxygen plant will be located on a plot of land adjacent to the Project Site. The oxygen Supplier will guarantee a supply of more than 100.000 cubic meters per hour of 95% pure oxygen.

7.4 Operating and Maintenance Agreement

The Sponsors propose to create a Company named ISAB Energy Services who will operate and maintain the Project. The shareholders will be Erg Petroli (51%) and Edison Mission Energy (49%).

7.5 IAS Agreement

Industria Acqua Siciliana is a local service company managing waste water discharges in the area and treating them in an existing waste treatment plant located in Priolo. IAS is a public/private consortium whose shareholders are at the same time users of the plant. All private shareholders/users bear the costs of the plant on a basis proportionate to their usage. Upon successful completion of negotiations with IAS, ISAB Energy has become a shareholder of IAS and will discharge to the Plant a maximum of 170 m^3/h of pre-treated water.

7.6 SNAM Metano Agreement

In order to cope with the fuel gas consumption of IGCC Complex, normally utilized for pilots of flare and hot-oil and Claus Units furnaces pilots, an agreement for the supply of natural gas was defined with SNAM Metano Company.

8.0 Financial Overview

The ISAB Energy IGCC Plant is financed through a limited recourse project financing scheme. All the tasks related to the successful completion of the project financing development have been completed in July 1996 (Financial Close). This result is particularly significant because this is the first time that an IGCC Plant is completely financed without using government funds. This has been a result of intensive activities aimed from one side to finalize main contracts with ENEL, Oxygen Supplier (Air Liquide), Feedstock Supplier (Erg Petroli), and other minor Parties and from the other side to demonstrate to the Technical Advisor (Stone & Webster) the robustness and well proven experience of the technologies involved.The Total Project Capital Costs amount to Lire 1.900 billions. Funds made available from ISAB Energy's financing arrangements will be expended over a construction period of approximately 3.5 years and repaid over a period of approximately 8.5 years. At the earlier of commercial operations the Sponsors will infuse equity in the amount of 25% of actual Project Costs.

9.0 Project Status and Schedule

The Project will be built under a lump-sum turnkey construction contract (LSTK) which will incorporate terms already agreed between ISAB Energy and the Contractor pursuant to a Memorandum of Understanding signed in March 1995. The Contractor, Consorzio SnamProgetti - Foster Wheeler - Energy, was selected after completion of a competitive international bidding process. After a six months Open Book Phase (June-December 1995) and the definition and signing of a LSTK Contract including engineering, construction and taking over of the Plant, a detailed engineering has been developing from Febraury 1996. On July 17, 1996 the notice to proceed and the official start-up of Construction activities took place. Soil consolidation works and earth movements started in September and are currently in progress. At the current date (early December 1996), engineering development was at 30% of the schedule whereas material purchase orders at 5%. Completion of engineering is foreseen by October 1997, whereas the start-up of commercial operation is scheduled by the end of 1999.

References

1. J.C. HOLDSWORTH, G.D. TOBIN: "Processing Options for Heavy Residues"; Foster Wheeler Publ., 1993.

2. A. ARIOLDI, G. BERGAMASCHI: "La gassificazione del tar in Raffineria"; La Rivista dei Combustibili, Vol. XVL, fasc. 10, 10 ottobre 1991.

3. F. BIFULCO: "La gassificazione dei residui petroliferi pesanti associata al ciclo combinato per la produzione di energia elettrica e ambiente"; Atti del Convegno ATI 1992.

4. G.BELLINA, G.CAMMARATA, M.CIANCIO, A.FICHERA, L.MARLETTA; "First and second law analysis of an IGCC with Tar Gasification"; Atti del Convegno ATI 1996 "Energy and Environment towards the Year 2000".

5. EEFE; "Assessment of the External Costs of Integrated TAR Gasification Combined Cycle"; April 1995.

Operating Experience with Air Blown 200 ton/day Entrained Flow Coal Gasifier

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Abstract:

The New Energy and Industrial Technology Development Organization (NEDO), the Engineering Research Association of Integrated Coal Gasification Combined Cycle (IGC Research Association) newly succeeded in the development of an air blown two stage entrained bed coal gasification technology for power generation with the achievement of one month stable and continuous operation at the 200 ton/day IGCC pilot plant. Based on these results, newly developed Mitsubishi IGCC system "HERCULES" composed of the air blown two stage entrained bed coal gasifier, the fix bed hot and dry gas clean-up and the 1300-1500℃ class gas turbine enables the highest net thermal plant efficiency. "HERCULES" attains high reliability and economic feasibility of generation cost equivalent to that of conventional pulverized coal fired power plant with the simplicity of the system under the Mitsubishi's single responsibility.

1. Introduction

The future fossil-fueled power plants are strongly required to achieve superb environmental performance with excellent economy from the standpoints on the global environment preservation. Furthermore, coal is the most important and dominant fuel on the viewpoint of the energy conservation on global scale and the security of energy resources, because coal is overwhelmingly abundant and distributed in many countries in comparison with the other fossil fuels (oil and natural gas). Therefore the development of new technologies for coal firing power generation has being accelerated worldwide.

Integrated Coal Gasification Combined Cycle (IGCC) is regarded as the most powerful candidate for the new coal firing power plant in the next generation because of the following two points. One is its significantly high net thermal efficiency of 43-50+% which can be achieved with the increasing of gas turbine inlet temperature. The other is its superior environmental performance which is attained by means of the conversion process from "dirty" coal to "clean" purified gas and the ash disposal of glassy slag.

Under these circumstances, several IGCC demonstration plants are now under construction or in demonstration operation in Europe and USA.

In Japan, as the national R&D project, a 200 ton/day entrained bed coal gasification combined cycle pilot plant was constructed and is now in the final stage of the test operation since 1991 by the IGC Research Association under the entrust of NEDO. MHI is in charge of the development of the pressurized air blown two stage entrained bed coal gasifier and the fixed bed dry gas clean-up system in this project. These essential technologies for IGCC have almost verified at this pilot plant with the achievement of one month stable and continuous operation in March 1995.

This air blown entrained bed gasification concept is more suitable for power generation compared with the oxygen blown ones which had been developed mainly for chemical plants in Europe and USA, because it is not necessary to use a large amount of the auxiliary power to produce oxygen . However, the air blown entrained bed gasification has difficult points in

nature and had not been succeeded until our first success in April, 1995. The key point is to balance the calorific value of product gas fuel and the combustion temperature.
The status and future perspective of the air blown entrained bed gasification power generation technology are introduced in the followings.

2. Particulars of dry feed air blown two stage entrained bed coal gasifier

The air blown entrained bed gasification combined cycle is able to attain high net thermal efficiency with dry coal feed and dry gas clean-up because of a great merit of no necessity of oxygen generation by the air separation unit and no latent heat loss in the gasifier and gas clean-up system compared with the oxygen blown IGCC system especially with slurry coal feed and wet gas clean-up. (See Table 1.) However, the nitrogen in the air as the gasification agent lowers the combustion gas temperature in the gasifier ,so that it needs special attention to the ensured discharge of melting ash and sufficiently high coal gas calorie for the stable combustion at gas turbine.

As the effective solution of this point, MHI adopted two stages gasification concept, which is consisted of combustor and reductor. (See Figure 1.) This concept realizes the ideal process based on the function required. The combustor, where coal and recycled char are burnt at high temperature under high air/coal ratio condition, supplies sufficient heat to the reductor for gasification while attaining few unburned carbon in ash and smooth exhaust of molten slag. The reductor ,where char is gasified through heat absorption process, attains high gasification efficiency under low air/coal ratio condition using the heat from the combustor.

This technology had been developed with the 2 ton/day PDU (Process Development Unit) gasifier by MHI and Central Research Institute of Electric Power Industries (CRIEPI) from 1981 to 1991 and was adopted to the gasifier of the 200 ton/day pilot project.

This air blown two stage gasifier also has an advantage that the ash particles in coal gas are quenched at the reductor from molten state to the solid state by means of coal gasification absorption reaction without special quenching medium or radiant cooler, which enables fairly small size of the gasifier, regardless of the fact that its coal gas flow is roughly double of the oxygen blown ones. Thus It is considered to be a great advantage compared with the oxygen blown one stage entrained bed gasifiers which need a large size radiant cooler and/or a quenching system to mix cold recycled coal gas. The radiant cooler requires at least the ten(10) times resident time of the air blown gasification reaction time. The gas flow after quenching by cold gas mixing at the oxygen blown gasifier becomes nearly double. Therefore, there is no big difference of the size among the air blown gasifier and the oxygen blown ones.

3. Result of the 200 ton/day air blown entrained bed coal gasifier operation

3.1 Outline of the 200 ton/day pilot plant

The research and development project of the 200 ton/day pilot plant for an entrained bed coal gasification combined cycle is sponsored by the Japanese government (MITI/NEDO) and all the Japanese utility power companies. Under the entrust by the NEDO, the project is directed by the IGC Research Association which is a technical association composed of the members from ten domestic utility power companies and CRIEPI. The pilot plant is located at Nakoso power station of Joban Joint Power Co., at Iwaki City, Fukushima Prefecture and has been operated by the IGC Research Association since 1991.

The system schematic diagram of the pilot plant is shown in Figure 2. This plant, mainly consists of the air blown gasifier (gasifying 200 ton/day of coal), the hot and dry gas clean-up unit and the low calorie coal gas fired gas turbine (12.5MWe of capacity), is able to be tested and verified the performance, operational flexibility, reliability, and durability of air blown IGCC technologies. The dry gas clean-up is composed of two parallel systems, one is granular bed gas filter and fluidized bed desulfurization system, and the other is candle ceramic filter and fixed bed honeycomb desulfurization system. A product coal gas incinerator and its flue gas deSOx/deNOx facility have been installed to allow an independent gasifier operation. Heat recovery steam generator and steam turbine have not been installed, since they are well-established technologies. MHI is in charge of the design and manufacturing of air blown 200 ton/day gasifier ,the fixed bed hot and dry gas clean-up unit and the testing stand for the full pressure and full scale of a gas turbine combustor. The MHI's fixed bed hot and dry gas clean-up unit is composed of the dust removal system of a porous ceramic filter and honeycomb shaped stationary oxidized iron desulfurization unit. (See the Figure 3 and 4) This system realizes a significantly compact size and a completely water free system and superb clean-up performance under the high gas temperature condition of 350~450"C, which enables an effective use of the sensitive heat of the coal gas as the gas turbine heat input. And without moving materials, handling system including valves, pipes and controls are greatly simplified, which increased system reliability and economy.

The key dates of this project is shown in Table 2. The accumulative operating hours is 4770 hours of the gasifier, 1586 hours of the fixed bed clean-up and 1643 hours of the power generation respectively, including the one month continuous operation of 789 hours. The target of this project is to develop the air blown IGCC with pure domestic technologies in cooperation with the Government and the power companies. It has successfully accomplished with the satisfactory results. The test at the pilot plant was completed by the end of march 1996. After the test operation, this project was finally finished with the evaluation of the materials of the dismantled parts.

3.2 Outline of the operating experience of 200 ton/day gasifier

The test operation of the 200 ton/day coal gasifier has conducted since June 1991 using a domestic coal (Taiheiyo coal) and an Australian coals (Moura coal and Workwarth coal) of which properties are shown in Table 3. Some initial troubles were encountered and the proper improvements of the facilities and the operating procedures were made. After these initial adjustment, the remaining issue was the slagging problem which is slag built-up and at the wall of the lower part of the reductor and occasional plugging of slag tap at the bottom of the combustor.

The IGC Research Association investigated the cause of slag problem and made extensive analytical studies at MHI's Nagasaki laboratory. The IGC Research Association conferred with NEDO and the modification of the gasifier was conducted from July to October 1994. After the modification, neither slag built-up nor slag tap plugging occurred. And it is concluded that slagging trouble have been completely solved and the principle of the quenching effect with coal gasification absorption reaction at the second stage is fully verified. With this success, the reliability of the gasifier was highly enhanced and obtained the superior performance data to the target value as shown in Table 4.

3.3 Result of the One Month Reliability Test of the 200 ton/day Gasifier

The 200 ton/day gasifier had achieved the one month continuous operation for 789 hours (about 33 days) from 3rd March to 5th April 1995. During this test period, the operating of the gasifier had been continuously stable, and the test had completed according to the planned schedule. Figure 5 shows the major operating parameters of the gasifier in this test.

It is confirmed that the gasifier can maintain sufficiently high heating value of the produced coal gas for the stable combustion at the gas turbine and also keep stable discharging of molten slag through the slag hole with 100% recycling of char. Throughout this test, there was no trouble of any other auxiliary equipment, so that the reliability of whole system of the gasifier was confirmed.

The superior air blown gasification performance data have been collected. 70% of cold gas efficiency, 99.9% of carbon conversion efficiency are obtained, which satisfies to attain the highest net thermal plant efficiency at a demonstration plant. All of the ash in the coal fed to the gasifier is converted to non-leaching and compact glassy slag with negligible small unreacted carbon, which enables to realize the simple ash handling and the economical ash disposal without the fly ash including unreacted carbon.

3.4 Result of the One Month Reliability Test of the 20 ton/day Fixed Bed

Gas Clean-up

For the purpose of verifying the performance, reliability and durability of the MHI's fixed bed dry gas clean-up technology, a 20 ton/day capacity pilot facility was installed next to the 200 ton/day gasifier.

This pilot facility had also achieved for 765 hours continuous operation during the same period of the long term reliability test operation of the 200 ton/day gasifier. This test shows the excellent performance i.e., 1~5 mg/Nm3 of dust concentration and 20~50ppm of sulfur content (H2S.COS) at the outlet of the clean-up and the durability of the ceramic filter element, the oxidized iron desulfurization agent and the whole system.(See Figure 6.)

In this operation, the sulfur recovery system had been tested. However considering the actual market situation and infrastructure, the gypsum recovery system have been tested. MHI's gypsum recovery system is a pressurized system, which enables to produce *a* -phase gypsum and is extremely low heat loss due to its compact size.

Another advantage of this fixed bed clean-up equipment is easiness to scale-up its capacity because of the simple scale up criteria that the only quantity of the element is increased in accordance with the gas flow.

4. Mitsubishi IGCC system (HERCULES)

Based on the successful results of the 200 ton/day pilot plant, the development at the pilot phase of the air blown two stage entrained bed coal gasification and the fixed bed hot and dry gas clean-up have nearly completed, while obtaining the base data for scaling-up to a 2000 ton/day class demonstration plant.

IGCC may be more complicated system compared to conventional fossil plant, so system integration and simplification is very important. MHI can supply all the key components of IGCC i.e., gasifier, gas clean up system and combined cycle plant components such as gas turbine, steam turbine, etc. Therefore, as the system engineering and component manufacturer, MHI can drastically simplify the system based on the optimized integration among the interdependent main components under the single responsibility. This enables to realize high reliability, superior operational flexibility, simple maintenance and low installation cost.

Now, MHI propose the Mitsubishi's totally coordinated IGCC system, which is named "HERCULES" (High Efficiency and Reliability Coal UtiLizaton system with Economy and Simplicity). The feasibility study on the demonstration plant and commercial plant of "HERCULES" has been executed, which predicts its superb performance and economical feasibility.

The HERCULES system is composed of the air blown two stage gasifier , the fixed bed dry gas clean-up unit and the 1300℃ or 1500℃ class gas turbine, Table 5 and 6 show the major technical specification and Figure 7 shows the system diagram of HERCULES.

Adopting the fixed bed gas clean-up, which is a very compact in size and water-free system, contributes to the simplicity of the whole plant system and the significant reduction of the plant area. Furthermore, the dual-fueled gas turbine combustor eliminates the starting-up air compressor and a starting-up coal gas incinerator with deNOx/deSOx equipment.

The small sized air separation unit (ASU) is installed in order to produce nitrogen as the inert gas for the pressurized coal/char handling and the seal gas at the gasifier. The remaining oxygen at ASU is mixed with the air to the gasifier, so that the oxygen concentration in the gasifier will be somewhat higher than 21 vol.%.

The predicted performance data indicated in Table 5 are based on the condenser vacuum pressure of 722 mmHg. and the ISO condition of gas turbine performance (15 ^{0}C of atmospheric temperature and 60% of air humidity.) The design coal is an Australian bituminous coal, Moura coal.

The gross output is 323~493 MW and the capacity of the gasifier is 1800~2800 ton/day.

The net thermal efficiency of HERCULES is predicted to be 47.5+ LHV% It achieves the 20% reduction of the C02 gas emission compared with the conventional pulverized coal firing power plant in Japan , which enjoys the superb thermal efficiency with the extremely high steam condition of 246 bar and 566/566°C. Auxiliary power consumption rate is estimated to be less than 7~8%, which is considerably low as compared with 10~15% of the oxygen blown system.

The SOx/NOx emission at the stack inlet is anticipated to be 20 ppm with the advanced desulfurization technology and the new deNOx sy'stem. The dust concentration is predicted to be less than 10 mg/Nm3 at the inlet of the stack.

Ash is disposed in the form of glassy slag, of which advantages are as follows :

The volume of slag is half of the fly ash discharged at the conventional coal fired boiler with an electrostatic precipitator, which affect the size of the ash disposal area.

Non-leaching characteristics attains the simple treatment of drainage water and economical ash disposal.

Minimum carbon loss enhances the thermal plant efficiency and the commercial value of slag.

Moreover, in HERCULES, high quality sulfur or *a* -gypsum can be obtained as the salable byproducts at a gas cleani-up unit, which can meet the various requirements of the market.

MHI is also ready to supply the large capacity coal gas fired gas turbine. We have the proven technologies of the 1350°C class natural gas fired gas turbine, 701F/501F, and the 1250 °C class blast furnace gas (BFG gas) fired gas turbine. BFG gas has the similar property of coal gas. Its calorie is 970 kcal HHV/Nm3 (710 kcal LHV/kg), which is nearly same as that of coal gas. MHI had been performed the coal gas firing test with the full pressure and full scale test stand of a gas turbine combustor at the 200 ton/day pilot plant. Its capacity corresponds to one (1) can of the gas turbine for utility use, 701F/501F. This test shows the stable combustion and the low NOx conversion rate of 30~40 % (at 1000ppm of NH3), which is required at the combustion of the ammonia rich coal gas. Furthermore the 1500°C class natural gas fired gas turbine of 701G/501G has been developed. The commercial unit will start operation in 1997~1999.

The gasifier and the gas clean-up unit can be arranged in single steel structure. The required plant area per power output is reduced by 20% in comparison with the conventional coal fired power plant including the pollution control facilities i.e., the electrostatic precipitator and the deSOx/deNOx system.

The capital cost estimation of HERCULES executed at the feasibility study predicts 120% at the demonstration phase and 95% at the commercial phase compared to the conventional coal fired power plant. Consequently HERCULES has economical feasibility on the generation cost equivalent to that of the existing coal fired power plant.

5. Conclusion

The successful results of the R&D at the 200 ton/day pilot plant verified the IGCC technologies of the air blown two stage coal gasifier and the fixed bed gas clean-up. The development of air blown IGCC has stepped up to the demonstration phase. The Mitsubishi HERCULES composed of these technologies will realize the highest net thermal efficiency in the world and superb environmental performance with a feasible economic condition.

Reference :

1. Ishikawa, T. Hamamatsu, H. Moritsuka, T. Toda ,S. Ishigami .T.Furuya : "Current Status of IGCC Development in Japan", VGB kraftwerkstechnik Vol. 72 No. 5 1992 p.405
2. Toyoda, O. Shinada, S. Kaneko, K. Tokuda : "Experience of 200 ton/day Coal Gasifier Operation", JSME/ASME International Conference on Power Generation '93 Vol. 2 1993 p. 538
3. Takahata : "Development of the 200 ton/day entrained coal gasification pilot plant", Proc. 4th The Japan Institute of Energy Congress No. 7 1995 p. 180

Type	Gasifying Agent		Coal Feed		Gas Clean-up	
	Air	Oxygen	Dry	Wet	Dry	Wet
NEDO/IGC/ANC	○		○		○	
Oxygen-Blown Type A		○	○			○
Oxygen-Blown Type B		○		○		○

Table 1 Comparison of IGCC Systems

■ Key Dates of the Project

Start of Erection	Jan. 27, 1988
First Ignition of Gasifier	Mar. 3, 1991
Start of Gasification	Jun. 25, 1991
Completion of Test	Mar. 1996

■ Accumulative Total Operating Hours

Gasifier	4770 hrs
Fixed Bed Dry Gas Clean-up System	1586 hrs
Power Generation	1643 hrs
Longest Continuous Operating Hours	789 hrs (Mar. 3–Apr. 5, 1995)

Table 2 Key Dates and Operating Hours of 200 ton/day Pilot Plant

Item	Unit	Taiheiyo	Moura	Warkworth
Higher Heating Value	kcal/kg	6,550	7,160	7,080
Proximate Analysis				
Moisture	%	5.3	3.7	3.1
Fixed Carbon		35.8	56.2	53.3
Volatile Matter		46.7	30.9	33.2
Ash		12.1	8.7	10.4
Fuel Ratio	–	0.76	1.82	1.61
Ultimate Analysis				
Carbon	%	68.2	78.3	74.4
Hydrogen		5.71	4.60	4.84
Nitrogen		0.98	1.54	1.56
Oxygen		12.28	7.31	8.11
Sulfur		0.18	0.46	0.41
Ash Fusion Temperature				
Initial Deformation	°C	1,250	1,350	1,480
Softening		1,315	1,410	>1,500
Hemispherical		1,325	1,415	>1,500
Fluid		1,360	1,440	>1,500

Table 3 Property of Test Coals

	Unit	Predicted	Result
Kind of Coal	—	Moura	Moura
HHV	kcal/m³N	1,000	1,000~1,100
Cold Gas Efficiency	%	>67	67~74
Carbon Conv. Efficiency	%	>97	99.9
Char Recycle	%	100	100

Table 4 Performance Data of 200 ton/day Gasifier

Cycle	Hz	50	60
Gasifier	—	Air Blown Two Stage Entrained Flow Type	
Capacity	t/d	2600	1800
Dust Removal	-	Ceramic Porous Filter	
Desulfurization	-	Fixed Bed Type	
Gas Turbine	-	701F (1350°C class)	501F (1350°C class)
Gross Output	MW	493	323
Net Output	MW	444	290
Net Efficiency	%LHV	47.5	47.5
SOx/NOx Emission	ppm	20/20	20/20
Dust Emission	mg/Nm³	10	10

Table 5 Major Specification of HERCULES

Type	Pressurized Two Stage Partially Oxidized Entrained Bed Gasification
Gasification Agent	Air
Coal Feed System / Feed Gas	Dry Lock Hopper / Nitrogen
Char Feed System / Feed Gas	Dry Lock Hopper / Nitrogen
Furnace Structure	Pressure Vessel & Water Wall Furnace
Gas Cooling System	Reductor Coal Gasification Reaction
Syngas Cooler Type	Water Tube
Slag Disposal System	Wet Lock Hopper
Char Recovery System	Cyclone / Ceramic Filter
Char Recycle System	Direct Feed to Combustor

Table 6 Major Specification of 2000 ton/day Demonstration Gasifier

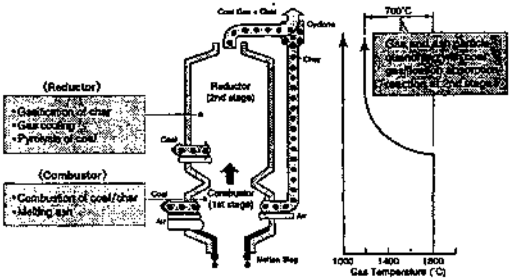


Figure 1 Principle of Air Blown Two Stage Entrained Bed Gasifier

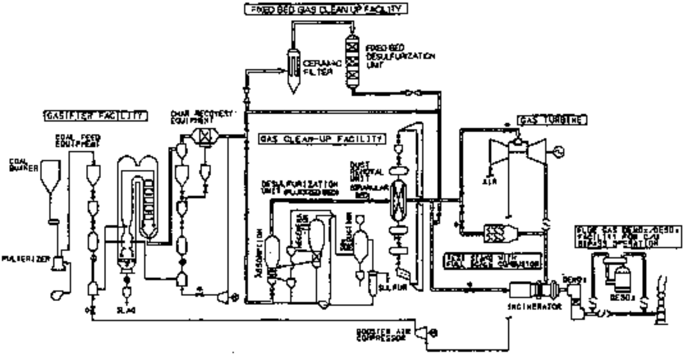


Figure 2 200 ton/day Pilot Plant System Flow Diagram

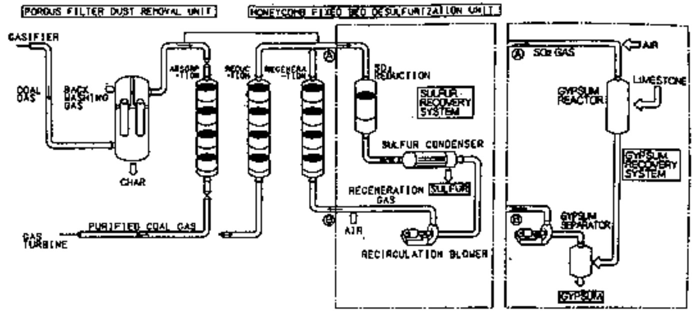
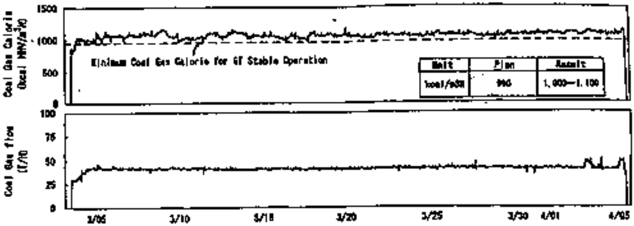


Figure 3 Fixed Bed Hot and Dry Gas Clean-up System

Honeycomb Fixed Bed Desulfurizatio Agent

Porous Ceramic Filter Element

Figure 4 Fixed Bed' Hot & Dry Gas Clean-up Element



Unit	Plan	Result
kcal/m3N	960	1,000~1,100

Figure 5 One Month Continuous Operation Data of 200 ton/day Gasifier

	Result	Plan
Sulfur Concentration (Inlet)	570~670ppm	Max. 1500ppm
ditto (Outlet)	20~50ppm	Less than 100ppm

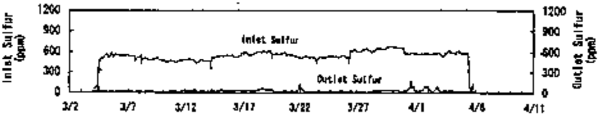


Figure 6 One Month Continuous Operation Data of 20 ton/day Fixed Bed Gas Clean-up

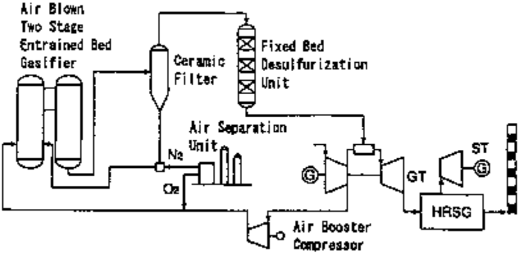


Figure 7 System Flow Diagram of the Mitsubishi HERCULES

IGCC Development Study on 6t/D Coal Texaco gasifier Test Facility

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1. Abstract

Gasification test of heavy oil and Orimulsion® was conducted with our gasification pilot plant (CGT), and also hot gas clean up test was done. First of all, we confirmed a steady operation of both gasifier and hot gas clean up system with these feed stock. And, high gasification performance was attained on IHI-Texaco gasifier for these fuels. Hot gas clean up test showed high desulfurization efficiency and stable regeneration of sorbent for long term operation using zinc oxide sorbent instead of iron oxide. From these tests, we confirmed heavy oil and Orimulsion® are excellent feed stock for Integrated Gasification Combined Cycle.

2. Introduction

The integrated gasification combined cycle system (IGCC) has being prospected as a next-generation power plant as high efficiency and its environmental performance. In recent, the commercialization of IGCC system is under way in many countries . To improve the gasification technology suitable for Japan, Tokyo Electric Power Company (TEPCO) , Kansai Electric Power Co., Inc. (KEPCO) and Ishikawajima-Harima Heavy Industries Co., Ltd. (IHI) are jointly conducting the gasification test and hot gas clean up system test with a 6 t/d Texaco type gasification pilot plant. IGCC features it's feed stock flexibility, for example coal, heavy oil and etc. We have already tested 15 kinds of coal and operated coal gasification tests about 5000 hours since 1987.

Heavy oil and orimulsion are the suitable feed stock for the Texaco coal gasifier which is fed coal as coal water mixture, in respect of facility utilization and gasification reaction. We carried out gasification tests to research gasification performance, operation condition, and gas clean up test with CGT pilot plant. This paper present the test results of this pilot plant.

3. Gasification tests

3.1 Test facility

IHI built a 6t/d pilot plant (CGT Coal Gasification Test facility) based on the Texaco process. The specification of CGT is shown in Table 1 and the flow sheet is shown in Fig.l. The fuel is pressurized by feed pump and fed to the gasifier. Heavy oil is high viscosity at room temperature, and for stable fuel feeding heavy oil should be heated to high temperature to degrease fuel viscosity. We provide steam trace for fuel tank and fuel line in CGT. Heavy oil is controlled maximum temperature at 200 degree C. Oxigen is used as oxidant. In case of heavy oil gasification, we controlled gasification temperature by steam as moderator. Orimulsion®is contained moisture about 30wt%, so we need not steam as moderator for gasification. In the gasifier at high temperature and high pressure, fuel reacts with oxygen (partial oxidation), and are converted to syngas whose main components are CO and H2. A char recycle system is not provided since the amount of unconverted char is very low. Ash in the fuel is fused and turn to slag. The slag is discharged through the lock hopper system. But heavy oil and Orimulsion has low ash contain, so recovery slag is small in quantity. Syngas is cooled by syngas cooler located under the gasifier and fed to the hot gas clean up system. Finally desulfuerization gas is incinerated by the flare stack.

3.2 Gasification fuel

The analysis data of gasification fuel used in CGT is shown in Table 2. We selected C grade heavy oil, asphalt as heavy oil and orimulsion®. From table 2, heavy oil has a low ash and water. Sulfur content in fuel is higher than that of coal. Orimulsion is an emulsion fuel consisting of Orinoco tar and about 30wt% water. Orimulsion* highly contents heavy metals, specifically vanadium and magnesium.

These fuel are cracked for light hydrocarbon gas (such as acetylene) and fixed carbon into the gasifier. The light hydrocarbon gas reacts with oxygen, steam and syngas (CO and C02) in the gasifier by partial oxidation at high temperature. Basic reaction in the gasifier are shown in equation (1)~(5) .

$$CmHn + \frac{m}{2}O_2 \leftrightarrow mCO + \frac{n}{2}H_2 \qquad (1)$$

$$CmHn + mO_2 \leftrightarrow mCO_2 + \frac{n}{2}H_2 \qquad (2)$$

$$CmHn + mCO_2 \leftrightarrow 2mCO + \frac{n}{2}H_2 \qquad (3)$$

$$CmHn + mH_2O \leftrightarrow mCO + (m + \frac{n}{2})H_2 \qquad (4)$$

$$CmHn + 2mH_2O \leftrightarrow mCO_2 + (2m + \frac{n}{2})H_2 \qquad (5)$$

These are very radical reaction in the gasifier. Another gasification reaction is the reaction between fixed carbon and gas components . But the amount of fixed carbon is a small, so it can not give big influence for the gasification performance. Heavy oil and orimulsion have good gasification performance and those are the suitable fuel for the gasification.

3.3 Gasification operating results

For each fuel, the operations were carried out smoothly and the gasification performances were stable and good. Long term operation is prospected.
From CGT test results, we have confirmed the stable gasification temperature for each heavy oil. Orimulsion was supplied into the gasifier at 50 degree C to keep its quality and good splay from the burner. After the test runs, 2-3 mm thickness of ash and uncovered carbon (soot) deposition on syngas cooler heating surface were found, while no ash deposit on other area. Those deposited ash and soot were very soft, and could be blow off easily.

3.4 Gasification performance results

The main indexes for the gasification performance are carbon conversion, cold gas efficiency and O/C ratio, and these are defined as follows.

$$\text{Carbon conversion} = \frac{\text{C in syngas (kg)}}{\text{C in fuel (kg)}} \times 100\ (\%)$$

$$\text{Cold gas efficiency} = \frac{\text{Heat value of syngas (J)}}{\text{Heat value of fuel (J)}} \times 100\ (\%)$$

$$\text{O / C ratio} = \frac{\text{O in oxidant (atom)}}{\text{C in fuel (atom)}}\ \text{(atom / atom)}$$

The relationship between O/C ratio and carbon conversion and cold gas efficiency are shown in Fig. 2, 3. High carbon conversion were achieved at higher O/C ratio. In each tests, cold gas efficiency increases by reducing O/C ratio. Maximum cold gas efficiency values were 79% for orimulsion gasification and 82% for heavy oil gasification. In case of asphalt gasification, carbon conversion on same O/C ratio tend to rise with steam ratio. Because oxygen atomic in steam that supplied into the gasifier to control gasification temperature contribute gasification reaction, so carbon conversion tend to rise with steam ratio. In view of these facts, the balance of steam and oxygen is very important for heavy oil gasification. The relationship between gasification temperature and carbon conversion and cold gas efficiency are shown in Fig. 4, 5. High carbon conversion was achieved at higher gasification temperature . But heavy oil gasification, gasification temperature was determined oxygen and steam balance. So carbon conversion and cold gas efficiency can not be evaluated with gasification temperature. Syngas composition of heavy oil and orimulsion are shown in Table 3. Syngas heating value of heavy oil gases are higher than that of orimulsion gas.

The main results from these gasification tests are as follows.

(1) Very stable operation and excellent gasification performance of high carbon conversion and high cold gas efficiency were confirmed.
(2) Heavy oil and orimulsion are very promising fuel for IGCC power plant.
(3) In case of heavy oil gasification, steam supplied into the gasifier contribute to gasification performance.

4. Desulfurization test

4.1 Background

The syngas generated at the gasifier is reductive gas, so the sulfur in the fuel is reduced to H2S and COS. Many wet type gas clean up system using some liquid solvents are available. But we have been developing the higher performance and efficient system using oxide metals sorbent, what we call Hot Gas Clean-Up System with fluidized bed reactors. The sorbent is continuously sulfurized, regenerated and recycled between the desulfurizer and regenerator, so the operation and performance are stable. This hot gas clean-up process has been developed on a national project. The first hot gas clean-up pilot plant for the 40t/day coal gasifier was constructed in Yubari, Hokkaido. On the basis of R&D results with this pilot plant, 200t/day IGCC pilot plant was constructed and finished the demonstration test at Nakokso, Fukushima Prefecture. We tested using the iron oxide as sorbent at these two pilot plant. We achieved total 3000 hours on stream performance at Nakoso pilot plant and got useful desulfurizing performance data.

We think more effective sorbent is required for more stringent environmental requirement with heavy oil or orimulsion fueled gasification. Because the syngas with Texaco type gasifier content much moisture, we need to change the desulfuring sorbent from iron oxide to zinc oxide. The following equations shows reactions of desulfurization and regeneration of iron oxide and zinc oxide.

$$Fe_2O_3 + H_2 \rightarrow 2Fe_3O_4 + H_2O$$

Desulfuer stage

$$Fe_3O_4 + 3H_2S \rightarrow 3FeS + 4H_2O$$

Regenerate stage

$$2FeS + \frac{7}{2}O_2 \rightarrow Fe_2O_3 + 2SO_2$$

Zinc oxide

Desulfuer stage

$$ZnO + H_2S \rightarrow ZnS + H_2O$$

Regenerate stage

$$ZnS+\frac{3}{2}O_2 \rightarrow ZnO+H_2O$$

More moisture was generated on desulfurization stage with iron oxide than zinc oxide, so the desulfurization efficiency decrease at high concentration of moisture. Fig.6 shows the comparison between zinc oxide and iron oxide, H2S equilibrium outlet concentration. We confirmed that theoretical equilibrium value decrease at the high concentration level of moisture. So, IHI test the hot gas clean-up system with zinc oxide sorbent at CGT(Coal Gasification Test facility) to confirming desulfuring efficiency and applicability to this process.

4.2 Hot gas clean up system of CGT

Fig. 7 and Table.4 show hot gas clean-up system at CGT. Firstly, the syngas generated at Texaco type gasifier is led to desulfurizer with two stage fluidized bed. Desulfurized syngas with dust is treated by finally dust removal system. The regenerator is one stage fluidized bed reactor. The regeneration gas with a few % of 02 (other is N2) is heated up before led to regenerator. Same as the desulfurizer, the dust removal system is provided at outlet of regenerator for removing sorbent dust. The sorbent are transported form first stage desulfurizer to the transport pot by gravity. And, the sorbent is lifted up to regenerator though the raiser with lift gas which consist of N2 . The regenerated sorbent is transported to the second stage desulfurizer by gravity. By down comer pipe at the desulfurizer, the sorbent is transported from second stage to first stage. In IHI type Hot Gas Clean-Up System, the sorbent are continuously sulfurized, regenerated and recycled between desulfurizer and regenerator. The sorbent circulation amount is controlled rotary feeder between desulfurizer and sorbent transport pot.

4.3 Desulfurization operating results

The desulfurizer is operated at 400-500 C , and about 680-700 C for regenerator. The dust removal system is operated well.

The Hot Gas Clean-Up System is operated over total 200 hours at the orimulsion gasification gas. We confirmed the steadiness of the system with zinc oxide sorbent.

4.4 Desulfurization test results

The desulfurization performance is evaluated by total sulfur concentration at outlet the desulfurizer. Generally, the desulfurization performance is higher at the low level sulfur concentration in sorbent. In increasing of sulfur concentration in sorbent, the desulfurizaton performance is decreasing. The sulfur content in sorbent is defined as follow.

$$\text{Sulfur Content in Sorbent(wt\%)} = \frac{\text{Absorbed Sulfuer Weight(kg)}}{\text{Sorbent Weight(kg)}} \times 100$$

The regeneration performance is evaluated by the sulfur content in sorbent at outlet regenerator. Both of two indexes(total sulfur concentration and sulfur content in sorbent) are used for analysis of test data.

Fig . 8 shows the trend of total sulfur concentration at outlet desulfurizer . The total sulfur concentration decrease rapidly just after starting regeneration. After that continuous operation have been performed for over 65houes. In the midst of operation, the sulfur content in sorbent decrease by increasing the circulation flow of sorbent. So, We achieved higher desulfurization performance. Fig.9 shows the desulfurization test result. We confirm the total sulfur concentration at outlet desufurizer is decreasing with decreasing the sulfur content in sorbent. And, Fig. 10 shows the regeneration test result. The retention time for sorbent at regenerator is about 30-40mins to decrease the sulfur content in sorbent to less than 1.0 wt%.

We confirmed follows.

1)The desulfurization performance is increased with decreasing the sulfur content in sorbent.
2)Hot Gas Clean-Up System with zinc oxide sorbent was proved it's good and steady performance.
3)The sorbent retention time at regenerator of 30mins is enough for regeneration.
4)Zinc oxide sorbent is applicable at Hot Gas Clean-Up System and is confirmed high desulfuerization performance.

5. Conclusion

The environmental requirement have become more stringent. IGCC is expected to be realized in the near future in view of its high efficiency and environmental superiority. A series of gasification tests goes on now at CGT and the main theme of recent tests are gas clean up test, improvement gasification performance and plant reliability.

Table 1 CGT outline

Gasifier type	One stage Entrained flow Texaco type
Capacity(Coal)	250kg/hr
Coal feed system	Coal water slurry feed
Capacity(Heavy oil)	180kg/hr
Capacity(Orimulsion)	240kg/hr
Pressure	1.96Mpa (20kgf/cm^2)
Temperature	Approx. 1400℃
Oxidant	Oxygen
Pressure of feed steam (Only heavy oil gasification)	3.43Mpa (35kgf/cm^2)
Gas cooling	Syngas cooler

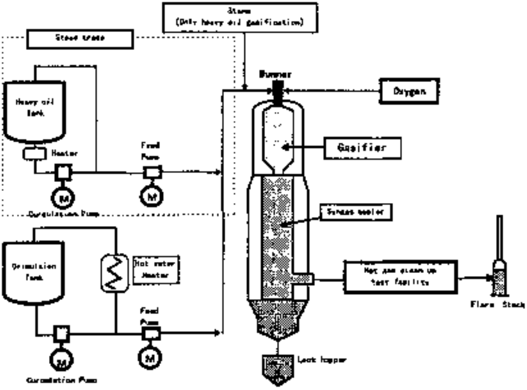

Fig.1 CGT flow sheet

Table 2 Fuel composition

	unit	C grade heavy oil	Asphalt	Orimulsion
Carbon	wt%	85.2	83.8	59.9
Hydrogen	wt%	11.2	10.1	7.23
Nitrogen	wt%	0.17	0.35	0.48
Oxygen	wt%	0.80	0.45	1.04
Sulfur	wt%	2.63	5.25	2.74
Ash	wt%	0.02	0.07	0.28
Moisture	wt%	0.01	0.01	28.3
HHV	MJ/m^3(normal)	43.5	41.6	29.8

Table 3 Syngas composition

		Heavy oil (Asphalt)	Orimulsion	Coal
CO	%	50.0	43.9	44.6
CO_2	%	4.42	9.31	15.4
H_2	%	43.9	45.4	34.6
CH_4	%	0.16	0.06	0.01
H_2S	%	1.22	0.93	0.17
HHV	MJ/m^3(normal)	12.3	11.7	10.3

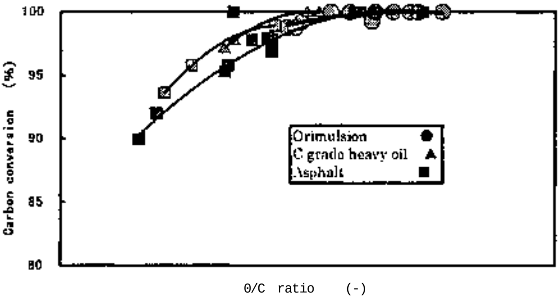

Fig.2 Relationship between O/C ratio and carbon conversion

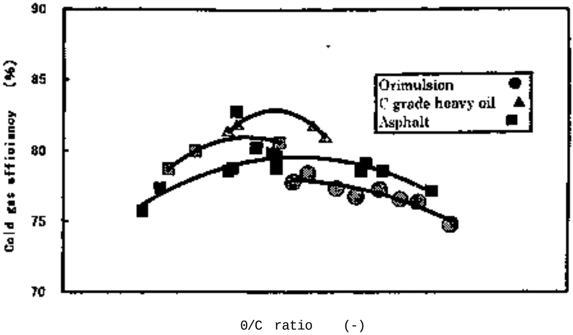

Fig.3 Relationship between O/C ratio and Cold gas efficiency

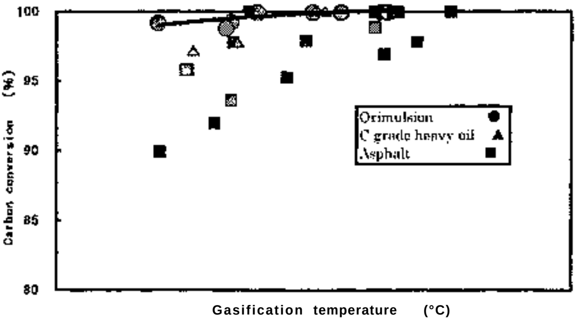

Fig.4 Relationship between Gasification temperature and carbon conversion

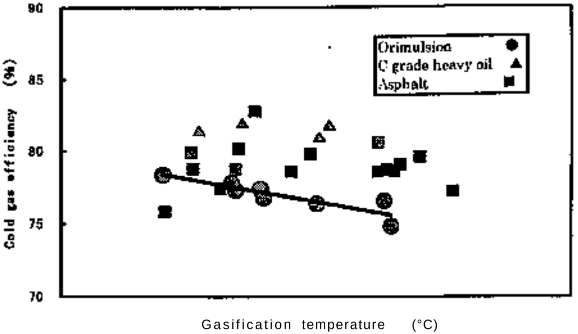

Fig.5 Relationship between Gasification temperature and carbon conversion

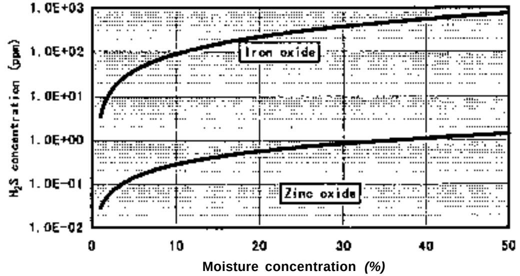

Fig.6 Comparison between zinc and iron oxide about H_2S equilibrium value

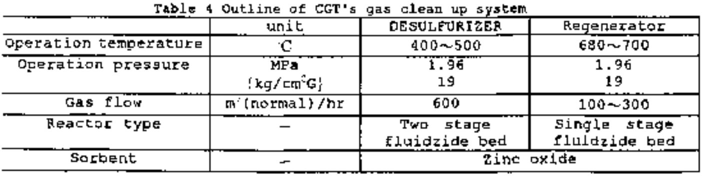

Table 4 Outline of CGT's gas clean up system

	unit	DESULFURIZER	Regenerator
Operation temperature	℃	400～500	680～700
Operation pressure	MPa {kg/cm^{2}G}	1.96 19	1.96 19
Gas flow	m^3(normal)/hr	600	100～300
Reactor type	—	Two stage fluidzide bed	Single stage fluldzide bed
Sorbent	–	Zinc oxide	

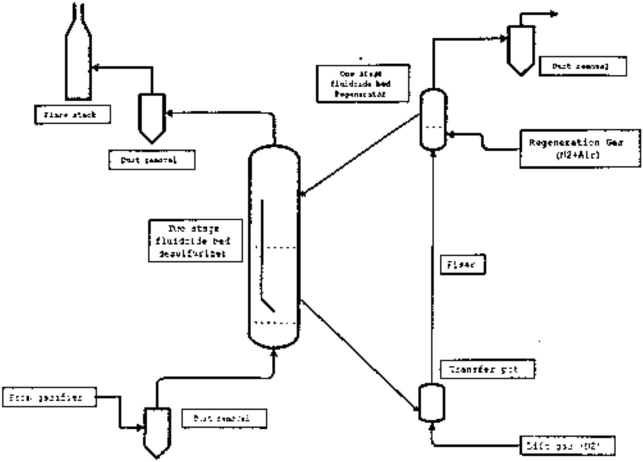

Fig7. Flow sheet of gas clean up system

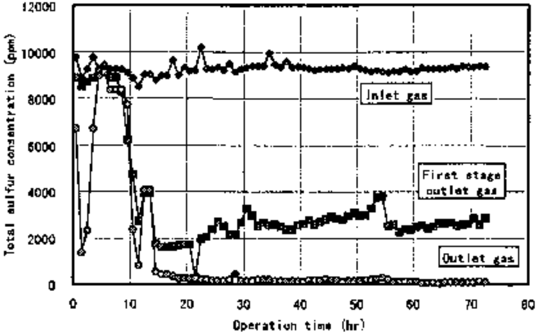

Fig. 8 Desulfurization test trend

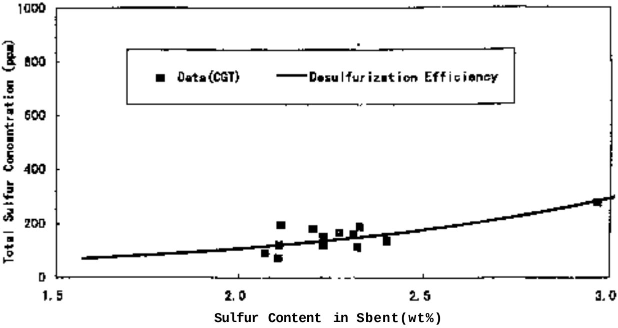

Fig. 9 Sulfr Content in Sorbent vs Total Sulfur Concentraion

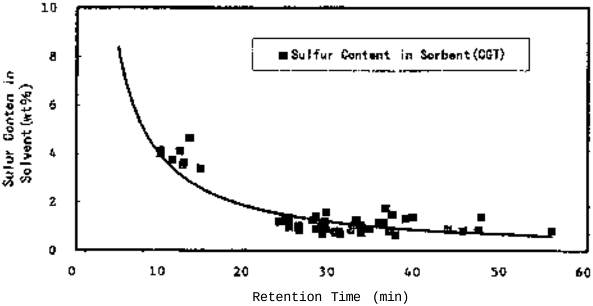

Fig. 10 Retention Time at Regenerator vs Sulfur Contnt in Srbent at Outlet Regenerator

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api ENERGIA 280 MW IGCC PLANT: A DESCRIPTION OF THE PROJECT FROM THE TECHNICAL AND CONTRACTUAL POINTS OF VIEW

Igino M. Chellini, Paolo V. Chiantore (api anonima petroli italiana S.p.A)
Roberto Del Bravo, Francesco Starace (ABB Sae Sadelmi S.p.A.)

ABSTRACT

api ENERGIA S.p.A., Roma, Italy is building a 280 MW Integrated Gasification Combined Cycle plant in a refinery at Falconara Marittima, on Italy's Adriatic coast, The IGCC will use heavy oil residues produced in the refinery. This paper provides the latest information on the status of this project as well as background information on plant history, technical features, integration with the refinery and contractual framework.

This is one of the three IGCC plants under construction in Italy following the liberalization of the electricty production sector and the introduction of governative decrees regulating the transfer, exchange and wheeling of the electricity.

The project, sponsored by api anonima petroli italiana, a private italian oil company, and ABB Sae Sadelmi, an italian engineering contractor of the ABB group, is a commercial IGCC plant, realized through a limited recourse international project financing.

The api ENERGIA venture achieved financial closing on November 22, 1996.

PROJECT HISTORY

The api ENERGIA IGCC project is the result of years of optimization studies carried-out to overcome the constraints that refineries are expected to face within this turn of the century.

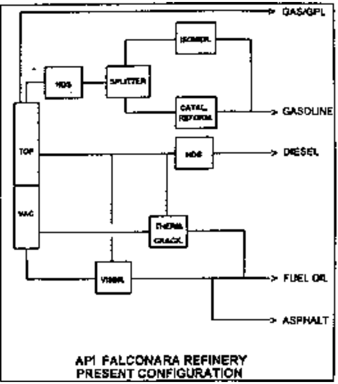
Demand for heavy fuel oil products in Western Europe is rapidly declining. Only low or very-low sulfur fuel oils will continue to have a market, due to environmental regulations.

Therefore the most immediate challenge for the refineries is the reduction of the amount of sulfur in the fuel oil or the conversion of residues to more valuable products.

The refinery is located in Falconara Marittima, Marche region, Italy with a 80,000 bpd (4 million ton/year) crude oil capacity. The refinery is owned and operated by api raffineria di ancona S.p.A., a wholly owned company of api anonoma petroli italiana S.p.A., holding and operating company of the api group..

The api group, owner of the Falconara refinery, is one of the largest private industrial group in the Italian oil sector and, with over Lit 5,500 billion (3.7 billion US$) consolidated annual revenues, is one of the main industrial groups in Italy api supplies, refines, markets, trades, distributes and sells oil and oil products.

The Falconara refinery is a oil refinery based on a thermal conversion scheme. The production cycle is typical for refineries of this capacity with distillates yield of more than 70% and asphalts production of more than 8%. The distillates are produced to meet the most stringent european quality specifications. All the units have been installed, completely modernized or revamped over the past ten years to maintain economical and technological competitiveness in the oil market, while complying with the highest safety and environmental criteria. The refinery cycle includes atmospheric and vacuum distillation, gasoline desulfurization, isomerization and catalitic reforming, gasoil desulfurization, thermal cracking of vacuum gasoil and visbreaking of vacuum residues (See the attached simplified flow diagram).



Falconara is in the center of the Adriatic coast. The refinery is strategically located in an area where no other refinery is present: this gives to api a significant local market advantage over competing Italian oil companies.

The two nearest refineries are more than 300 km (Venice) and 500 km (Taranto) away respectively. Additionally the Marche region is the poorest Italian region in terms of electric power production.

In order to eliminate the present production of high sulfur fuel oil, api examined many alternatives potentially available :

- operate on low sulfur crudes
- realize new plants to convert the residues into more valuable products

The first option is strategically very weak, leaving the refinery strongly dependant on a limited amount of foreign sources of supply, with a high degree of risk for the refinery survivability.

The second option was developed according to various alternatives based on the realizatiaon of the following new plants:

a) vacuum gasoil desulfurization, FCC, alkylation, flexicoker

b) vacuum gasoil hydrocracking, FCC, alkylation, flexicoker

c) atmospheric residue desuifurization, residue catalytic cracking

d) partial oxidation of visbreaking residue (IGCC)

As a first look, it was clear that options a), b) and c) involved a significant restructuring of the heavy-end section of the refinery, including modification or dismission of existing and well-operating units, such as thermal cracking or visbreaking, as well as a significant revamping of all the utilities production systems.

This table compares the alternatives studied from various points of view:

OPTION --->	a) VGO desulfuriz., FCC, alkyl, flexicoker	b) VGO hydrocr., FCC, alkyl, flexicoker	c) Atm. residue desulfuriz., RCC	d) IGCC
Safety	- more storage tanks - higher risk	- more storage tanks - higher risk	- more storage tanks - higher risk	- no additional risk
Production compared to api market (*)	- gasoline excess - gasoil deficit - no bitumen	- gasoline excess - no bitumen	- gasoline excess - no bitumen	- balanced
Environmental constraints	- coke disposal - exhaust with a lot of particulates - spent catalysts	- coke disposal - exhaust with a lot of particulates - spent catalysts	- exhaust with a lot of particulates - spent catalysts in very large amount	- minimal
Power balance	- small export	- small export	- import	- large export
Economic parameters (base=100 for case d)	- investment 107 - gross margin 75 - pay-out time 150	- investment 105 gross margin 85 - pay-out time 140	- investment 80 - gross margin 86 - pay-out time 100	- investment 100 - gross margin 100 - pay-out time 100
Overall impact on refinery cycle	- medium	- medium	- high	- minimal

(*) expected 1999 market, with no HSFO request

As a result of the comparison alternative d) was selected for the significant amount of advantages over all the other cases studied.

For the implementation of such a project, api decided to form a dedicated joint venture : on february 20, 1992 in Rome the api group and ABB Sae Sadelmi S.p.A. formed api ENERGIA S.p.A as a joint venture, special purpose company. The project company was estabilished to build, own and operate an IGCC plant at Falconara.

ABB Sae Sadelmi S.p.A. is a wholly owned subsidiary of ABB Italia S.p.A. which is a member of the ABB Asea Brown Boveri group. ABB Sae Sadelmi is one of the largest contractors in the global power business, active in 38 countries. ABB Sae Sadelmi's business include the design and construction of power plants, transmission lines, railway lines, airports, telecommunication systems, oil and gas plants, industry drives and robotics, environmental and water plants and the manufacturing of electrical generators, boilers, steel structures and towers. ABB Italia group includes 50 companies in Italy, 28 factories and has annual revenues of more than Lit 3,000 billion (2.5 billion US$).

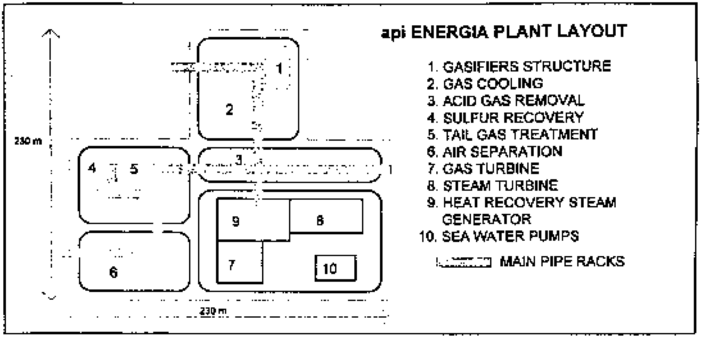
In the years 1993 to 1995 the api ENERGIA project development took place, based on studies of the appropriate project configuration and financing. During this period api ENERGIA was one of the pioneers as large Independent Power Producer in a country where the electricity has always been a monopoly and where project financing schemes of this kind and proportions had never been applied. The complex structure of the project was put in place by developing the contractual framework through which all the involved parties, including sponsors, contractors, suppliers, lenders, licensors and consultants are now working to implement the IGCC plant. The venture has executed a turn-key engineering, procurement and construction (EPC) contract with a temporary association of company within the ABB group, the "ABB Consortium", that includes ABB Sae Sadelmi (consortium leader), ABB Lummus Global Gmbh (Germany) and ABB Power Generation Ltd (Switzerland).

PLANT DESCRIPTION

The Process

The api ENERGIA IGCC plant is designed to gasify high-sulfur heavy oil produced by the Falconara refinery: the design feedstock is a vacuum visbroken tar from Arabian Heavy crude. The gasifiers are also able to process many other feedstocks, including tars produced by other crudes as well as other types of refinery residues such as vacuum and atmospheric residues, heavy gasoils and heavy crudes.

api ENERGIA will lease an area of approximately 47000 squared meters on the refinery's premises. A simplified plant layout shows the area occupied by the IGCC plant.



The IGCC process includes two technically integrated sections: the Syngas Manufacturing Process Plants (SMPP) and the Combined Cycle Power Plants (CCPP).

The SMPP is designed to convert the hydrocarbon feed into a clean syngas by using technologies licensed by Texaco, UOP, Parsons and ABB Lummus Global.

The SMPP produces a clean syngas mainly composed of CO and H2; this gas is then mixed with nitrogen to prepare a low-NOx fuel gas to be used in the gas turbine of the CCPP. In addition this section produces a pretreated water blowdownand, a liquid sulfur stream and a metal-rich sludge.

The CCPP is dedicated to the generation of electric power and steam through a combined cycle plant designed by ABB Power Generation. It mainly consists of a conventional cycle with a Gas Turbine (GT), a Heat Recovery Steam Generator (HRSG) and a Steam Turbine (ST). The main equipment of the CCPP is the ABB Type 13E2 Gas Turbine-Generator set designed for both syngas and back-up diesel fuel operation. The machine is equipped with dry low NOx burners.

The CCPP has a wide rangeability of operation due to the possibility of switching over from syngas to back-up fuel.

A major effort has been made by ABB to optimize the integration between the CCPP and the SMPP in order to maximize plant efficiency: there are steam extraction and admission devices in the steam cycle and in the ST for the integration with the gasification and with the refinery.

Main Plant Units

Texaco Gasification

The core of the SMPP is the Texaco Gasification System. api ENERGIA selected this technology for its significant commercial experience and environmental superiority.

The Texaco licensed section includes

- one Feed Preparation section,

- two Quench Gasification and Scrubbing sections, each rated at 56% of the plant throughput requirement (with Arabian Heavy feedstock),
- one Carbon Extraction unit

Grey Water Treatment section

A blowdown water stream frorm the Texaco gasification, containing metals is treated in the Grey Water Treatment unit to prepare a final water blowdown suitable for the existing refinery bio-treatment.

The unit performs physical/chemical treatment for destruction of cyanides and precipitation of salts and metals The resulting sludge is dewaterd to a minimum of 40%wt solids: this cake is discharged from this unit and sent to external plants for the recovery of the metals (Vanadium and Nichel). The treated grey water is filtered and stripped from sour gases before disposal via the refinery bio-treatment.

Gas Cooling and COS Hydrolisis

This section has been designed by ABB Lummus Global, with the aim of recovering the medium/low temperature duty available in the syngas after water scrubbing. The unit includes:

- a gas cooling train to recover the waste heat of the scrubbed gas by generation of steam at three pressure levels. Condensate formed during gas cooling is collected and returned to the syngas scrubber of the Texaco unit.
- a COS hydrolisis reactor to convert the small amount of COS produced in the gasifier to H2S for removal in the Acid Gas Removal unit.
- a gas expander for the recovery of the pressure energy of the syngas (pressure in gasifiers is about 65 bar)

Acid Gas Removal

A selective physical process, UOP's Selexol system has been selected, mainly because of its extensive and succesfull experience with Texaco syngas. In the unit a circulating Selexol stream contacs the syngas in an absorber in order to selectively absorb the sulfur compounds (mainly H2S). The solvent is then regenerated in a stripper producing an acid gas containing the H2S to be sent to the Claus unit.

Sulfur Removal and Tail Gas Treatment

The recovery of the sulfur contained in the H2S of the Selexol acid gas is accomplished in two Claus units licensed by Ralph M. Parson. These are conventional Claus systems using oxygen instead of air.

The Claus units are followed by a Tail Gas Treatment section thus allowing an overall sulfur recovery of 99.9%.

Air Separation

This unit is licensed by Praxair and produces the oxygen required for the gasification and the Claus plant (70 t/h NOC) and the nitrogen for syngas conditioning (127 t/h NOC). It is based on conventional criogenic air fractionation.

Gas Turbine

The ABB 13 E2 machine is an advanced, single shaft, heavy duty, industrial gas turbine with annular combustor chamber. The machine has a published ISO rating of 164 MW with natural gas. With the syngas produced in the Texaco gasifiers the gross output increases to 189 MW. The axial compressor of the machine has 22 stages, a pressure ratio of 16.7 and one row of variable inlet guide vanes, that ensures high efficiency operation even at part load. This parameters are specific for low-BTU gas firing. The gas burned in the annular combustor expands in a 5-stage turbine, equipped with an air-cooling system to protect rotor, blades and vanes against high temperature.

The gas turbine is able to operate from part load up to its saturation point, after which the excess syngas can be fired in the supplementary firing system realized in the HRSG.

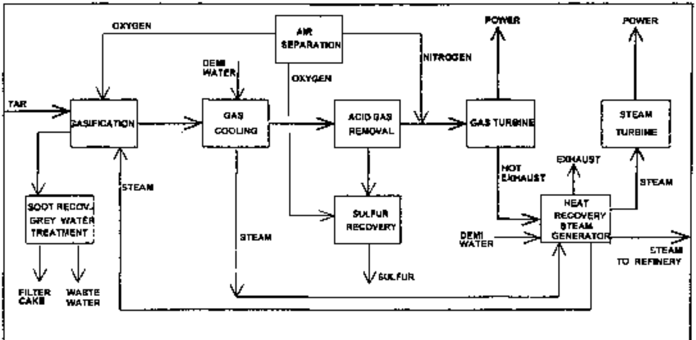
Heat Recovery Steam Generator

The HRSG is an horizontal heat recovery natural circulation boiler with three steam pressure levels, supplementary firing system and single stack: it recovers the heat of the gas turbine exhaust and must be regarded as the main equipment acting in the steam cycle of the entire plant.

Steam Turbine

The Steam Turbine is a multiple inlet extraction condensing type turbine with two cylinders. Live steam will pass through the high pressure / medium pressure casing then cross-over into the twin exhaust low pressure turbine. During NOC the steam turbine exports Medium Pressure Steam and Low Pressure Steam to the refinery. Piping for export of High Pressure Steam is also provided.

An auxiliary Steam Generator (back-up boiler) will be provided to insure continuous steam export to the refinery even when the gas turbine is unavailable for operation.



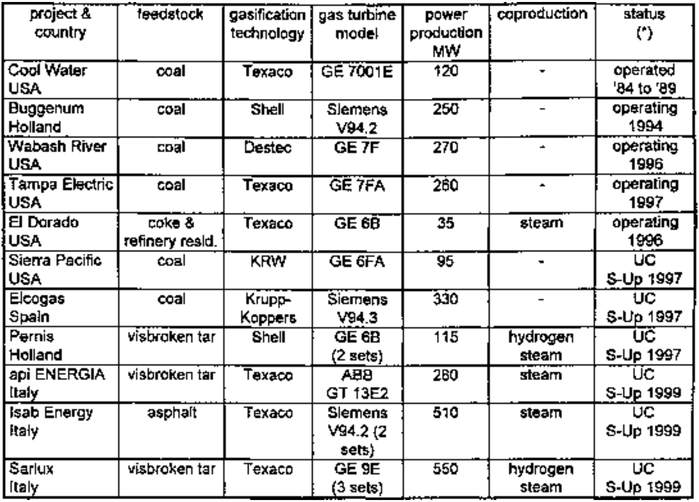
SIMPLIFIED BLOCK FLOW DIAGRAM OF THE api ENERGIA IGCC PLANT

Auxiliary Systems

The IGCC complex will be equipped with all the necessary auxiliary systems including cooling water (mixed system with an open seawater circuit for large users and a closed clean water circuit for the other users), demi water, air, nitrogen, water and fuel gas networks, firefighting, flare, storages, electrical distribution, buildings, etc.

The Technology

The IGCC tehcnology as such is quite a new approach in the field of power production. Experience of IGCC power plants is limited: the following table gives a list of IGCC plants which have commercial operating experience or are currently being constructed. Excluding Cool Water, the first IGCC plant, only few other has already gain commercial experience.



project & country	feedstock	gasification technology	gas turbine model	power production MW	coproduction	status (*)
Cool Water USA	coal	Texaco	GE 7001E	120	-	operated '84 to '89
Buggenum Holland	coal	Shell	Siemens V94.2	250	-	operating 1994
Wabash River USA	coal	Destec	GE 7F	270	-	operating 1996
Tampa Electric USA	coal	Texaco	GE 7FA	260	-	operating 1997
El Dorado USA	coke & refinery resid.	Texaco	GE 6B	35	steam	operating 1996
Sierra Pacific USA	coal	KRW	GE 6FA	95	-	UC S-Up 1997
Elcogas Spain	coal	Krupp-Koppers	Siemens V94.3	330	-	UC S-Up 1997
Pernis Holland	visbroken tar	Shell	GE 6B (2 sets)	115	hydrogen steam	UC S-Up 1997
api ENERGIA Italy	visbroken tar	Texaco	ABB GT 13E2	280	steam	UC S-Up 1999
Isab Energy Italy	asphalt	Texaco	Siemens V94.2 (2 sets)	510	steam	UC S-Up 1999
Sarlux Italy	visbroken tar	Texaco	GE 9E (3 sets)	550	hydrogen steam	UC S-Up 1999

UC = under construction, S-Up = Start up year

The two key technologies of an IGCC plant are the gasification and the gas turbine. All the other systems of an IGCC plant are conventional and widely used in refeining and petrochemical plants.

The experience of those IGCC plants that have already operated for some time shows that the majority of the problems encountered were attributable to the operation on coal and coal ashes; only the Buggenum plant has experience considerable problems related to the gas turbine, which in any case has been considered unique of that machine. Other typical problem were related to the high degree of integration of certain schemes.

Regarding the gasification technology, there are numerous examples of commercial applications operating on different feedstocks. Texaco, that is the world leader licensor of gasification technologies, has licensed more than 100 commercial plants, of which approximately 60% based on liquid feed, 30% on gaseous feed and 10% on solid feed.

Texaco technology has been in commercial operation since more than 40 years. This means that the technology is well known, and although there is limited experience in using the syngas as fuel for gas turbines, there is no major concern in using this technology.

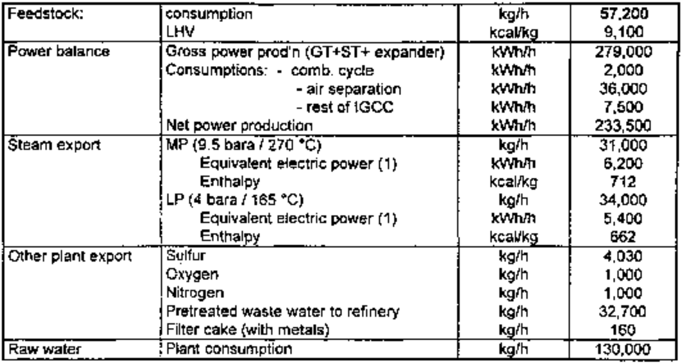
Two alternative route are feasible under the Texaco licence: the indirect gas cooling system, using a syngas Waste Heat Boiler or the direct gas cooling system using the Direct Quench gasifier. For this project api ENERGIA selected the Direct Quench gasifier for its reliability and ease of operation and maintenance; in fact, for feedstocks with metal content up to 800 ppm, such as the maximum expected for the api ENERGIA project, there is the potential risk of plugging in the waste heat boiler tubes.

The gas turbine as such is a common equipment in the power industry: thousands of these machines are in operation world-wide in simple or combined cycle operation. Compared to the standard application with natural gas or gasoil feedstock, limited experience is available with low-BTU fuel gas, such as that produced with gasification. Nevertheless, the modification to be realized on the machine are well known. The ABB GT 13E2 is the direct development of the GT 13E, the main modification being the combustion system incorporating an annular combustion chamber. More than 40 GT 13E2 gas turbines have been sold since 1993. A major test has been conducted on the combustor of this machine with a synthesis gas identical to the one that will be produced in the api ENERGIA IGCC plant to study the turbine behaviour and define the modifications to the standard design.

The other technologies, i.e. air separation, gas cooling, COS hydrolisis, acid gas removal, sulfur recovery and tail gas treatment as well as the integrated power generation of a combined cycle, are all based on standard processes widely applied in chemical and petrochemical plants.

Plant Performance

The following table shows the main plant data with a feedstock derived from crude operation with a mix of 65% arabian heavy and 35% arabian light.



Feedstock:	consumption	kg/h	**57,200**
	LHV	kcal/kg	**9,100**
Power balance	Gross power prod'n (GT+ST+ expander)	kWh/h	**279,000**
	Consumptions: - comb. cycle	kWh/h	**2,000**
	- air separation	kWh/h	**36,000**
	- rest of IGCC	kWh/h	**7,500**
	Net power production	kWh/h	**233,500**
Steam export	MP (9.5 bara / 270 °C)	kg/h	**31,000**
	Equivalent electric power (1)	kWh/h	**6,200**
	Enthalpy	kcal/kg	**712**
	LP (4 bara / 165 °C)	kg/h	**34,000**
	Equivalent electric power (1)	kWh/h	**5,400**
	Enthalpy	kcal/kg	**662**
Other plant export	Sulfur	kg/h	**4,030**
	Oxygen	kg/h	**1,000**
	Nitrogen	kg/h	**1,000**
	Pretreated waste water to refinery	kg/h	**32,700**
	Filter cake (with metals)	kg/h	**160**
Raw water	Plant consumption	kg/h	**130,000**

(1) Electric power not produced due to steam extraction from steam turbine

Based on the above data it is possible to calculate an overall IGCC efficiency by including the equivalent theoretical electric power of the exported steam (see note 1 above):

- Net (theoretical) power production 233.5 + 6.2 + 5.4 = 245.1 MWh/h
- Tar consumption in kWh/h 57.2 t/h * 9,100 Mcal/t / 860 Mcal/MWh = 605.26 MWh/h

- *Net IGCC efficiency: 245.1/ 605.26 * 100 = 40.5%*

Similarly the cogeneration (power plus heat) efficiency could be calculated by considering the steam enthalpy instead of the equivelant power; in this case the calculation gives:

- *Net cogeneration efficiency: 47.2%*

Environmental Impact

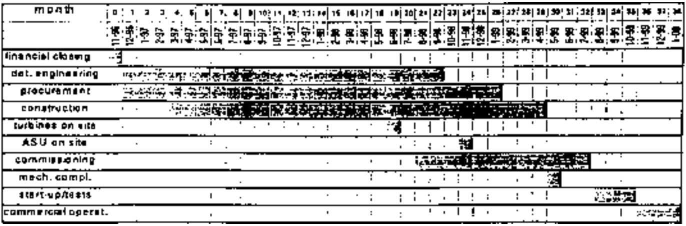
The IGCC plant as such is one of the cleanest system today available for power production, and certainly the cleanest of all with heavy, high sulfur fossil fuels such as coal or refinery residues. This was one of the reasons api ENERGIA selected this technology.

The following tables summarizes the main api ENERGIA IGCC plant stack emissions.

SO2	630 ton/year	50 mg/Nm3 dry at 15% O2
NOx	720 ton/year	53 mg/Nm3 dry at 15% O2
CO	115 ton/year	8 mg/Nm3 dry at 15% O2
Particulates	63 ton/year	5 mg/Nm3 dry at 15% O2

Project schedule

The api ENERGIA venture achieved financial closing in november 22, 1996. According to the EPC contract, the turn-key contractor has undertaken to deliver to api ENERGIA the completed plant, ready for commercial operation, in 35 months since financial closing. The plant will be completed in spring of 1999 while start-up and trial runs will occurr in the second half of the same year. The attached table shows the project schedule.



The following list of selected macro-numbers gives a better idea of the dimension of the project:

- 2 millions of man-hours will be used for the erection of the plant: during the peek period 900 people will be working at plant site
- more than 400 equipment will be installed
- 41000 m^3 of foundation excavation will be required
- more than 28000 m^3 of concrete will be used
- 16400 ton of steel will be required
- 430 km of cables will be used

PROJECT FINANCING

On November 22, 1996 the api ENERGIA IGCC project achieved financial closure: this is a venture financed through a limited recourse project financing scheme. This means that the api ENERGIA IGCC plant is a 100% commercial plant. We are proud to say that, through this achievement the demonstration phase of the IGCC technology for power production must be considered concluded.

To get such a result all the tasks related to the implementation of this project have been discussed and solved with the loan providers in order to be granted a financing.

The priority topic was of course an accurate cash flow analysis forecast. In fact, if the main economic indicators (IRR, NPV, debt/coverage ratio) have acceptable values, the possibility to obtain a project financing of this kind is substantially based on the recognition of the lenders that the forecast on that cash flow model is realistic and that the risk associated with it are sufficiently and proportionally carried by the project participants.

The economic model has been deeply studied by the sponsors and takes into account all the peculiar conditions of the italian market and legislation for this plant and in particular the tariff to be applied. The Italian legislators has developed a tariff structure that compensates for the high capital costs of clean power generation systems (such as IGCC) that use residues with no commercial value. The two main driving forces behind this approach are:

- promote energy saving

- use all the possible domestic resources considering the strong dependance of the country on oil, coal and gas import for enrgy production.

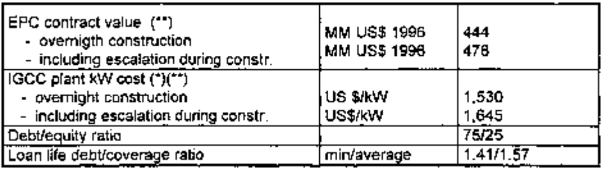
Anyway, the flat value of the tariff over the twenty years life of the electricity contract is averaging international standards. In fact for the IGCC plants the average value for the 20 years project life (values for the year 1995, Lit/USD=1,550) is 7.0 cents/kWh (108.5 Lit/kWh). During the first 8 years the tariff is 9.0 cents/kWh (139.4 Lit/kWh) because it includes a component to take into account the repayment of the financing, while in the remaining 12 (or more) years the tariff drops to 5.7 cents/kWh (87.9 Lit/kWh) because it does not include that component.

The other major topics discussed and accepted by the lenders have been the analysis of the risks related with the project and the guarantees to be provided by the participants. The main area of risks include the maturity of the technology and its capability to be the base for a reliable plant, the power sales agreement, the feedstock supply and the integration with the refinery, mainly the steam supply agreement. All of these issues are today covered by detailed contracts which include guarantees and liabilities.

Audits on the refinery survivability over the next 25 years have been carried out by independent firms to verify the capability of the hosting industrial structure to maintain a competitive position in the Italian marketplace, thus providing the adequate support to the initiative over its minimum 20-years life. These studies highlight that the refinery is small-medium size and therefore suffers some disadvantages in terms of economy of scale, but it has a large advantage based on its location and on the higher transportation costs that any competitor would have to incur in entering its market. Additionally the introduction of the IGCC plant will enhance the refinery profitability.
The positive assesments of these studies have been an additional element provided to the lenders to positively judge on the bankability of the project.

All of these aspects, which are the results of years of studies, negotiations and agreements, have been the basis for the succesfull financial closure of the api ENERGIA IGCC project.

The main economic data of the project are summarized in the attached table.



EPC contract value (**) - overnigth construction - including escalation during constr.	MM US$ 1996 MM US$ 1996	444 476
IGCC plant kW cost (*)(**) - overnight construction - including escalation during constr.	US $/kW US$/kW	1,530 1,645
Debt/equity ratio		75/25
Loan life debt/coverage ratio	min/average	1.41/1.57

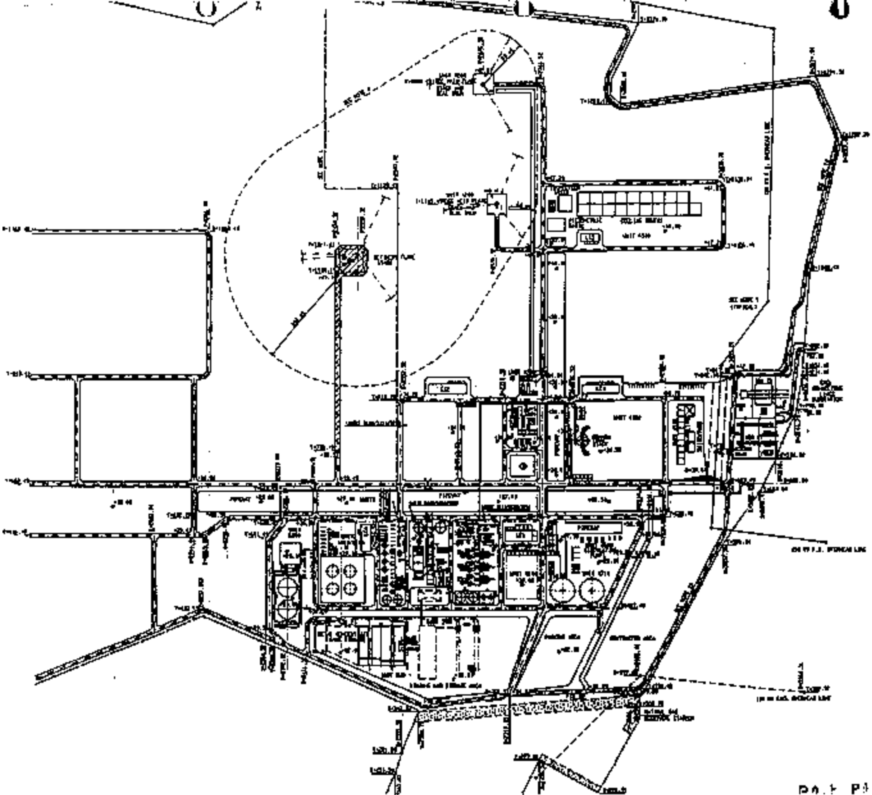
(*) power production includes the equivalent production of the exported steam

(**) including Air Separation Plant

CONTRACTUAL FRAMEWORK

A number of contracts govern the api ENERGIA IGCC project. The following list describes the main companies involved in the contractual frame.

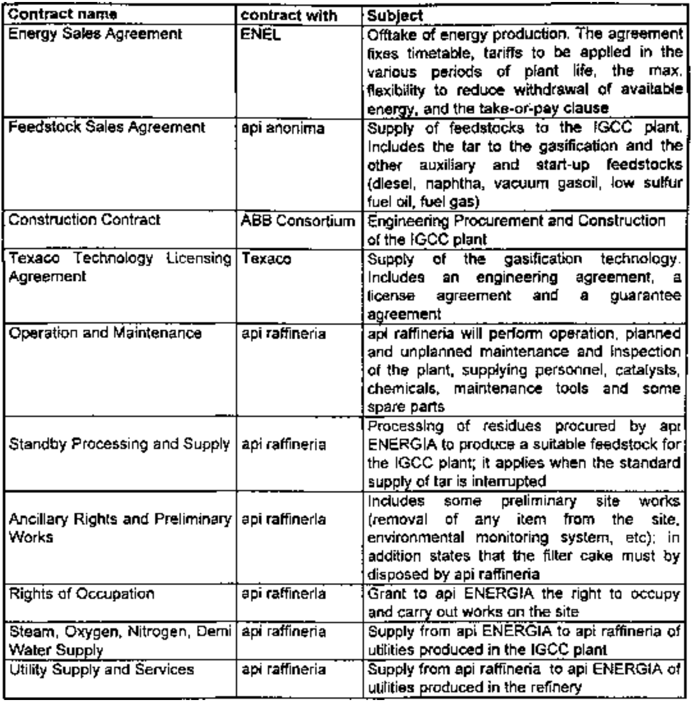
- **api ENERGIA S.p.A. (I),** 51% owned by the api group (I) and 49% by ABB Sae Sadelmi S.p.A. (I) is the owner of the IGCC project.
- **api anonima petroli italiana S.p.A. (I),** holding company of the api group, supplies the feedstocks
- **api raffineria di ancona S.p.A. (i),** company of the api group, is responsible of the operation and maintenance of the IGCC plant
- **ABB Consortium,** led by ABB Sae Sadelmi S.p.A., includes ABB Lummus Global GmbH (D) and ABB Power Generation Ltd (CH), is the Turn-key EPC contractor



Plot Plan

- **ENEL S.p.A. (I),** is the offtaker of the electric power
- **Texaco Development Corporation (USA),** provides the gasification license

This table gives on overview of the contracts related to the api ENERGIA project



Contract name	contract with	Subject
Energy Sales Agreement	ENEL	Offtake of energy production. The agreement fixes timetable, tariffs to be applied in the various periods of plant life, the max. flexibility to reduce withdrawal of available energy, and the take-or-pay clause
Feedstock Sales Agreement	api anonima	Supply of feedstocks to the IGCC plant. Includes the tar to the gasification and the other auxiliary and start-up feedstocks (diesel, naphtha, vacuum gasoil, low sulfur fuel oil, fuel gas)
Construction Contract	ABB Consortium	Engineering Procurement and Construction of the IGCC plant
Texaco Technology Licensing Agreement	Texaco	Supply of the gasification technology. Includes an engineering agreement, a license agreement and a guarantee agreement
Operation and Maintenance	api raffineria	api raffineria will perform operation, planned and unplanned maintenance and inspection of the plant, supplying personnel, catalysts, chemicals, maintenance tools and some spare parts
Standby Processing and Supply	api raffineria	Processing of residues procured by api ENERGIA to produce a suitable feedstock for the IGCC plant; it applies when the standard supply of tar is interrupted
Ancillary Rights and Preliminary Works	api raffineria	Includes some preliminary site works (removal of any item from the site, environmental monitoring system, etc); in addition states that the filter cake must by disposed by api raffineria
Rights of Occupation	api raffineria	Grant to api ENERGIA the right to occupy and carry out works on the site
Steam, Oxygen, Nitrogen, Demi Water Supply	api raffineria	Supply from api ENERGIA to api raffineria of utilities produced in the IGCC plant
Utility Supply and Services	api raffineria	Supply from api raffineria to api ENERGIA of utilities produced in the refinery

THE IGCC PLANT WITHIN THE REFINERY

Integrations with the refinery

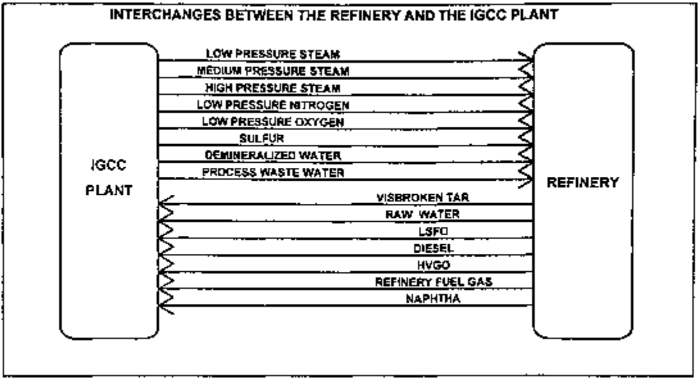
The impact on the existing refinery cycle will be minimal. The only modifications to the refinery include some pipework, tank storage rearrangement and utility services. In particular, the refinery has the responsibility of the following preliminary work necessary to allow api ENERGIA to construct and operate the plant:

- relocation of the refinery flare system
- interconnecting facilities for utilty exchange with the refinery
- control room

- rearrangement of area of some existing tanks
- relocation of firefighting training area
- relocation of one electrical substation
- provide facilities for the IGCC construction works (fencing, power, potable water, phone/fax lines..)

Some of this activities, in particular the reieocation of the main existing flare stack and rearrangement of the firefighting system have already been completed.

The following table shows the major interchanes between the refinery and the IGCC plant:



Benefits to the refinery: environment and economics

With the construction of the IGCC plant and the completion of the desulfurization units the refinery will become a "white refinery" with no High or Low Sulfur Fuel Oil (HSFO/LSFO) production. In fact, in addition to the IGCC, the refinery plans to build a new unit for the deep desulfurization of the thermal cracking residue, in order to produce a Very Low Sulfur Fuel Oil (VLSFO) with 0.25 %wt sulfur, suitable for direct firing in existing power stations without installing flue gas desulfurizers.

=> *$S0_2$ emissions*

The IGCC plant will dramatically reduce the emissions of $S0_2$: despite the fact that, after the start-up of the IGCC plant the Falconara site will produce 285 MW instead of the 15 MW today produced with steam boilers, the overall refinery emissions will be reduced. This will be achieved by shutting down the existing, conventional power/steam station that burns fuel oil. The results will be:

- overall SO_2 emissions from the refinery reduced by 30%
- SO_2 immission on the ground in the area reduced by 80%
- global reduction of SO_2 emissions by 380% due to the elimination from the market of 600,000 ton/year of heavy fuel oils responsible for the production of approximately 20,000 ton/y of SO_2.

=> *Economics*

It has been estimated that the api refinery margins will increase about 0.7÷0.8 $US/bbl when the IGCC plant will be in operation. The elimination of the HSFO, the increased processing flexibility, the long term contracts related to the IGCC plant as well as the utility integrations between the IGCC plant and the refinery, will be the main reasons of this improvement.

This estimation has been confirmed by independent consultants: studies indicate that refinery performances as measured according to world-wide accepted indices will measurably improve in several areas, allowing api refinery to reach higher positions in the international refinery rankings.

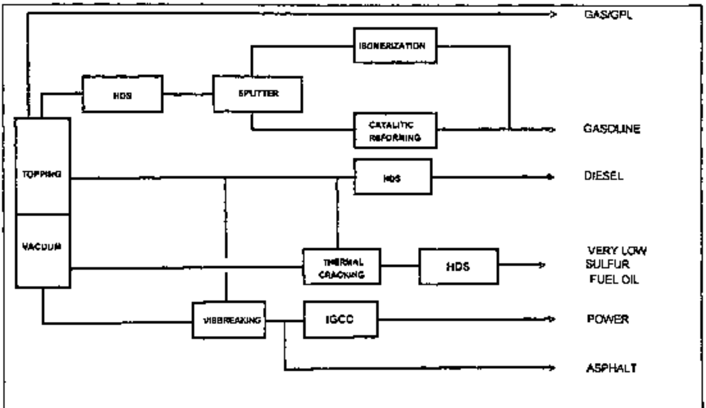
The year 2000 api refinery

The "year 2000 api refinery" will be a modern integrated site of petroleum refining and power production; by using advanced and flexible technologies the refinery will have the following advantages:

- ability to process heavy crude slates, with the possibility to maximize refinery profits by selecting the most economical crudes available in the mediterranean market at any time.
- ability to produce high quality and clean transportation fuels as mandated by the European Union.
- ability to produce high quality, high value and clean fuel oil for direct use in existing domestic or mediterranean power stations without the need to install desulfurization systems
- ability to optimize the production of high quality asphalt with the excess IGCC feed, to cover local market demand
- very low environmental impact, obtained by eliminating the old power and steam generation refinery station and integrating the refinery steam network with the IGCC plant

Adequate storage facilities will provide constant feedstock availability to the IGCC plant, while the auxiliary boiler in the IGCC plant will ensure constant steam delivery to the refinery even when the IGCC plant is out of service.

The attached diagram shows the refinery scheme after the major modifications expected by the year 1999: these include the IGCC and the thermal tar desulfurization plant.



API FALCONARA REFINERY CONFIGURATION IN 1999

References

1. F. Starace, C. Napolitano, R. Del Bravo "api Energia 250 MW IGCC Plant in Italy, Consideration Relevant to Project Financing of IGCC Projects" 13th EPRI Conference, S.Francisco, oct. 19-21, 1994
2. R. Del Bravo, G. Pignatelli "The Gasification Process Associated to Electrical Production: a New Alternative to Refineries Deep Conversion, api Energia 250 MW Commercial Plant in Italy" WEFA Energy Workshop, London, dec.2, 1994
3. K.A. Sheikh, H. Wen "Advances in Gasification Power Plant Technology" Power-Gen Europe, Cologne, may 1994
4. G.L. Farina, L. Bressan "IGCC Projects in Italy" 13th EPRI Conference, S.Francisco, oct. 19-21, 1994
5. D.L. Heaven "Gasification Converts a Variety of Problem Feedstocks and Wastes" Oil & Gas Journal, may 27, 1996
6. L.R. Aalund "Gasification, Polygeneration Capture Interest of Refiners" Oil & Gas Journal, dec 9, 1996
7. R. Del Bravo, F. Starace, I.M. Chellini, P.V. Chiantore "Italian IGCC Project Sets Pace for New Refining Era" Oil & Gas Journal, dec 9, 1996

A KEY ASPECT OF INTEGRATED GASIFICATION COMBINED CYCLE PLANTS AVAILABILITY

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One of the most important attributes of the Integrated Gasification Combined Cycle (IGCC) plant is the ability to assure continuous and reliable operation. This ability to supply continuously electric power close to the maximum capacity is one of the most important factors that can guarantee the return on the plant investment and give confidence to the financing institutions.

During the design stage several important decisions must be taken to ensure the realization of a highly reliable IGCC unit; selection of proven technologies, adequate provisions of redundancies, selection of equipment with demonstrated reliability in similar services, duplication of valves and key instruments and adequate spares of equipment in critical services.

Following the experience gained with the activities performed for the three Italian IGCC complex a representative case study has been developed showing the design approach to the IGCC plant.

CASE STUDY DESCRIPTION

The IGCC plant considered is composed of several units where the feedstock (heavy refinery oil, typically visbroken vacuum residue) is gasified and the gas produced (syngas) is cleaned from impurities and polluting compounds and then conditioned before being burnt in the gas turbine of a combined cycle to generate power.

The block flaw diagram (fig. 1) of the case study is attached and should be followed when reading the description reported herebelow.

The feedstock (visbroken vacuum residue) is preheated and gasified in quench type Texaco gasification reactors (two reactors and two scrubbers); the syngas produced is immediately washed with water to finalize the quenching step and mechanically remove unburned carbon and solid impurities that were present in the feedstock.

The gasification is accomplished with oxygen in the presence of steam as reaction moderator.

After quench and scrubbing, dirty water (grey water) is routed to the unburned carbon and metals and ashes recovery system and then recycled or dumped to maintain the adequate concentration of dissolved salts in the water circuit.

The scrubbed syngas is cooled down to a temperature where the concentration of water in vapour phase is adequate to perform the COS hydrolysis reaction and then routed to the reactors. After reaction the syngas is furtherly cooled down to enter the H_2S removal section. The cleaning is achieved with an amine solution that chemically capture the H_2S present in the syngas and is subsequently regenerated. The offgases from regeneration are routed to a sulphur plant for pure sulphur recovery.

The clean syngas is expanded to the pressure necessary to feed the gas turbine of the combined cycle and humidified as necessary to achieve the desired NOx production during the combustion in the gas turbine.

The combined cycle is composed of one gas turbine, one waste heat recovery boiler and one steam turbine.

An air separation unit provides the oxygen for the gasification reaction.

A multiple cells cooling tower system is used to supply cooling medium for the condenser of the steam turbine and other minor users. Other utilities units are provided for the correct operation of the plant.

An economically reasonable capacity for the IGCC plant as described is 250÷300 MW gross power output.

DEFINITIONS AND METHODOLOGY

The terms and definitions used in this study analysis make reference to EPRI definitions.

A brief list of the most important definitions is reported herebelow.

Availability is the probability that a plant, component or other element is in operating conditions at any given time. It can be expressed as available hours (AH) divided period hours (PH).

Equivalent Availability is the ratio of the equivalent, at 100% capacity, hours of operation and the period hours. Empirically is equivalent to actual energy producted during the period hours and energy which could have been produced by the plant during the period hours if operating all time at full (100%) capacity.

Reliability is the probability that a plant, component, or other element, will operate satisfactorily for a given period of time.

Operating Reliability is the ratio of the equivalent, @ 100% capacity, hours of operation and the period of time that scheduled maintainance is not being performed.

Mean Time Between Failure (MTBF) is the interval of time, a component operates without interruption.

Mean Time to Repair (MTTR) is the time required to repair **a** component after **a** forced outage.

Both, MTBF and MTTR, are the average values of statistical data collected over periods of several years. Consequently all the availability - reliability parameters, developed for a plant on the base of MTTR and MTBF of single component, have a similar meaning; the probability of recording these values in an operating unit increases with the number of years of operation.

The methodology of availability assessment for a plant includes:

a. A definition of the programmed maintenance schedule to establish the downtime required to maintain the equipment of the plant in an optimum status. A good and well organized maintenance program is the basis to improve the plant reliability and achieve an optimum overall availability

b. An assessment of the forced downtime periods for a plant.

In this paper the availability assessment for the described IGCC complex is discussed, indicating which are the most important steps to be followed and which are the precautions to be adopted in order to assure that the data used in the analysis are applicable to an IGCC.

PROGRAMMED MAINTENANCE SCHEDULE

The overall IGCC complex maintenance programme scheduled for the first nine years of operation is summarized in the attached table. (Table 1)

The table shows for the most important equipment the expected maintainance periods expressed in hours over the first nine years of life of the IGCC plant.

A proposal of maintainance interventions is made with the aim to minimize plant shutdowns grouping the single equipment shutdowns within the time frame foreseen for the equipment with the most extended maintainance intervention. This means that maintainance people should be organized to act on several equipment simoultaneously.

Downtime hours are indicated for each maintainance period and the loss of plant capacity is shown in terms of percent of load; equivalent outage hours are calculated multiplying the shutdown hours and the fraction of power plant load lost.

The maintenance program is based on Vendors information for licensed units, packages and main items and on FWI experience for other items such as exchangers, pumps and control valves.

Herebelow a list of the most important considerations that have been done in order to prepare the attached maintenance plan.

As can be easily deducted from the table, the gas turbine, the steam turbine, the quench gasifiers and the expander only have an impact in determining the equivalent planned maintenance period. All other equipment maintenance has been scheduled within the time frame allowed by the time dedicated to major equipment maintenance.

The maintenance program has been prepared with the idea to minimize the overall plant downtime; every three years is programmed approx. one month of complete shutdown where the overall plant is maintained. However within the three years, in case of gas turbine and steam turbine inspection a general shutdown is required but limited in time.

Gasification Unit - Carbon Recovery - Water Treatment

Quench gasifiers

The main items of these Units to be considered are the quench gasifiers. For the three critical components of the gasifiers the following frequency and time of maintenance is recommended:

Component	**Frequency**	**Time required for intervention, hours**
Burner	every six months	6÷10
Quench ring	every two years	156÷180
Refractory	every three years	680÷760

The maintenance of the burner and of the quench ring will be performed separately for each gasifiers, while the replacement of the refractory will be carried out for the two gasifiers during the general shutdown period foreseen every three years. The layout of the structure holding the reactors, the lifting facilities and the drop areas, shall be studied in order to allow the maintenance operations concerning one reactor while the other one is in operation.

Exchangers

Maintenance of ammonia stripper reboiler will be planned during the IGCC general shutdown every three years. It is good practice to provide a connection for direct injection of steam in order to isolate reboiler without shutting down the unit (and consequently the IGCC plant), should more frequent cleaning operations be required.

The same feature should be applied to the Sour Water Stripper Unit.

Several exchangers on dirty service are present in the Unit; all of them should be spared particularly when handling the feedstock or arranged in order to minimize the plant capacity reduction when a shell is put out of service for cleaning operations. Exchangers with a new design particularly suited for viscous fluids are available on the market but they are not yet fully proven.

Syngas Heat Recoverv and COS Hydrolysis Units

The critical item of this unit is the hydrolysis reactor requiring catalyst replacement. The catalyst replacement will be every three years during the IGCC general shutdown (the expected life is over four years).

To improve the lifetime of COS hydrolysis catalyst, two guard reactors having a minimum catalyst holding capacity shall be installed upstream the hydrolysis reactors; one guard reactor is in stand by while the other is in maintainance or operating and the switchover of the main stream from one to the other is done without plant shutdown.

Expansion and Saturation Unit

The following maintenance schedule have been planned for the expander:

- major maintenance and inspection every three years, duration two weeks
- complete inspection of generators every six years, duration three weeks

The above mentioned program can be achieved only if the expander is selected with stringent requirements (typically API). Minor maintainances around the equipment should be performed with the unit in operation.

The maintenance of the other components of the unit can be carried out during the IGCC complex scheduled shutdown periods.

Acid Gas Removal Unit

General maintenance requirements are:

- change out of cartridge filter elements (estimated time is 6 hours every 2÷6 months); this operation can be carried out leaving the unit operating at 100% capacity
- other items requiring periodic maintenance, such as hydraulic turbines, regenerator reboiler and plate heat exchangers, are fully spared and therefore can be maintained without shutting down or decreasing the Unit capacity.

Sulphur Recovery and Tail Gas Treatment Units

The sulphur recovery system is generally composed of two lines operating in parallel each covering 100% of design capacity. Therefore the scheduled maintenance of one line can be carried out while the other line is operating, without decreasing the overall capacity of the IGCC Complex.

Each line requires a 15 day period each year for general maintenance.

Combined Cycle Unit

The main items involved in the maintenance program are the Gas Turbines, the Heat Recovery Steam Generator and the Steam Turbine.

Gas turbine and associated generator

The maintenance program of the gas turbine is summarized herebelow:

Year of operation	Type of intervention	duration, hours
first	combustor inspection	72
	combustor overhaul	96
second	combustor overhaul	96
third	hot part inspection	504
fourth	combustor overhaul	96
fifth	combustor overhaul	96
sixth	major overhaul	676
	combustor inspection	72
seventh	combustor overhaul	96
eighth	combustor overhaul	96
ninth	hot part inspection	504

The above maintainance program is similar to the one foreseen for a gas turbine burning natural gas; it is recommended the gas turbine manufacturer demonstrates confidence in burning syngas.

The maintenance of the generator can be performed during the above schedule foreseen for the associated gas turbine.

Heat Recovery Steam Generator

The maintenance of critical components can be carried out during the maintenance of the gas turbine. If salts depositions are expected in the HRSG cold section the exchanging surfaces shall be provided with fins with increased spacing and sootblowing facilities be installed.

Steam Turbine

The maintenance program is summarized in the following table.

Type of intervention	Equivalent ST operating time	Duration weeks
minor overhaul	10000	2
minor overhaul	25000	2
major overhaul	50000	4

Auxiliary units

The maintenance schedule for auxiliary units like cooling water system, plant and instrument air system, flare system, etc.... will not affect the overall plant downtime, in fact they should adequately provided with spare equipment in order to guarantee always continuous operation capability.

On the basis of the assumptions made and of the design features introduced in the IGCC plant it is possible to calculate an equivalent maintainance outage time of 352 hours per year for the first twenty years of plant operation. The loss of electric power production related to the maintainance program is equivalent to 352 hours of the IGCC plant running at 100% capacity.

UNPLANNED OUTAGE EVALUATION

The evaluation of the IGCC Complex unplanned outages can be performed through an availability assessment.

To accomplish this evaluation, FWI normally uses the UNIRAM methodology, developed for EPRI and conceived to consider partial outages as well as full outages. Thus the IGCC Complex is not assumed to be always 100% available or 10C% unavailable (binary operation) but is assumed to have several intermediate levels of power output capability, depending on the operating condition of the complex.

Each operating condition of the IGCC complex, called an IGCC complex state, has an associated capability.

Each state is determined by evaluating which of the components of the complex are available for operation and which are not available because of failures (i.e. unplanned outages).

This approach uses the reliability/maintainability data of each component of the IGCC Complex to predict the IGCC transitions from one state and relevant capability to another one, producing time-varying prediction of the operating conditions of the complex.

The main elements of the availability assessment methodology are shown in the Figure 2.

To define the operating states and relevant capabilities of the IGCC complex a dedicated scheme (called availability block diagram (ABD)) giving the functional configuration of the plant from the power production point of view, must be built from the IGCC complex process scheme.

Fig. 3 attached (Availability assessment logical structure) depicts the ABD for the case study. All units of the IGCC complex are reported with indicated the percentage of the power that is available from the complex when they are in operation. Each unit is splitted in equipment following the same approach; in the figure 3 the combined cycle block is blown up showing its relevant equipment and their influence on power production.

Blocks connected by lines represents components of the plant and from the availability point of view, behave as logic switches: if the block is in a non operational state, the effect on the logic chain is to bring the whole plant to a state of reduced capacity or non operation, depending on the fraction of the power production supported by the blocks and on their connection along the chain.

With the IGCC Complex ABD and a complete data base, including Availability, Mean Time Between Failures (MTBF) and Mean Time to Repair (MTTR) for the IGCC Complex components, it is possible to evaluate quantitatively the IGCC Complex unplanned outages.

The duration of these outages can be minimized with a suitable design decision after highlighting the critical item from the plant availability point of view.

In the following are summarized typical values of availability for main units of the IGCC Complex.

Gasification Units

The point-estimate availability of each gasifier/scrubber deriving directly from Texaco information is 99.1%.

Air Separation Unit

The point-estimate availability of this unit is equal to 98.0%.

This value has been reached by means of suitable modification to the unit configuration originally proposed by its vendor, e.g. the addition of a liquid oxygen storage vessel which capacity is a compromise between the cost and increase of expected availability.

Carbon Recovery

The high availability of this unit, equal to 99.8%, has been reached with suitable improvements to the conventional design, like fully redundant exchanger and double control valve when dirty service is expected.

Water Treatment

Buffer tanks have been added to reach a point-estimate availability of 99.3%.

Syngas Cooling and COS Hydrolysis

The value of the availability of this unit is 99.0%.

Syngas Expander

The availability of the syngas expander, specified according to API Standards, reaches the value of 99.8%.

Sulphur Removal Unit

The availability of this unit is equal to 99.9%. The equipment sparing as described for the maintainance program highly contributes to the increase of the availability figure.

Sulphur Recovery and Tail Gas Treatment

The calculated availability of each unit is:

- Sulphur recovery: 99.8 %
- TGTunit: 99.3%

Following refineries operating experiences, the best IGCC Complex availability can be reached by adopting a fully redundant configuration for both units in spite of the relatively high values of the calculated availabilities.

Combined Cycle Unit

The combined cycle can operate at different derated power production capabilities due to unplanned outages of its main components.

The percentages of time during which the Combined Cycle operates at different loads is:

Load (%)	Percentage of time (%)
100	96.56
84	0.15
61.6	0.19
0	3.10

Auxiliary Units

The availabilities of the auxiliary units are:

- Cooling water: 99.98 %
- Instrument/Plant Air: 99.9 %
- Electric System: 99.7 %
- Control System: 99.7 %

In Table 2 attached, a set of reliability and maintainability data for the main components of the plant used in the preparation of this case study assessment are reported.

CASE STUDY AVAILABILITY RESULTS

The basic measure of availability obtained from the methodology explained in preceeding paragraph is the IGCC complex operating reliability which reflects the impact of unplanned outages and partial unplanned outages.

Operating reliability is similar to equivalent availability, except that equivalent availability accounts for planned outages for maintenance as well.

The operating reliability measure can be combined with the planned outages for maintenance to obtain prediction of plant availability and equivalent availability.

Table 3 shows the expected percentage of time the plant is available at various powers (states).

The IGCC Complex availability performances for the studied case are:

Operating Reliability:	91.37%
Availability:	88.84%
Equivalent Availability:	87.7 %

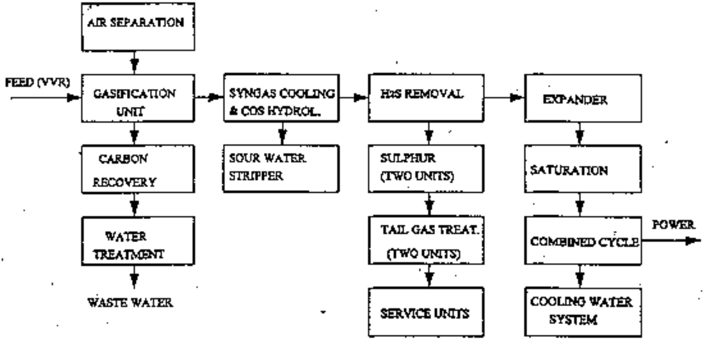
CONCLUSIONS

It is writers opinion that the assessment methodology proposed is the best tool to predict the availability performances of a plant during design stage.

To improve the availability performances it is possible to act in few direction only, i.e.:

- to increase the quality of the equipment purchased (decrease MTBF)
- to put more redundancies in the various systems (increase availability), but be cautious not to increase excessively the plant cost composed with the overall availability increase achieved.
- to improve the maintenance program and the logistic for spare equipment supply (MTTR and maintenance schedule)

Fig. 1 - IGCC Complex Block Flow Diagram



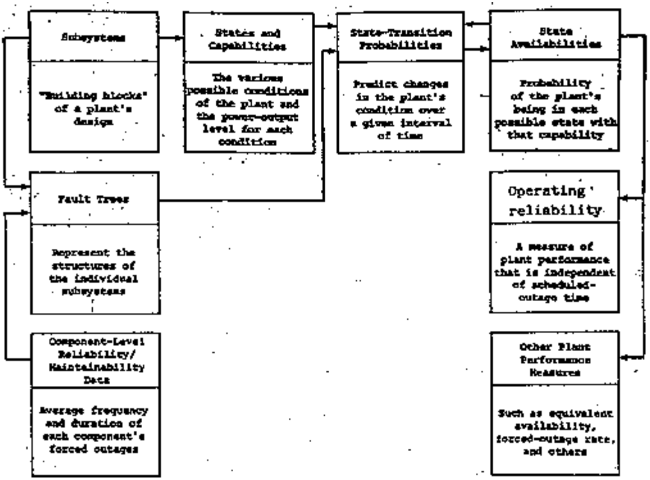


FIG. 2 - MAIN ELEMENTS OF THE AVAILABILITY ASSESSMENT TECHNOLOGY

FIG. 3 - AVAILABILITY ASSESSMENT LOGICAL STRUCTURE

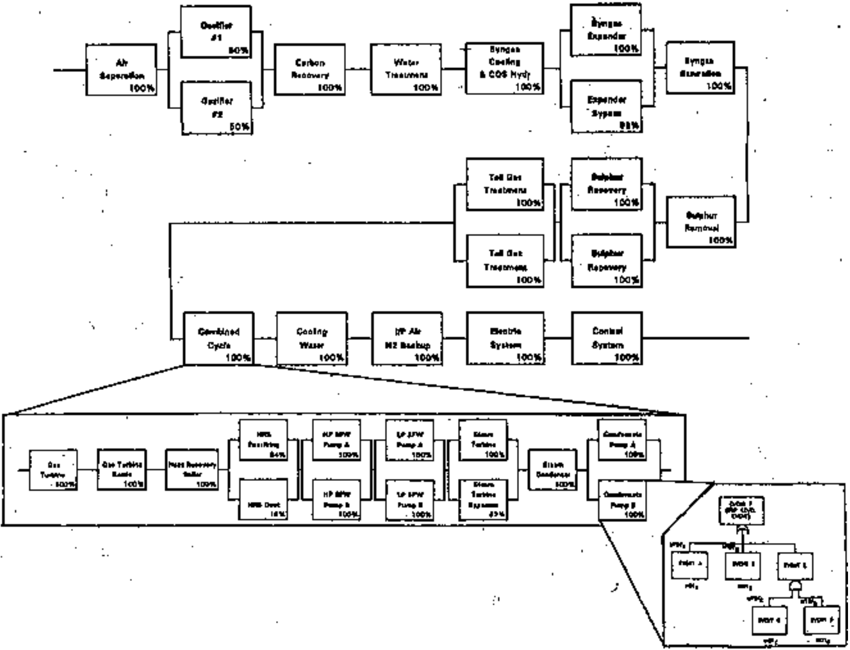


Table 1. Overall IGCC Complex maintainance program acheduled (or tha first nina years

equipment	years 1st				2nd							3rd		
Gas turbine	72			96						96				804
Steam turbine				336										336
Gasifier	8			8	8			168				8		720
Gasifier	8			8		8				96	72		8	720
Expander														336
Sulphur unit			24	336			192							360
Sulphur unit		360							192					360
Downtime hours	72	360	24	336	8	8	192	168	192	96	72	8	8	720
% capacity lost	100	0	0	100	50	50	0	50	0	100	50	50	50	100
Equivalent outage hours	72	0	0	336	4	4	0	84	0	96	36	4	4	720
Yearly eq.outage hours	·			408							224			728

equipment	4th							5th					6th		
Gas turbine						96						96			744
Steam turbine															672
Gasifier	8			168				8				8	8		720
Gasifier		8				96	72			8		8		8	720
Expander															804
Sulphur unit			182						360						360
Sulphur unit					192						264				360
Downtime hours	8	8	192	168	192	96	72	8	360	8	264	96	8	8	744
% capacity lost	50	50	0	50	0	100	50	50	0	50	0	100	50	50	100
Equivalent outage hours	4	4	0	84	0	96	36	4	0	4	0	96	4	4	744
Yearly eq.outage hours							224					104			752

equipment	7th					8th							9th		
Gas turbine					96						96				804
Steam turbine															336
Gasifier	8				8	8			168				8		720
Gasifier			8		8		8				96	72		8	720
Expander															336
Sulphur unit		360						192							360
Sulphur unit				264						192					360
Downtime hours	8	360	8	264	96	8	8	192	168	192	96	72	8	8	720
% capacity lost	50	0	50	0	100	50	50	0	50	0	100	50	50	50	100
Equivalent outage hours	4	0	4	0	96	4	4	0	84	0	96	36	4	4	720
Yearly eq.outage hours					104							224			728

note:. maintainance program from year 4 to yaar 9 can be repeated from yaar 10 to yaar 15 and from yaar 16 to yaar 21 and so on with a cyclic pariod of 6 years.

. avaragad on twanty yaars of operation the equivalent planned outage hours are 352

Table 3 - Expected Availability Performaneea

Period Hours (PH): 8760

Equivalent Planned Outage Hours (EPOH): 352

Operating State	State Probability [%]	Plant Capacity [%]	Operating Hours [h/y]	Equivalent Hours [h/y]	Equiv. Outage Hours [h/y]
1	89,941	100,0	7562,2	7562,2	0,0
2	0,1416	98,0	11,9	11,7	0,2
3	0,0002	84,0	0,0	0,0	0,0
4	0,2311	75,0	19,4	14,6	4,9
5	0,0007	61,6	0,1	0,0	0,0
6	2,2453	50,0	188,8	94,4	94,4
7	0,0007	25,0	0,1	0,0	0,0
Total per year			7782,6	7682,8	99,6
5	7,4391	0,0	625,5	(FOH)	
Total hours per year			8408,0		
6	·	0,0	352,0	(EPOH)	
Total hours per year			8760,0		

TABLE 2 - EQUIPMENT RAM DATA

Class	Type	MTBF (h)	MTBF Range (h)	MTTR (h)	MTTR Range (h)	Availability (%)
Blowers	All	87600	29200 - 292000	16	12 - 20	0.99982
Burners	WHB postfiring	29200	-	36	-	0.99877
Bypasses	Turbine/expander bypass	176200	87600 - 876000	10	8 - 16	0.99994
Control Loops	With valve bypass					
	- Dirty fluid	87600	-	12	-	0.99986
	- Clean fluid	8760000	-	12	-	0.999999
	Without valve bypass					
	- Dirty fluid	1750	-	12	-	0.89319
	- Clean fluid	8760	-	12	-	0.99863
Compressors	Centrifugal	5850	2500 - 17500	12	8 - 24	0.99795
	Reciprocating	2900	1750 - 17500	16	12 - 32	0.99451
Condensers	Water cooled					
	- Generic vapour	16000	10000 - 87600	10	8 - 24	0.99938
	- Steam	47400	-	24	-	0.99949
	Air cooled					
	- (see exchangers and fans)		-	-	-	-
Demineralizers	All	8760000	-	8	-	0.999999
Drums	Generic	876000	292000 - 2920000	8	6 - 18	0.99999
	WHB steam drums	49000	40000 - 55000	10	5 - 15	0.99979
Exchangers	Shell & tube					
	- Generic	97500	-	16	-	0.99984
	- Gas/gas	43900	-	16	-	0.99963
	- Gas/water	11000	-	16	-	0.99855
	- Water/water	43800	-	16	-	0.99963
	- Air/water	438000	-	16	-	0.99996
	- Oil/gas	11000	-	16	-	0.99855
	- Oil/water	17500	-	16	-	0.99809
	- Oil/oil	438000	-	16	-	0.99996
	Tube bundle	58400	8760 - 438000	16	8 - 24	0.99973
	Reboiler	2500	-	16	-	0.99404
	Plate	8760	-	36	-	0.99591
Fans	Axial	17500	4380 - 8760	16	12 -24	0.999087
Filters	Generic	876000	176200 - 8760000	10	8 -16	0.99999
Pumps	Centrifugal					
	- Each pump	5850	3500 - 35000	24	12 - 48	0.99591
	- Two pumps fully redundant	584000	-	12	-	0.99998
	Reciprocating	5850	3500 - 35000	24	12 - 48	0.99591
Reactors	Generic	175200	87600 - 438000	24	12 - 48	0.99996
Tanks	Generic	8760000	-	20	-	0.999998
Towers	Cooling water (seawater)	438000	-	12	-	0.999973
Turbines	Gas turbine	2400	-	48	-	0.980392
	Steam turbine	11000	-	22	-	0.999004

EFFICIENT AND RELIABLE OIL-BASED IGCC PLANT CONCEPT COMBINING PROVEN TECHNOLOGICAL COMPONENTS: ADVANCED BURNER TECHNOLOGY, WASTE HEAT RECOVERY, SOOT PROCESSING

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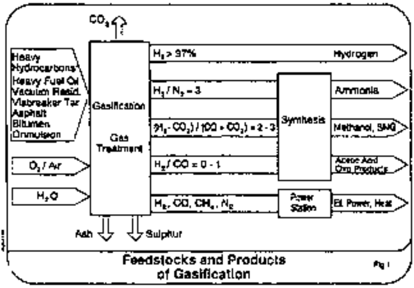
INTRODUCTORY REMARKS

Feedstocks and products

The gasification of liquid and solid fuels is a conversion process which offers a variety of applications. In recent years, the prospect of using heavy hydrocarbons as feedstocks for gasification processes has become increasingly attractive. This feedstock group includes heavy fuel oil and heavy residues from oil refining (such as vacuum residue, visbreaker tar and asphalt) as well as natural bitumen (see Fig. 1). In this context, Orimulsion™ , an emulsion containing Orinoco Bitumen and water, deserves special mention. It is offered on the world market at reasonable terms. The essential constituents of a typical heavy hydrocarbon feedstock are carbon (approx. 85 % by wt), hydrogen (approx. 10 %), sulphur of a high concentration (3 - 5 %) and a certain amount of ash in the order of 0.1 to 0.3 %. The ash usually has a high content of heavy metals, in particular vanadium and nickel. The vanadium concentration of Orinoco Bitumen can be as high as 300 to 500 ppm, visbreaker tars may have vanadium and nickel contents of up to 800 ppm and more.



Feedstocks and Products of Gasification

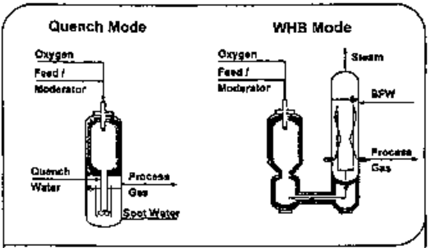
Fig 1

The main components of the crude gas from the gasification of these feedstocks are CO and H_2 in a molar ratio of approx. 1.1. In downstream facilities, the crude gas is conditioned so as to meet the requirements of the final products which are summarised in Fig. 1: pure hydrogen, the group of synthesis products and fuel gas for power and heat generation. The principle of combined cycle power generation with integrated oil gasification (IGCC) offers decisive advantages compared to a classic steam-turbine power station: The overall efficiency is increased and gas purification under pressure is more efficient, which results in drastically reduced emission levels comparable to those of a natural-gas-based power station.

Modes of gasification

A highly reliable and flexible partial oxidation process for the conversion of the fuels in question is the Texaco Gasification Process entailing a long record of experience. Approx. 100 plants based on this process have been constructed worldwide, using a variety of gaseous, liquid and solid feedstocks. The fuel is fed to a process burner where it is mixed with the oxidant and gasified at a high pressure and a temperature in the order of 1400 °C. A certain quantity of soot is obtained as a gasification byproduct mainly consisting of unconverted carbon and ash (slagged).

Fig. 2 shows two different gasification schemes: in the direct quench mode, the hot crude gas from the gasifier is cooled by direct contact with evaporating water. This method is very simple and inexpensive. However, it is much more efficient to utilise the sensible heat of the crude gas to generate HP steam in a downstream waste heat boiler (WHB mode). When a waste heat boiler is provided, the concentration of heavy metals (vanadium) in the crude gas should not exceed certain limits in view of the adverse effect this would have on the service life of the steam boiler. It is obvious that burner performance, soot formation and the method of soot handling determine the method of crude gas cooling to be selected and hence influence the efficiency of the overall process. These aspects will therefore be given special attention.



Gasification of Liquid Feedstocks
TEXACO Process

Fig. 1

OIL-BASED IGCC PLANT CONCEPT

Process configuration

Based on the practical experience gained over many years, a process concept applying the IGCC principle has been developed for the generation of electric power from heavy hydrocarbon feedstocks,. High efficiency and operational reliability is to be expected when using this concept which is characterised by three important elements:

- Advanced burner technology yielding a high carbon conversion rate.
- Crude gas cooler for the production of high-pressure steam.
- Soot separation by filtration and further processing of the soot to recover valuable metals, i.e. vanadium and nickel.

These process elements are described in more detail below.

The process configuration is illustrated in the simplified block flow diagram (Fig. 3): Heavy hydrocarbon feedstocks are gasified with the aid of oxygen supplied by an air separation unit (ASU). Compressed surplus air from the compressor of the gas turbine unit can be used as feed air for the ASU. This integrated concept offers certain advantages resulting in overall energy savings. The crude gas is cooled in a waste heat boiler producing high pressure steam. To increase the overall efficiency, a downstream medium-pressure steam generator can be provided. The particulates (soot) are separated from the gas by wet scrubbing and are then contained in the soot water. The dust-free process gas is cooled further and subsequently undergoes H_2S removal, the sour gas being treated in the sulphur recovery unit to produce elemental sulphur. Prior to admission to the combustion chamber of the gas turbine unit, the conditioned fuel gas is humidified, utilising process heat at a low temperature level. This measure counteracts the formation of NO_x and increases the capacity of the gas turbine set Nitrogen from the ASU can be admixed to the fuel gas to minimise NO_x formation. The fuel gas is then subjected to combustion and is expanded in the gas turbine to produce electric power. The exhaust gas from the gas turbine has a high heat content which is utilised in the downstream heat recovery steam generator to produce HP and MP steam. This steam, along with the steam produced in the crude gas coolers of the gasification section, is expanded in a steam turbine, thus generating additional electricity.

As Fig. 4 shows, the overall efficiency is favourable for all the feedstocks considered if a waste heat boiler is arranged downstream of the gasifier. If the waste heat boiler is replaced by a direct quench, the overall efficiency will considerably decrease by approx 5 to 6 percentage points since only steam of moderate pressure can be produced in the gasification section. This comparison shows that the incorporation of a crude gas cooler for HP steam generation is a decisive advantage.

The studies are based on a gas turbine V94.3 of Siemens KWU, which results in an overall net power output of the IGCC power station in the order of 300 MW.

SPECIAL PROCESS ELEMENTS

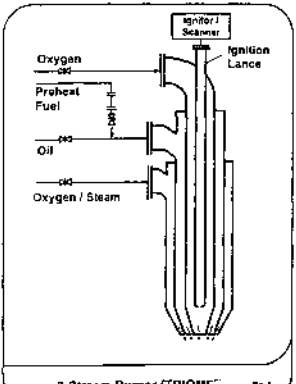
Technical basis

The process concept described above is based on special experience gained during the operation of coal and oil gasification plants. For more than twenty years Uhde, in its capacity of engineering partner to Ruhrchemie and Ruhrkohle, has played an active part in the development of Texaco Coal Gasification technology. A demonstration plant was designed, built and operated in Oberhausen with great success from 1978 until the end of 1985. The technical solutions found and the results obtained in this demonstration unit formed the basis for the design of the subsequently built commercial-scale plants. As the general contractor, Uhde designed, constructed and commissioned (in 1986) the SAR (Synthesegasanlage Ruhr) coal gasification plant in Oberhausen which produces hydrogen, oxo syngas (capacity: 50,000 m^3/h STP of $H_2 + CO$ in total) and food-quality CO_2. For economic reasons, the plant was modified 1991 and has been using heavy oil residues as a feedstock since then.

The first basic engineering package for Texaco-based gasification of Orimulsion was prepred by Uhde in 1987/89 based on pilot plant tests in Texaco's Montebello research facilities.

Several process components originally developed for coal gasification have been applied for oil gasification with great success. In this context, the following items deserve special mention.

Process Burner

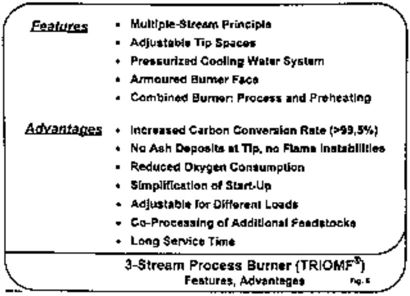


3-Stream-Burner (TRIOMF™) Fig. 5

In the early phase of process development in the Oberhausen demonstration plant mentioned above, a conventional two-stream process burner was used which showed a poor performance in coal operation, resulting in insufficient carbon conversion. To overcome these difficulties, an advanced burner type was developed in close cooperation of the plant owners with their engineering partner.

This three-stream burner successfully replaced the conventional two-stream burner. It was further improved and optimised in the commercial-scale SAR coal gasification plant After conversion of the plant in order to be able to use oil feedstock, the three-stream-burner was adapted to meet oil gasification conditions and showed an excellent performance which even exceeded all expectations. The carbon conversion rate was increased from 98 % (typical for a two-stream burner) to more than 99.5 %, with the result of a drastically reduced soot formation, thus obviating the need for soot recyclling to the gasifier to improve energy efficiency.

The sketch (Fig. 5) illustrates the burner configuration: a certain portion of the oxidant is fed to the centre of the burner whereas the remaining portion is injected via the outer annular space. If required, provision can be made for the injection of additional feedstocks via a separate central lance. Waste water contaminated with hydrocarbons has been co-processed successfully in this manner.



Features
- Multiple-Stream Principle
- Adjustable Tip Spaces
- Pressurized Cooling Water System
- Armoured Burner Face
- Combined Burner: Process and Preheating

Advantages
- Increased Carbon Conversion Rate (>99,5%)
- No Ash Deposits at Tip, no Flame Instabilities
- Reduced Oxygen Consumption
- Simplification of Start-Up
- Adjustable for Different Loads
- Co-Processing of Additional Feedstocks
- Long Service Time

3-Stream Process Burner (TRIOMF®)
Features, Advantages Fig. 6

Further technical features of the burner are summarised in the table (Fig. 6). The tip spaces can be adjusted so as to adapt the burner to different load conditions and to enable the use of the burner for preheating the refractory lining, thus simplifying the start-up procedure. The burner tip is cooled by means of a pressurised cooling water cycle, the burner face is armoured for protection. A great deal of experience as well as manufacturing know how is required to select the most suitable material, to design the burner tip and to meet the requirements regarding tight tolerances and durable attachment of the protective liner.

Soot handling

The soot formed during gasification is removed from the crude gas by scrubbing with water. In subsequent process steps, the soot is separated from the water phase and can be treated further to permit recycling to the gasifier. This is usually accomplished by carbon/naphtha extraction. An attractive alternative process route for soot water treatment is filtration, a method derived from char water handling and applied in coal gasification technology.

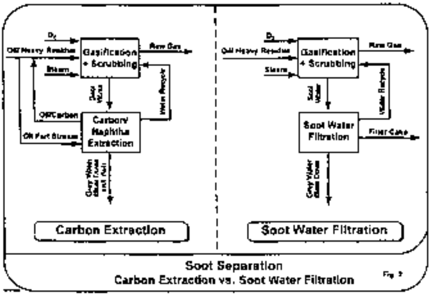
The sketch (Fig. 7) illustrates the process configuration for carbon extraction and soot water filtration. In the conventional process (left-hand side), the soot water is mixed with naphtha which acts as an extractant In a decanter, the soot/naphtha mixture is separated from the water phase which includes almost the entire ash portion. The carbon-bearing naphtha is brought into contact with a partstream of the oil feedstock. This mixture is then treated in a naphtha stripper where the naphtha is recovered and returned to be mixed with soot water again. The extracted carbon along with the feed oil partstream is recycled to the gasifier.

In contrast to this method, the alternative treatment concept (right-hand side) is very simple: The soot water is passed through a filter press and a filter cake is produced which (as was found) contains virtually all heavy metals and thus represents an excellent metal recovery source. The soot is not recycled to the gasifier. With this concept the vanadium concentration in the process section is kept below the critical level.

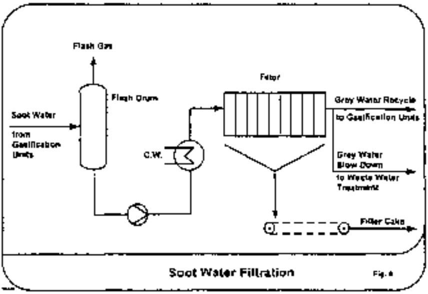
The soot water filtration concept has the following main advantages: favourable capital investment cost low energy consumption, no naphtha consumption and handling, simplified plant operation, high operational reliability

In order to minimise the amount of carbon black, a high carbon conversion rate is a fundamental prerequisite for the implementation of the soot water filtration concept The special three-stream process burner (TRIOMF® Burner) described earlier meets these requirements.

The simplified process flow diagram (Fig. 8) shows the process configuration of the soot water filtration unit: The soot water from the gasification unit is expanded and stripped with LP steam in the flash drum in order to remove any dissolved gases. Carbon water pumps withdraw the hot soot water from the flash drum and convey it via the filter feed cooler to the filter where the mixture is dewatered to a solids content in the order of 20 % by weight Two filters are arranged in parallel: One is in operation while the other is being cleaned. A quasi-continuous operation is achieved by automatically switching from one filter to the other after completion of each filter cycle. The filter cycle ends when the time has elapsed, on account of a high differential pressure or by a manual stop initiated by the operator. The filter cake is transported away by a conveyor. The filtrate from the filter is pumped back as make-up grey water to the gasification units. A blow-down partstream of the filtrate is fed to the waste water treatment unit



Soot Separation
Carbon Extraction vs. Soot Water Filtration
Fig. 7



Soot Water Filtration
Fig. 8

Since early 1992, a soot water filtration unit working by the process described has been in operation in conjunction with the oil gasification unit in the SAR plant at Oberhausen, Germany, giving excellent results.

Waste heat recovery

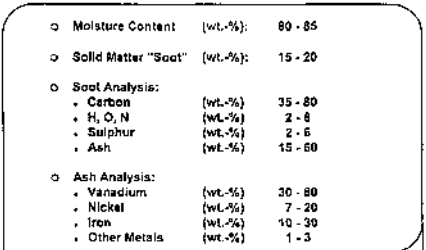
The utilisation of the process heat downstream of the gasifier for the production of high-pressure steam has a very favourable effect on the overall thermal efficiency of the IGCC power station. The application of a waste heat boiler for crude gas cooling, however, strongly" depends on the concentration of heavy metal compounds in the gas. A high concentration of heavy metals, vanadium in particular, may adversely affect plant operation in view of the risk of erosion, corrosion and deposits in the facilities downstream of the gasifier. Application of a crude gas cooler in conjunction with soot recycling is therefore not recommended. As practical experience has shown, the inlet section to the tube coils of the waste heat boiler is particularly critical due to high heat flux and its direct exposure to heavy metals attack will result in a limited service life.

By applying a three-stream burner with a high carbon conversion rate along with a soot filtration step, recycling of soot is no longer required. Thus the metals build-up in the gasification and water recycling systems is minimised. During plant operation by this method, neither indications for vanadium-induced erosion and corrosion nor ash deposits were found in the crude gas cooler and the other equipment downstream of the gasifier. The wear rate is very low, resulting in a considerably extended service life of the heat exchanger tube coils and their inlet sections.

RECOVERY OF HEAVY METALS FROM SOOT

Fundamentals

The table (Fig. 9) summarises characteristic data of the filter cake obtained through soot filtration as described earlier. The filter cake has a high moisture content in the order of 80 to 85 %. The solid matter is mainly made up of carbon and ash containing heavy metals in varying concentrations depending on the origin of the gasifier feedstock. Thus the filter cake is an excellent source for the recovery of valuable metals, such as vanadium and nickel. A proprietary process (CASH$^{®}$ technology) has been developed to produce an ash concentrate from soot filter cake from which these metals can be reclaimed in a conventional metallurgical process. This puts high demands on the quality of the metal ash produced as regards the residual carbon and sulphur contents, for instance.



○ Moisture Content	(wt.-%):	80 - 85
○ Solid Matter "Soot"	(wt.-%):	15 - 20
○ Soot Analysis:		
. Carbon	(wt.-%)	35 - 80
. H, O, N	(wt.-%)	2 - 8
. Sulphur	(wt.-%)	2 - 6
. Ash	(wt.-%)	15 - 60
○ Ash Analysis:		
. Vanadium	(wt.-%)	30 - 80
. Nickel	(wt.-%)	7 - 20
. Iron	(wt.-%)	10 - 30
. Other Metals	(wt.-%)	1 - 3

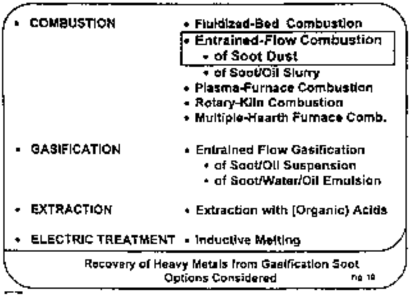
Typical Filter Cake Characteristics
Feedstock: Heavy Residue

Fig. 9

During the development of the process for the production of an ash concentrate of adequate quality, various process options were considered (see Fig. 10). These included different methods for combustion and gasification of the soot as well as extraction with acids and inductive melting. As regards the selection and design of the most appropriate treatment method, the following factors are of utmost importance: The moisture content and mechanical properties of the filter cake, reactivity of the soot and the ash melting behaviour which, in particular, strongly depends on whether certain vanadium oxides (V_2O_5) might be formed.

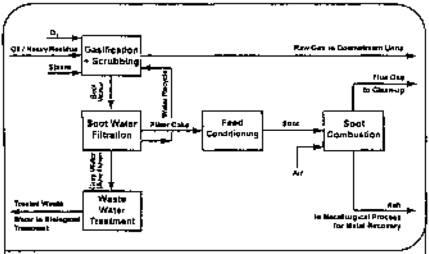
After comprehensive experimental investigations and technical/economic comparisons of the options considered, the process was based on the principle of entrained-flow combustion of soot dust The concept mainly comprises the following consecutive steps (see Fig. 11):

- Soot water filtration method to produce a filter cake containing the heavy metals.
- Soot conditioning to convert the soot into a product suitable for subsequent combustion.
- Soot combustion to obtain concentrated ash for metal recovery.



- **COMBUSTION**
  - Fluidized-Bed Combustion
  - Entrained-Flow Combustion
    - of Soot Dust
    - of Soot/Oil Slurry
  - Plasma-Furnace Combustion
  - Rotary-Kiln Combustion
  - Multiple-Hearth Furnace Comb.
- **GASIFICATION**
  - Entrained Flow Gasification
    - of Soot/Oil Suspension
    - of Soot/Water/Oil Emulsion
- **EXTRACTION**
  - Extraction with (Organic) Acids
- **ELECTRIC TREATMENT**
  - Inductive Melting

Recovery of Heavy Metals from Gasification Soot
Options Considered
Fig. 10



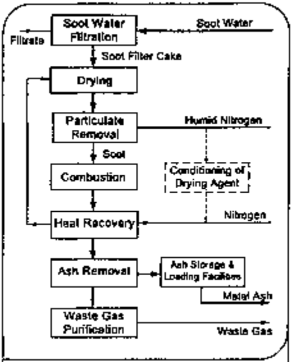
Recovery of Heavy Metals
Soot Processing by Combustion
Fig. 11

Process configuration

The block flow diagram (Fig. 12) shows the process sequence in more detail: the soot water filtration step has already been described. In order to incinerate the soot in the selected combustion system, the soot filter cake has to be dried to a low residual moisture content In addition, the dried soot has to be conditioned in such a way that a certain grain size distribution is achieved. The proper conditioning is accomplished in the drying unit by applying proven processes of combined drying and pulverisation.

Inert gas is used as the drying agent in order to exclude the risk of dust explosion. Usually nitrogen will be available as a by-product from air separation facilities supplying the oxygen for the gasification unit

In the particulate removal unit, the entrained dried soot is separated from the vapours/inert gas mixture by cyclones and filters. The humid nitrogen is discharged to the atmosphere or recycled after condensation of the vapours (see above). The dried soot produced is collected in a silo blanketed with inert gas.



Recovery of Heavy Metals from Soot
CASH® Technology Fig. 12

A patented method is applied for the combustion of the dried soot dust: the rotary combustion chamber system Loesche-Brinkmann. The soot dust is metered and tangentially injected into the combustion chamber via different nozzles by pneumatic transport In view of the high portion of vanadium and nickel as well as the presence of trace contaminants such as sulphur in the soot feedstock, particular combustion conditions are observed and the materials of construction of the combustion chamber and the downstream cooling system have to satisfy special demands.

Oxygen supply, combustion temperature and residence time are controlled and adjusted so as to exclude the formation of vanadium oxide compounds which have a low melting temperature, on the one hand, and to ensure complete combustion of the carbon, on the other hand. As a result a powdery dry ash containing vanadium and nickel in a concentrated form is obtained. The carbon conversion rate achieved in the combustion chamber is distinctly above 99 %.

The sensible heat of the hot flue gas from soot combustion is utilised in a gas cooler for heating the drying agent i.e. nitrogen, in the heat recovery section. A portion of the metal ash product is collected in the bottom section of the heat recovery system, the entrained fine metal ash being removed from the waste gas in the filter system of the ash removal section. After dust removal the waste gas has a residual dust content of less than 1 mg/m^3 (STP).

The metal ash product obtained is cooled and then conveyed to the ash storage & loading facilities where it is kept ready for being transported away. The concentrated metal ash is a valuable product which can be further treated in a conventional metallurgical process to recover vanadium and nickel.

Depending on the quality of the soot filter cake, certain concentrations of SO_x are to be expected in the dust-free waste gas from the soot combustion section. The waste gas can be fed to existing facilities (for instance, a Claus unit) or to an additional facility for desulphurisation. The concentration of halogenes and other non-atmospheric trace compounds is negligible.

Each of the process steps of the metal recovery concept described represents proven technology under identical or comparable operating conditions. The first commercial-scale rotary combustion chamber for gasifier soot which was commissioned in autumn 1996, confirmed the good performance predicted.

Plant capacity and economic aspects

Taking an Orimulsion ™ - based IGCC power plant with a net output of approx. 300 MW_{el} as a basis, the quantity of dry soot to be processed in the metal recovery section is in the order of 250 kg/h. With a typical vanadium content in the feedstock of approx. 350 ppm, the annual output of concentrated metal ash contains approx 200 t of vanadium and 501 of nickel. Since these metals can be reclaimed from the ash concentrate in a conventional metallurgical process, the application of the CASH® technology will be able to make an appropriate contribution to the profitability of the overall process. The combination of this measure with the advanced three-stream burner and soot water filtration technologies has a favourable effect on energy consumption, the capital cost and the operational reliability of the overall plant The technological components described can also be used to enhance the performance of existing gasification plants for heavy hydrocarbon feedstocks.

REFERENCES

/1/ W. Schafer, M. Trondt, J. Langhoff, W. Konkol, J. Hibbe):
.Texaco-Kohlevergasungsverfahren. Erfolgreiche Entwicklung eines Flugstrom-Vergasungsverfahrens durch Ruhrkohle AG/Ruhrchemie AG",
Erdol und Kohle-Erdgas-Petrochemie vereinigt mit Brennstoffchemie, Bd. 40, Heft 5,1967, PP. 209-216

/2/ J. Keller
.Diversification of feedstocks and products: Recent trends in the development of solid fuel gasification using the Texaco and the HTW process",
Fuel Processing Technology, 24 (1190), pp. 247-268

/3/ Texaco Development Corporation:
.Texaco Gasification Process for gaseous and Liquid Feedstocks", brochure 1993

/4/ W. Klosetal.:
.Orimulsion ™ IGCC Study",
Technical Process Information, Jan. 1995, unpublished

/5/ H. Wulfert
Product information of Loesche GmbH, 1996

START-UP OF THE FIRST COMMERCIAL SOOT ASH REMOVAL UNIT (SARU)

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Introduction

SARU was developed in 1991 jointly by Shell, Lurgi and an SGP-licensee as a new, more economic method of soot/ash removal for the Shell Gasification Process (SGP, partial oxidation of gaseous or liquid hydrocarbons). SARU is intended for processing refinery residues with high heavy metals (V, Ni) content or other applications where a soot recycle is unwelcome. The soot and ash are filtered out of the slurry using an optimized filtration method. The filter cake is then subjected to controlled oxidation in a multiple hearth furnace. The ash components are recovered as oxides, for use in the metallurgical industry.

The first commercial installation of a SARU was planned with the SGHP project at Shell's Pernis refinery near Rotterdam, where 1650 t/d of heavy residue will be gasified. This plant is scheduled for start-up in early 1997 (Ref. 1). Meanwhile this project was overtaken by a smaller, fast-track SGP-based syngas project. This plant in South Korea gasifies 144 t/d of heavy fuel oil and also employs a complete SARU. It was started successfully in October/November 1996.

The current paper describes the development of the process and its main features and then reports on the start-up and first operating experiences. With these two applications the new technology will be firmly established as SGP's preferred soot removal process.

Development of the Process

Background: Resource recovery instead of recycling

The Shell Oil Gasification Process (SGP) is a well proven process for the partial oxidation of liquid hydrocarbons [1]. This process is used to convert predominantly heavy residual oils from refineries into a raw syngas. Soot is obtained as a byproduct as the oil is reacted with steam and oxygen. This soot is removed from the gas in a two-stage scrubber together with the feedstock ashes. Depending on the feedstocks, the solids-laden scrubber effluent (carbon slurry) may also contain vanadium and nickel in addition to the soot.

Modern processes for treating this carbon slurry have so far largely been aimed at recycling soot to the gasification process. This was desirable for ecological and economic reasons (to avoid waste and ensure maximum carbon conversion). However, processes used for this purpose such as pelletizing and naphtha soot recovery are relatively expensive in terms of both capital investment and operating cost. Soot preparation alone accounts for a considerable part of the total capital investment in the gasification plant.

Residual oils have lately become heavier: deasphalters for instance, produce tars and asphalts which are liquids above 180 °C. In the case of such residues, soot recycling has to be ruled out simply because it would increase the viscosity of the oil. The factors which originally were in favour of soot recycling were thus turning against it and suggested that a new process concept would be desirable. Figure 1 shows a diagram of SGP with soot recycling and with the new alternative without recycling.

First steps: Basic R&D

An older SGP plant with a pelletizing system for soot recovery/recycle experimented with heavier and dirtier oils and found that aspects of cost and environmental protection might make a new soot processing system desirable.

In the first stage of the new system, the soot and ash were to be jointly eliminated by filtration and the filtrate as usual recycled to the process to feed the scrubber after the surplus was withdrawn from the loop. In a second stage, the filter cake was to be burnt and the resulting high-vanadium slag marketed.

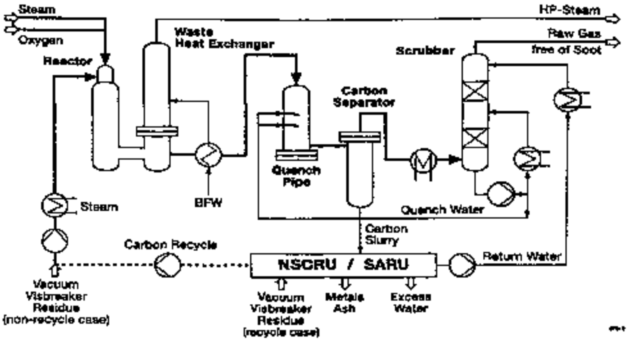

Figure 1

Successful filtration tests had been made with a continuous belt filter press. Also, several incineration tests were made with different partners, but the results of these tests were not entirely satisfactory.

One process developer was confronted with this problem and could offer a new process concept based on its involvement with vanadium refinery residues in the form of petrol coke. These tests made around the mid 80s had led to fundamental new insights into the incineration of such materials.

If vanadium-containing material is incinerated, there is always a risk that the low melting point of V_2O_5 (680 °C) is exceeded and the material agglomerates. Extensive kinetic and pilot tests in the R & D facilities proved that the soot-vanadium filter cake containing some 80 % moisture can be continuously incinerated in a multiple hearth furnace (MHF) - a well proven "classical" metallurgical equipment. Owing to its low slag and high vanadium content (60 - 70 % V_2O_5) the resulting ash is a useful secondary raw material for vanadium production.
(A detailed description of this R & D effort is given in [2].

Further Improvements
The other developer also went ahead with research in the same direction. In this case, it was not an existing plant but one that was being built which gave the first impulse. Shell was looking for a new soot processing concept for its extensive refinery renovation project in Pernis, Rotterdam. Under the similar conditions - very heavy feedstock oils, lower capital investment and operating cost - the results were comparable: a two-stage process consisting of a filtration stage followed by controlled incineration of the filter cake.

Preliminary laboratory filtration tests were made and followed by a five-day test with a membrane filter press at a suitable SGP plant. These tests also produced a handable filter cake with a solids content of more than 20 %. Incineration tests performed with the cake in a multiple hearth furnace at a commercial laboratory in Belgium produced results which were similar to those of the process developers.

With two different filter types - the belt filter press and the membrane filter press - tested and proven reliable, it was still seen worthwhile to expand the available technology range, so further tests were made with different filter types at various licensee's plants. As a result of these test, today a multitude of filter types is applicable, making the process flexible for different site conditions and economics.

From the** concept **and** testing stage **right up to the process

As a results of the research and development effort, a plant engineering firm must have a complete process to offer for which guarantees can be given.

This step was taken jointly by engineers and developers of all parties concerned. The results of laboratory and field tests were analysed and the design parameters and guidelines were agreed upon.

This led to the soot-ash removal process sketched out in figure 2.

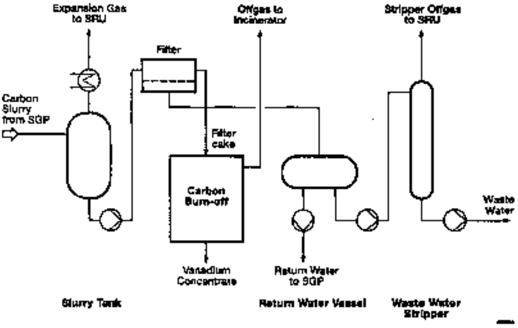

Figure 2

The carbon slurry from the SGP unit is flashed to atmospheric pressure in the slurry tank. The slurry is then filtered on an automatic filter to recover a filter cake with about 80 % residual moisture and a clear water filtrate. The filter cake is subjected to the controlled oxidation process in the multiple hearth furnace. The bulk of the filtrate is returned to the SGP process as wash water. Surplus water is routed to a waste water stripper and from there to treatment. An eye will also have to be kept on the furnace off-gas since it contains CO and $S0_2$.

Especially when new plants are being designed, this off-gas can be discharged together with the flash gases to an appropriately designed Claus unit. Where this is not possible - for instance if existing plants have already been modified or retrofitted - the conventional flue gas treatment processes can be used for working up the off-gas ex multiple hearth furnace, e.g. catalytic CO post-combustion and Sulfacid scrubber system.

The furnace ash is of metallurgical quality with a minimum of residual soot and high metal concentrations, i. e. the process yields a marketable product rather than a waste material. Vanadium pentoxide concentration is typically around 70 %. At 400 ppm vanadium in a feed stream of 100 t/h this translates to 320 t/a of vanadium.

Two fully erected plants and cost estimates for different applications and plant sizes have proved that the capital investment for the process is significantly lower than that for the existing processes.

An additional benefit of this development is increased feedstock flexibility, both for the SGP itself and for the refinery as a whole. The carbon recycle currently practised has the feature of also recycling some of the ash, so that the charge pump, burner and reactor system see a higher content of ash than that present in the fresh feed. The elimination of the recycle thus allows one to use feeds with considerably higher ash contents than previously. Current experience of over 1,000 wppm vanadium at the reactor inlet would be directly applicable to residues of this quality. This is certainly an important feature when reviewing the possibility of introducing heavier (and cheaper) crudes into the refinery.

Current status and future prospects

Two SARU plants of very different capacities (11:1) have been built in the meantime. One of them - Lucky Yochon, S. Korea - was started successfully last year under the supervision of Shell and Lurgi personnel and the second part of this paper gives a report of this start-up. The other - at Shell's Pernis SGHP project - now is in the pre-commissioning stage.

The important design features of these two realised SARU plants are:

minimum operator involvement

availability and reliability at least equal to those of the gasification itself

total on-site facility, i. e. processing from carbon slurry to furnace ash all in one on-site unit

minimum environmental impact, i. e. the water loop of the soot processing is fully integrated with the gasification, offgases are treated or disposed of safely and the furnace ash is clean to metallurgical specification.

The new process - SARU - is seen now as commercialized. Shell and Lurgi decided that SGP shall in future be offered in combination with SARU.

Start-up Report: SARU at Yochon, South Korea

Gasification soot in water generates high viscosities at already low solids concentrations. This phenomenon causes the relatively large wash water requirement as typical for gasification processes. Therefore the SGP and the SARU are connected through a large water circulation loop. The first process step of the SARU is the separation of soot and water. Filtration produces a filter cake and filtrate. The filtrate is routed back to the water wash section of the gasification. Therefore, the continuity of supply of filtrate and the reliability of the filtration is of the utmost importance.

Fully automated membrane chamber filter presses were selected of the several available filtration methods for both projects. These filter presses have a batch mode of operation. Therefore continuous-batch connections with some holdup are required.

The start up of the first SARU successfully demonstrated all the filtration requirements:

The filter presses produced a cake with a typical dry solid content of 20 % without any use of chemical additives such as flocculants. The filtrate is visually clear. The quality of the cake and the filtrate proved to be independent of the gasification process in terms of soot production quantity and quality.

The membrane filter press indeed worked fully automated. The operator attendance was limited to some small regular preventive maintenance like lubrication of moving parts.

The continuous-batch connections performed very well. The filtrate flow to the wash water was steady and continuous.

> No additional waste water was produced. All the utility water for e. g. filter cloth wash was clear filtrate and eventually recycled back. Of course some water is bleeded out of the circulation loop since the gasification process is a net water producer. All the bleed water was successfully processed by stripping and standard effluent water treating.

The filter cake is subsequently transported to the multiple hearth furnace. The produced cake has very good handling characteristics. The cake appears dry in spite of the 80 % water content. No sticking or clogging was observed in the batch-continuous intermediate cake storage.

The next SARU process step comprises the carbon burn off of the filter cake. This in the end generates the product ash. The main challenge of this step is dealing with the low melting point of vanadiumpentoxide (typical 680 *C). The burn off is carried out in a multiple hearth furnace in a controlled way. No melting or sintering was observed in the end product or in the furnace itself. The remaining combustibles content in the product ash was constantly observed below 1 %. The furnace operation turned out to be very robust. No upsets in ash product quality were observed resulting from feed changes in terms of quality or quantity.
All the dust produced by the furnace was recovered from the flue gas and recycled to the multiple hearth furnace. So indeed all the metals in the gasification feedstock ended up as ash from the burn off step.

The remaining flue gas was thermally treated along with other gaseous effluents from the same site in a thermal incinerator.

Summary

The start up of the first commercial Soot Ash Removal Unit was a success. The process requirements in terms of quality of the filtrate, quality and handleability of the filter cake and quality of the product ash were easily met. The process turned out to be very robust. No negative effects of soot quantity or quality due to changes in the gasification section were observed in the SARU part. The operator attendance was limited to some small preventive maintenance. The impact to the environment proved to be minimal since no additional waste water is generated and all the metals in the gasification feedstock end up as product ash. SARU has definitely demonstrated its position as preferred soot processing process for SGP.

Acknowledgements

The authors wish to thank all parties involved in the development of the process for their valuable contributions which made the new process an immediate success.

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Lurgi Metallurgie GmbH, Frankfurt am Main, Germany
Lurgi Ol • Gas • Chemie GmbH, Frankfurt am Main, Germany
Shell International Oil Products B.V., Amsterdam, The Netherlands

References

1. SGP - A flexible process for bottoms upgrading
S. Posthuma, B. Benckhuijsen, M. Senden, SIPM, The Netherlands
IChem^E conference "Gasification: an alternative to natural gas", Nov 1995, London

2. SARU - Soot ash removal unit for Shell Gasification Process (SGP) for oil application - from the idea to the process
W. Liebner, G. Samant, Lurgi Ol • Gas • Chemie GmbH, Germany
Presentation at ACHEMA 1994, Frankfurt am Main, Printed in LURGI Annual Review 1994

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APPLICATIONS OF THE TOPHAT CYCLE IN POWER GENERATION

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Abstract

In the present paper some applications are discussed of the TOPHAT cycle[2] which comprises a Joule/Brayton cycle featuring quasi-isothermal compression of the air by saturating the air with water after each compression stage and a recuperator after the gas turbine for heating the compressed air. The first application illustrates the advantages of combining the TOPHAT cycle with the Optimized Gasification Combined Cycle (OGCC)[1] in which a coal-water slurry is used as a quench medium after a gasifier. Both this scheme and a similar scheme comprising dry coal pressurization result in efficiencies of about 52% with a slight advantage for the OGCC option. Further it is shown that a coal fired TOPHAT cycle including particulate removal at 900 °C (1 650 °F) and a hot gas expander can result in a simple power station without any need for steam and with efficiencies in the range of 45-50%. Such a scheme is even more advantageous for power production from wet hydrocarbon feedstocks as biomass, peat, lignites and Orimulsion®. The efficiency for biomass is >45% which is significantly higher than for gasification based power stations using feedstocks with such a high water content.

Introduction

Most advanced coal fired power stations which have been proposed during the past two decades were based on integrated combined cycles featuring gas turbines with ever higher inlet temperatures. This has resulted in a number of large demo plants which have a few important things in common:

- The capital cost in $/kW is well above that of modern conventional coal fired power stations whereas the efficiency of these plants is not or only slightly higher. Even for fully commercial gasification based power plants using the same principles as currently applied in the demonstration plants it is unrealistic that the somewhat higher efficiency will warrant the additional capital expense.

- [illegible] se for the high capital cost is the [illegible] no; of the major process/equipment items was specifically built for a gasification based power plant. All coal gasifiers were originally built for making synthesis gas, virtually all treating processes were adopted from the refining industry and in the gas turbines the same cycle was used as in aircraft turbines, that is without intercooling in the compressor and without reheat in the turbine. Moreover the fact that a combined cycle was used added to the higher cost of these plants.

- Using the Carnot criterium: $n = 1 - T_L/T_H$ it may be concluded that too little attention has been given to lower T_{Low}. The only attempt to cope with this omission have been the HAT cycles in their various embodiments.

- Despite the emphasis during the past years on exergy too little attention has been given to upgrade the low level heat in the gas turbine exhaust gases to that of the gas turbine inlet by applying recuperators.

- The operability of the plants is often difficult due to a.o. too much integration of the various units. This implies that the plants are most suitable for base load operation and large capacities and it is very questionable whether such plants will be required in the future.

Some of the attempts to cope with the above problems and challenges for that matter have already been discussed [1,2]. In the present paper these topics have been further worked out.

Coal Gasification based Power Stations

The combination of OGCC and TOPHAT

A simplified block scheme for such a plant is given in Figure 1. In a two-stage dry coal feed Destec type gasifier coal is gasified at 32 bar (460 psi) in a first stage with oxygen/steam under slagging conditions at 1 550 °C (2820 °F) and the remaining char is gasified with steam in a non-slagging second stage at 1050 °C (1920 °F). The hot gases leaving the gasifier are quenched with a coal-water slurry of 90 °C (194 °F) to 300 °C (570 °F) after which the dry coal is separated from the gas and transported with nitrogen to the gasifier. The sulphur removal is either accomplished by a hot gas treating such as the KEMA warm gas treating process or by flue gas desulphurization. In this and all other coal gasification based systems the stack gas temperature is such that there is sufficient heat left for raising the medium pressure steam for gasification, gas treating and -where required- for drying of the coal.

The fuel gas and the humidified air leaving the TOPHAT compressor (two stage with humidification after each stage) are both heated to 550 °C (1020 °F) in a recuperator before being combusted at 1350 °C (2460 °F) in the gas turbine. The exhaust gas fron tl gas turbine is routed through the re...r at-.' anc .ed for drying the coal, process steam generation and coal slurry preheating, etc. before being routed to the stack. The efficiency of the above scheme is 52.3 % based on the LHV of the coal.

The combination of TOPHAT compression and a gasifier with dry coal pressurization

The block flow scheme of this plant is similar to the previous one and is depicted in Figure 2. The coal is dried in a conventional dryer, pressurized in lockhoppers or another coal 'pump' and then routed to a gasifier which operates under similar conditions as in the previous case. The hot gases leaving the gasifier are quenched with water of 200 °C (390 °F). The efficiency of this scheme is 51.8 %. A possible advantage of the present scheme is that it can be operated without the danger of any tar formation. A disadvantage is that as long as lockhoppers have to be used for pressurization of dry coal the pressurization is more complicated than with a slurry system.

Using a dry coal feed pressurizing system it is possible to completely avoid tar formation with the result that in principle the filtering of the coal could be carried out at a temperature of 500 °C (930 °F). However, by doing away with the steam cycle and thus with the expensive syngas cooler it is more advantageous to lower also the filtering temperature in this case to 300 °C (570 °F) because then more heat is extracted in the recuperator which has a beneficial effect on the station efficiency.

The above scheme has also been studied for a combination of a cold (TOPHAT pressurized) nitrogen quench and a water quench but this resulted in a lower station efficiency due to the required compression energy for the nitrogen. The atmospheric nitrogen can still be usefully employed though buy using it for diluting the inlet air of the air compressor of the gas turbine. Doing this results in combustion air with an about one percent lower oxygen content as well as in a lower air inlet temperature. The lower oxygen content reduces the stoichiometric adiabatic flame temperature and hence the thermal $N0_x$ emission and the lower air temperature results in a lower compressor duty and hence in a higher station efficiency. This use of the surplus nitrogen from the Air Separation Unit is also applicable to the OGCC case described above.

Highly integrated coal gasification based power plants

It is of course possible to obtain higher efficiencies of up to 55% with coal gasification based power plants by applying higher pressures, more isothermal compression, reheat gas turbines, etc. Flow schemes with these features have been calculated but it is believed that these have only theoretical value because they are very complex and hence lead to expensive plants which are difficult to operate.

Coal fired TOPHAT cycle

The coal fired TOPHAT cycle was specifically developed with the purpose of obtaining an efficient, economical and clean coal fired power station. A block flow scheme of a coal fired TOPHAT cycle is depicted in Figure 3. Air is compressed in one, two or three stages and cooled by direct injection of water after each stage. The air is then heated in a recuperator before being used as combustion air and optionally also as quench air. The combustion takes place at a low pressure of 8 bara (120 psia). The combustor may comprise a PFBC, PF firing or a slagging combustor followed by a quench. During or after the combustion the gases are desulphurized with limestone and/or dolomite. After injecting an alkali getter the gases are filtered at 900 °C (1650 °F) and routed to a hot gas expander. The gas expander is rather simple as no blade cooling is required. The hot gases leaving the expander pass through the recuperator, and optionally through a heat exchanger preheating the water which is injected during compression, before being sent to the stack.

The resulting efficiencies for the various cases which were studied are given in the table below.

Station efficiencies for coal and biomass fired TOPHAT cycles.
Because of losses and the own power consumption 2 - 4 percentage points should be substracted to obtain realistic figures.

Temperature injection water, °C/°F =>	25/80	Boiling point at prevailing pressure in the compressor	200/390
COAL			
Injection stages 0			
1	43.8	45.1	45.5
2	47.3	47.7	48.0
3	48.5	48.7	48.9
6	49.5		49.6
BIOMASS			
2	54		

The data in the above figure show that most benefits of the water injection are already obtained after 2 or 3 in;"""-'' *n stage-' This implies that injection inside the ~^ essor is not per se required to reap the benefits of the TOPHAT cycle. The effect of the temperature of the water injected after each compressor stage is largest in case only one compressor stage is applied but becomes appreciably smaller in case of more injection stages. Nevertheless even in the case of three injection stages the effect is still about half a percentage point in case water of 200 °C (390 °F) is injected instead of water of ambient temperature (25 °C, 80 °F).

From the above data it can be concluded that the efficiency of this simple power station is only 2 - 3 percentage points lower than that of a gasification based power station in case 3 or more compression stages are applied. Further it can be concluded that this cycle has a similar efficiency as the most advanced PF steam cycle presently under construction in case 2 compression stages are applied. As the recuperator is rather simple because nowhere the metal temperatures exceed the 500 °C (930 °F) the capital cost of the coal fired TOPHAT cycle is expected to be much lower than for the gasification based schemes and the operation is much simpler.

The above scheme is especially attractive for wet feedstocks such as biomass (see Figure 4), peat, lignite, etc. because the remaining heat in the flue gases can be used to dry the feedstock. With biomass containing 50% water the efficiency is 50 % !

An additional advantage of the indirectly fired TOPHAT cycle for biomass is that it can be built for relatively small power stations of 5-10 MW_a which implies that biomass has to be transported over much smaller distances.

Further it is observed that for low pressure ratio Joule/Brayton cycles with low compressor outlet temperatures the recuperator can extract more heat from the stack gases resulting in relatively high station efficiencies considering the low gas turbine inlet temperature.

Coal pressurizing

The fact that the pressure of the coal fired TOPHAT combustor is only 4 - 8 bara (58 - 11 6 psia) makes coal pressurizing relatively easy and can make expensive lockhopper systems superfluous. An elegant system comprises a high bunker which is designed such that:

- the pressure in the combustor is lower than the corresponding head of the coal column in the bunker

- the upward velocity of the gases through the interstices of the coal bed in the bunker is equal or lower than the downward velocity of the coal bed in the bunker

- rm w p'ev. in the bunker.

Another elegant option is to use a 'Stamet' pump for transporting the coal against pressure. The principle of this pump is that the coal is entrained by the sides of a rotating spool by the same effect which causes hang-ups in coal bunkers thus making a virtue of necessity.

Conclusions

- With modern gas turbines combined with two-stage TOPHAT air compressors coal gasification based power stations can be built with an efficiency of 52 % based on the LHV of the fuel. Due to the absence of a steam cycle the capital cost of such power stations is lower than for conventional coal gasification based combined cycle stations whereas the operation of the plant becomes simpler. Integration with the Air Separation Unit is not required.

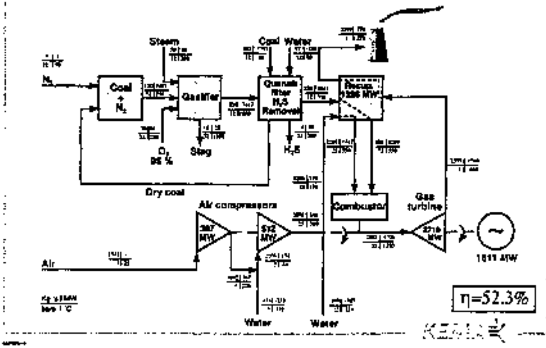
- To obtain the efficiency of 52 % a two-stage gasifier of the Destec type is required using a dry coal feed. The dry coal feed can be obtained by pressurizing dry coal in e.g. lockhoppers but it is more elegant to pressurize a coal-water slurry and use this as a quench medium for the gas leaving the gasifier as applied in the OGCC. Moreover the OGCC option results in about a 0.5 percentage point higher station efficiency.

- The coal fired TOPHAT cycle comprises the almost direct firing of coal in a gas turbine. The hot combustion gases do pass through a filter though and moreover an alkali getter is required. Although the alkali problem will probably limit the gas turbine inlet temperatures to 900 °C (1650 °F) the efficiency of these power stations range from 46 - 48 % provided at least 2-3 compression stages are used. The low gas turbine inlet temperature has the advantage that no blade cooling is required.

- For wet feedstocks a directly fired TOPHAT cycle may be advantageous as it results in a high station efficiency and a simple plant. For biomass with a 50 % moisture content the efficiency amounts to 50 %. Moreover the absence of the steam cycle has in this case the additional advantage that the plant can be built for small duties of 5-10 MW_e which is especially attractive for biomass because it can reduce the problems associated with the transport of this material with its very low energy density.

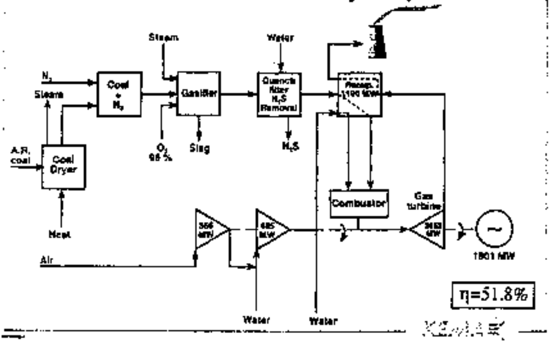
- Recapitulating it can be concluded that the TOPHAT cycle offers in all applications major advantages in terms of efficiency, capital expense and operability. In relation to this it is reminded that these advantages also apply to r -jtu. gas fired power stations as was illu.

The combination of OGCC and TOPHAT



η=52.3%

The combination of TOPHAT and dry coal pressurization



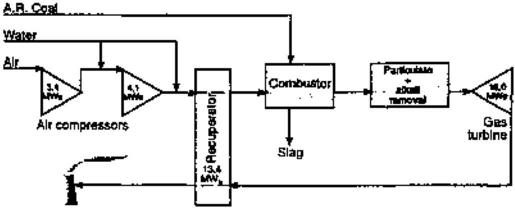
η=51.8%

References

1. Thirteenth EPRI Conference on Gasification Power Plants. *OGCC: Optimized Gasification Combined Cycle,* 1994, Sheraton Palace Hotel, San Francisco.

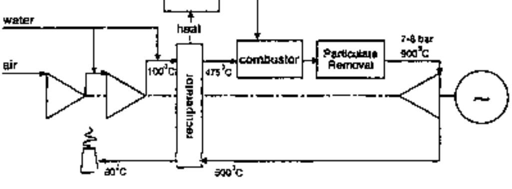
2. EPRI Conference on New Power Generation Technology. *TOPHAT and REX turbines,* 1995, ANA Hotel, San Francisco.

Coal fired TOPHAT cycle
two stage air compression
T. injection water 200*C



n=48,o%

Biomass fired TOPHAT cycle



η= 50 %

DYNAMIC SIMULATION: AN ENGINEERING TOOL TO OPTIMIZE ISAB ENERGY IGCC PLANT DESIGN, CONTROL AND OPERABILITY

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ABSTRACT

In the course of the execution of detailed engineering of ISAB Energy project, a Dynamic Simulation Study of the Integrated Gasification Combined Cycle Plant is under development.

Dynamic Simulation is an engineering tool necessary to check and to finalize the overall plant design.
The dynamic simulation which is commonly applied to CCU plants, is particularly important for IGCC Complex where the integration between the Gasification section and the CCU section, is essential for the correct and safe operation of the Plant. The simulation model is used to predict the transient behaviour of the IGCC plant subsequent to a planned or unplanned disturbance of the steady-state operation.
Complete plant responses (stream flows, temperatures, and pressures) to these events are predicted and evaluated for their acceptability.
The paper describes in detail the steps which are being followed in the development of a dynamic simulation study of the ISAB Energy IGCC Plant:

a) the dynamic model is built assembling equipment design data, operating data, control valve and controller data, process flow and logic diagrams;

b) the planned and unplanned events to be simulated are defined in accordance to the plant operating modes and the operating requirements specified by the electric power purchaser (i.e. plant load variations, gas turbine load rejection, disconnection from the electric grid, island operation etc.);

c) the simulations are performed and their results discussed. If necessary, modifications to equipment and control devices are implemented;

d) finally all the information relevant to the control of the plant, derived from the dynamic simulation are implemented in the control system (i.e. ramp for planned load change, controller parameters, advanced control strategy etc.).

PURPOSES OF THE DYNAMIC SIMULATION STUDY

The Dynamic Analysis is by now an engineering tool, commonly applied in the design of Combined Cycle plants. This is due to the operating features of these plants: frequent load variations, large operating flexibility, sudden disconnection of plant from the electric network, possible island operation. The expected transients are imposed to a plant consisting of sections with different time responses: very quick for the machines, much longer for the steam cycle.

FWI developed several dynamic simulation studies for Combined Cycle and Cogeneration Units: the first one for FIAT AUTO Mirafiori plant (2 x 60 MWe Units) in 1988, following for SERENE Project (8 x 50 MWe Units), for a Refinery Cogeneration Units (25 MWe) and for CENTRO ENERGIA projects (2 x 150 MWe).

The questions described for CCU are furtherly stressed in the IGCC plants, where the feed to the CCU is supplied by a Complex Plant connecting intrinsically the fuel and the power productions: the integration between the Gasification Section and the CCU section is essential for a flexible and safe operation of the Plant.

Based on these considerations a Dynamic Simulation Study has been planned for the ISAB Energy IGCC Project, and is under development.
It consists in building a dynamic simulation model describing the sections of the plant which are dynamically significant. Then the simulation model is used to" predict transient behaviours of plant variables such as temperature, pressure, flow, caused by planned or unplanned events.

The Dynamic Simulation Study performed during the engineering phase of the ISAB Energy IGCC project is aimed to the following targets:

Check of the equipment sizing; the dimensions defined on the basis of one or more operating and design cases, shall be suitable to withstand the transients which might prove more critical than the steady state operations.

Check of the control strategy, development and finalization of the control system; this includes definition of ad hoc control philosophy to solve particular problems, and to ensure that no undesirable or unsafe conditions are expected during transients; check of control valve size and characteristics.

Selection of safe operating procedures such as rate of load changes.

Estimate of controller parameters, allowing a shorter tuning on field.

For the study execution many detail information relevant to equipment geometry, control devices data and characteristics, are needed. It is performed as soon as these data are available from the Vendors, trying a compromise between the need to forward the check of the equipment and the control system, and the availability of the required data.

PROCESS AND CONTROL DESCRIPTION

The ISAB Energy IGCC Plant is designed to process heavy oil residues (i.e. Asphalt, Visbroken Vacuum Residue, Fuel Oil, etc) coming from the adjacent refinery.

The Plant is composed mainly of the following sections:

- Gasification: two Texaco Partial Oxidation Reactors using steam as moderator and oxygen as oxidant, of direct water quench type, each followed by one scrubber, to remove the soot and ash from syngas.

- Carbon Recovery and Recycle to recover soot from soot water and recycle it to the gasifiers.

- Syngas Heat Recovery section where raw gas from Gasification is cooled by generating steam and hot water, with separation and condensation of most water vapour. The catalytic hydrolysis of COS to H_2S is also achieved in this section.

- Acid Gas Removal where raw gas is scrubbed by means of formulated MDEA in order to selectively remove H2S, minimizing C02 co-absorption.

- Purified gas is repeated, expanded by producing additional electric power, and humidified with water heated in the above mentioned Heat Recovery Section.

- Finally syngas enters the Combined Cycle Unit composed of two identical trains consisting of the gas turbine, the heat recovery steam generator with post-combustion, the steam turbine.

In addition to these main sections, the IGCC Complex includes the Metals Recovery Section, the Sulphur Recovery and Tail Gas Treatment Section and all me Utility Systems required for the operation of the Plant.

The IGCC Complex control system is aimed to manage the electric power production of the five power generators connected to the national electric distribution grid.
The IGCC Complex operate either in the power control mode or in the feed control mode.
In the power control mode the amount of power produced" by the complex is a set point defined by the management of the power distribution grid.
In the feed control mode the amount of power produced by the complex is limited by the amount of available feed up to the maximum throughput capability of the complex. In essence the feed control mode is a specific case of the power control mode where the specified power output is the maximum possible.

In other words, during normal operation a fixed amount of electric power shall be produced or a fixed quantity of asphalt shall be destroyed. Variations in both requirements promote an unbalance between the syngas production and the syngas consumption. The unbalance produces a variation of pressure in the raw syngas header.
When the electric power production requirement is to be changed, a new set point is entered. This is compared against the measured power output, activating the signal to CCU to change the power output. This causes a variation of the syngas requirement and consequently of the clean syngas header pressure.

The clean syngas header pressure activates the expander controller to maintain constant the syngas pressure at the inlet of the Combined Cycles by throttling more or less on expander inlet and/or bypass. Consequently the raw syngas pressure in the Heat Recovery Section fluctuates and a variation of the syngas production from the Gasifiers is called. Steam, oxygen and hydrocarbon feed rates are changed to adjust the syngas production in order to match the clean syngas demand of the Combined Cycles and re-establish the raw syngas header pressure.
This type of control maintains constant the pressure at the gas turbine inlet, during the transient'period, while the raw syngas pressure upstream the expander is let to fluctuate, utilizing as a buffer for capacity adjustment, the large gas inventory existing in the system operating at high pressure.
The same procedure is applied when a feed rate variation is necessary.

The above described functions are performed through four main control systems:

> The IGCC complex controller which monitors and controls the raw syngas header pressure by sending signals to the gasifier control system and/or to the Combined Cycle master control system to balance the raw syngas make and the clean syngas consumption: the IGCC Complex controller allows the front end of the complex where syngas is produced to match up quickly to the back end of the complex, where syngas is consumed. This will minimize the response time of the complex to changes in plant power output or plant feed rate.
>
> The gasifier control system which signals the two individual gasification train control systems to control the train throughput of the respective gasification train. In so doing, the gasifier control system controls the total raw untreated syngas production from the two gasification trains.
>
> The Combined Cycle master control system which signals the two individual Combined Cycle Unit controllers to control the clean syngas input to one or both of the respective combined cycle units and in so doing controls the total treated syngas consumption in the two combined cycle units.
>
> During normal operation the unit controller optimizes the power generation efficiency by the distribution of load between the two trains, and within each one, between the gas turbine and the postcombustion.
>
> The raw syngas header pressure control system which maintains the raw syngas header pressure upstream of the gas expander to allow efficient heat recovery from the hot syngas from gasification.

The dynamic model of the ISAB Energy IGCC Plant describes the whole process involving syngas, starting from its generation in the gasifiers, through cooling, H_2S washing, expansion, humidification and combustion in the CCU. These systems are connected, from the operating point of view, by the mentioned main controllers. The other ancillary sections (i.e. Metal Recovery, Sulphur Recovery and Utility Systems) are not simulated as the associated dynamics are not significant.

DYNAMIC MODEL PREPARATION

The model for each section describes the main components, the piping and the associated control system; it integrates all the information necessary to evaluate the mass, thermal, hydraulic balances, predicting dynamically stream flows, temperatures and pressures during the transient.

Data Gathering

The following plant and equipment data will be assembled to build the dynamic model:

a) Process flow diagrams of the plant

b) Equipment physical data. This includes volumes, surfaces, dimensions, geometric arrangements and design characteristics of mechanical equipment in order to simulate off-design component behaviour for gasification and combined-cycle components and associated valves and piping.

c) Operating point data. Heat and mass balance for base-load operating condition. This includes all stream information (mass flow rates, pressures, temperatures, enthalpies, and compositions).

d) Controls and logic drawings for the equipment and plant. Control valves and controllers data.

e) Plant operating philosophy.

Model Preparation

The model will be built using a commercial dynamic simulation software of modular type. Some additional modules will be customized to describe adequately the ISAB Energy IGCC Plant.
The following steps are followed:

a) First, a model schematic is generated. This involves laying out the process which defines the scope of the model.

b) A diagram is then created which depicts the selected software modules and their connections used to simulate the process.

c) Most components can be simulated using modules from the standard software library. New modules for unique components will be developed, as necessary.

d) The next step is to superimpose a process heat balance with enough information to define the pressure, flowrate, enthalpy, and composition of each stream at operating point condition.

e) Drawings of control strategies are developed from the operating procedures and plant controls and logic drawings.

0 With the plant scope defined, all modules selected, all data gathered, a dynamic model of the plant is configured. The model will include all the main components (e.g., coolers, gasifiers, exchangers, drums, absorber, expander, saturator, combustors, turbines, heat recovery steam generators, valves and all associated piping) in the plant as a series of resistance and volume modules connected together in a thermal/hydraulic network.

g) Once the model is created and all appropriate variables initialized, a quick next step is to test the model at steady-state conditions to choose if the model variables match the heat balance at the operating condition, both in design condition and in offdesign condition.

h) The next step is to dynamically test the model. Test disturbances are introduced into the model and the system's response in terms of flows, pressures and temperatures observed. The system should pass from the original steady-state condition to a different final steady-state condition through a transient which can be properly discussed.

EVALUATION OF PLANT TRANSIENTS

Based on plant operating philosophy and operating experience with similar plant, planned, unplanned or upset events for dynamic evaluation will be identified.
Planned events are Gasification and Combined Cycle load variations. The simulation study is aimed to define the faster load ramp, accepted by the equipment, minimizing the impact on their life.

The expected upsets of the CCU operation like:

- sudden disconnection of the Unit from the national grid;
- island operation feeding only the CCU auxiliaries;
- trip of one gas turbine;
- trip of one steam turbine;
- trip of HRSG postcombustion system;

will be studied in their effects on the whole Plant operation.

The same will be evaluated for upset conditions relevant to the Gasification and Syngas Treatment Sections, such as Gasifier trip , Expander trip ect.

Once the planned and unplanned plant events are selected, the model will be exercised for each of the transients. Complete plant responses (stream flows, temperatures, and pressures) to these events will be predicted and graphically presented. These responses will be evaluated for their reasonableness and acceptability. If a certain response has the potential for equipment damage or other unsafe conditions, plant design modifications will be investigated to preclude such conditions. Revised plant configurations will be re-evaluated for their acceptability.

The final product will be a dynamic model of the ISAB Energy IGCC combined-cycle plant with plant response predictions for the identified planned and unplanned (upset) events.

CONCLUSIONS

At present the Data Gathering and Dynamic Model Preparation phases are under development. The modelling of the ISAB Energy IGCC Plant is expected to be completed within four-five months and three months are scheduled for the evaluation of the plant transients.
As already happened for the developed CCU dynamic simulation studies, the IGCC dynamic model is expected to be a powerful tool to check the plant design, to optimize the control system and to explore the plant operability.

REFERENCES

- Maderni L., Icardi G. and Fontana M., 1989, "Control System for a Combined Cycle", ASME International Gas Turbine & Aeroengine Congress & Exposition, Toronto - Canada.

- Ahluwalia K.S. and Domenichini R., 1989, "Dynamic Modeling of a Combined Cycle Plant", ASME International Gas Turbine & Aeroengine Congress & Exposition, Toronto - Canada.

- Giglio R., Cerabolini M. and Pisacane F., 1996, "The Dynamic Simulation of the Progetto Energia Combined Cycle Power Plants", International Joint Power Generation Conference, Houston - Texas.

THE PUERTOLLANO IGCC PLANT DYNAMIC SIMULATOR

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Project Manager

0. INTRODUCTION

Integrated Gasification Combined Cycle has emerged as a promising technologies for coal -based electric power generation. While each subsystem of IGCC plants have been in commercial use for a number of years, subsystems such air separation unit to produce oxygen, coal gasification and hydrogen sulphide removal are new to the power industry.
The importance of power plant dynamics in relation to safe economical operation has been recognized throughout the industry. Transients, which occur as a result of both planned and unplanned events, can impact reliability, availability, efficiency, operating and maintenance cost, and even the integrity of plant systems. Computer modelling is an effective tool for studying operating conditions, analyzing performance, and understanding and predicting dynamic behaviour of equipment and system under a variety of plant conditions. It is clear that multiple-subsystem, multiple-train IGCC plants especially would benefit from dynamic simulation.

The major application of power plant computer modelling have been engineering simulation studies and operator training simulators. These two applications traditionally have employed different approaches and have been based on divergent modelling criteria. Typically engineering simulation is performed on large mainframe computers operated in either a batch mode or a mode with limited user interaction, since its purpose is a comprehensive analysis of complex engineering problems. Training simulators have been developed for fast or real-time interaction between the user and the computer model: however, their analytical capabilities and general acceptance by the industry have been limited by the cost and performance of available simulators.
An efficient plant modelling technique, which can be applied to both in-depth engineering studies and real-time operation, would hasten the maturity of IGCC technology. In the long term, simulation analysis can provide a frame work for consistency in the evolution of IGCC technology and can be routinely applied to aid in the process design, to develop operating plans and procedures, and during commercial IGCC projects to train the utility staff in new technology operation.

1. THE PUERTOLLANO IGCC PLANT DYNAMIC SIMULATOR PROJECT

Recognizing the potential benefit of developing a powerful, state of the art dynamic simulator to support IGCC demonstration and commercialisation, Elcogas decided to built a plant dynamic simulator oriented to train plant personnel and conduct engineering studies.

The work is conducted as a joint development of Elcogas with Enel and EDF according to a specific Cooperation Agreement signed among the Parties and under the Project Management of ENEL.

The selected scope of the simulator is not a full-replica but the systems neglected fall into the conventional category (coal preparation, cooling water distribution, demi-water plant, ..) and the simplified ones correspond to minor impact on the main plant components operation (sulphur recovery).

The use of the LEGO environment, ENEL proprietary code, assure the availability of a reliable and state of art software package as the base of the simulation tool. At the end of the development, the LEGO system will handle many different process and control models of very complex nature and reproduce the advanced functionality features and alarm management of the Siemens Teleperm XP.

This project is part of the successful Thermie proposal entitled "Puertollano Project: Activities to improve the efficiency, availability and economics of the current and future IGCC".

The simulator consists in a detailed dynamic modelling of the various systems of the plant including the control. automation systems and alarm management.
The simulator represents a flexible and accurate tool to predict plant behaviour before commissioning. In particular it allows:

- the analysis of plant performance in off-design conditions
- the analysis of plant transients
- the optimization of load following operation of the plant in terms of dynamic response, thermodynamic performance and pollutant emission

- the assessment of the most suitable operational procedure for normal and emergency plant conditions
- the testing of the control and protection systems

The simulator, once validated and tuned with operation data, will be used to:

- Train plant operations on the CC/IGCC systems with graphic interface reflecting the actual plant control room to improve operator skills
- Perform malfunctions analysis to train the operator in abnormal plant operating conditions
- Conduct analysis of plant dynamic responses in design and off-design conditions for control concept evaluation and trip prevention
- Support the plant operation: optimization of the operation procedures
- Support process engineers in the plant transients studies: process and control, optimization
- Analyze specific phenomena through stand-alone models upon the identification of operational disturbances during the operation
- Enhance the safety key issues of the plant.

2. THE PUERTOLLANO IGCC PLANT

2600 tons/day of fuel, 50% raw coal and 50 % petroleum coke are milled and dried in the two mills of the coal preparation unit and fed to the Krupp Koppers entrained bed gasifier. The IGCC plant has been designed for a wide range of the fuels utilized in the ECC (see fig. 1).

The dry fuel is gasified with oxygen at purity of 85% in volume in the gasifier in an highly integrated process configuration.
The produced raw gas is quenched at the outlet of the gasifier with dedusted warm gas and cooled down in a HP and IP waste heat boiler system to approximately 235°C.
The coal gas is dedusted in two ceramic candle filters based on the LLB design and nitrogen is used for cleaning the candles.
Nearly 100% of the fly ash is recycled via a lock hopper system to the burners of the gasifier. The granulated and non-leachable slag is transported to slag extractors via the slag crusher and a lock hopj er system.

In a scrubbing systems, residual dust, halogens and alkalis are removed from the dedusted raw gas at approximately 125°C. Further treatment of the raw gas is performed in the MDEA process. A COS hydrolysis unit converts the COS of the raw gas to H_2S upstream of the desulphurization unit in which the H_2S is adsorbed in MDEA solution. After stripping , the clean solution is recycled to the absorber and the sour gas is routed to a Claus unit. The tail gas from the Claus unit is hydrogenated and recycled to the COS hydrolyse unit avoiding the use of an incinerator system, which would result in emission of sulphur components.
Only clean gas and liquid sulphur leave the plant.

The gasifier produces more than 4.300.000 m^3n raw gas per day which results in the generation of 335 MW (gross) electric power together with the steam produced from the gasification unit.
The clean coal gas, after saturation and mixing with N_2 is burned by the coal gas burners in the combustion chamber of the advanced gas turbine Siemens Model V.94.3 with an output of 190 MW (based on ISO condition). The sensible heat of the extracted air is used for clean gas saturation, low pressure steam production and preheating of impure nitrogen.

The gas turbine compressor delivers compressed air for the combustion chamber of the gas turbine and for the air separation unit. Besides the recycling of impure nitrogen from the air separation to the gas turbine, the clean gas is saturated with water to diminish NOx emissions and to increase plant efficiency and power output.
On the water/steam side, HP and IP-steam from the gasifier is superheated in the HRSG from Babcock & Wilcox Espafiola and used for power generation, the necessary feedwater being heated in the HRSG. In the feedwater tank, the condensate from gasification, feedwater for the HRSG and feedwater make-up for the overall system is degased.

The Siemens 145 MW steam turbine is of two casing design . In the first casing, HP and IP steam is expanded. In the second casing, LP steam is expanded through a double flow turbine. Steam from the HP section is reheated and again used in the IP part of the steam turbine. In front of the LP casing, LP steam from the HRSG is added. The exhaust steam flows to the condenser which is cooled via a cooling tower.

The oxygen plant consists mainly of separation columns with front end purification. In addition the unit includes a nitrogen and oxygen storage system and the necessary production compressor. The unit is designed to produce with very stringent specification 85% purity oxygen and 0.1% purity nitrogen for the coal gasification unit. Residual nitrogen, the oxygen content of which is limited to 2%, is added to the fuel gas added to the gas turbine combustion chamber for NOx formation limitation. The Air Separation Unit is fed by the gas turbine compressor: 300.000 m^3n/h air at 15 bar are cooled and purified before separation in a dual column distillation system. The first column operates at medium pressure and yields pure liquid nitrogen at the top. The low pressure column yields at the bottom oxygen and at the top both gaseous impure and pure nitrogen. Cold production is ensured by expanding gases through expansion turbines.

3. **THE** SIMULATOR

LEGOCAD is a modern and powerful real time simulation environment developed by ENEL oriented towards conventional power plant and supports the user in carrying out the mathematical models and the simulation studies.

LEGOCAD is made of a number of integrated tools covering every need in simulators construction .

Its core consists of a large easily readable library of mathematical models, summing up 25 years experience of many modelling specialists and allows the user to set up an overall plant by linking elementary modules and splitting the plant in several subsystems.

It allows also the user to create new modules regarding "not conventional" processes.

The general approach of the simulation is to use the fundamental principles of physic and chemistry based on the mass, momentum and energy conservation equations.

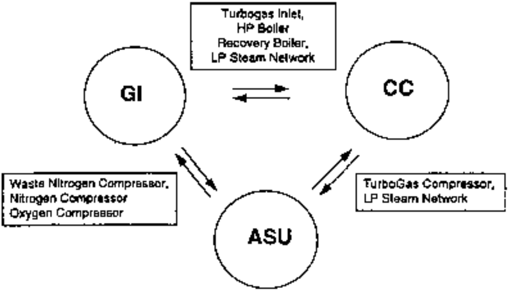
The advantages of this approach are:

1. it's not necessary to know the component operating curve in order to develop the model:
2. the models can work in different operating conditions including off-design conditions
3. modifying the geometrical and chemical-physical parameter, the model of the component can be implemented for various technologies.

The Puertollano IGCC dynamic simulator is a phased product corresponding to the phases of the Puertollano Project: in June '96 the complete Process Training Simulator for the Combined Cycle was completed in order to allow to train ELCOGAS plant personnel and at present a simplified version for the Advanced Processes (Gasification Island, Gas Cleaning and ASU) is available.

The complete IGCC Training Simulator is foreseen for September 1997 and, according to the real plant subdivision, the simulated areas are the following:

Gasification Island (GI)
Combined Cycle (CC)
Air Separation Unit (ASU)



The Gasification Island consists in 10 process tasks and 10 automation and control tasks:

- Gasifier and Heat Recovery System
- Slag Building
- Start up Burners
- Nitrogen System
- Fly Ash Recycle
- Nitrogen to the Burners

- Syngas Network
- Flame System
- Liquid System
- Data Treatment

In particular the simulator contains a 3-D code for predicting the behaviour of the Krupp Koppers high pressure entrained bed gasifier.

The Combined Cycle Island consists in 8 process tasks and 8 automation and control tasks:
- High Pressure System
- Low Pressure System
- Intermediate Pressure System
- Condensate System
- Gas Turbine System
- Steam Turbine System
- Auxiliary System
- Electric System

The Air Separation Unit Island consists in 7 process tasks and 7 automation and control tasks:
- Air Purification System
- Pure Nitrogen Network
- Oxygen Network
- Distillation Column System
- Cryogenic Exchangers 1 and 2
- Air and Nitrogen Preheating

The overall IGCC Simulator runs in real time on an Alpha Workstation DEC 600/5/266 AXP having a CPU power of 123 Mflops, 128 Mbytes RAM and 6 Gbytes Hard Disk.

4. RESULTS OF SIMULATION

As an example, a comparison between the results of simulation and the corresponding plant behaviour during a manoeuvre is reported below.
The manoeuvre is referring to the Project Training Simulator for the Combined Cycle, already started.
The manoeuvre consists in the rolling of the gas turbine up to the speed of 3000 rpm (4 / 12 min), in the setting of 3000 rpm (12 / 26 min), in the synchronization and the network connection at 50 Hz (26 min), in the ramp at 60 MW with a gradient of 8 MW/min (26 / 33 min) and in the setting at the load of 60 MW (33 / 75 min).
The thermal starting condition of the heat recovery is practically "cold" with the feedwater temperature and the inlet circulation rings at about 50-80°C, depending on the plant area, and with steam pressure equal to the atmosphere.
The starting condition cannot be replied on the simulator because it is a transient status coming from the cooling process which lasts several days and starts from unknown conditions.
A starting cold condition has been chosen.
The fluidodynamic starting condition of the water/steam circuit needs that ail the pumps are switched on (condensate extraction, feedwater and circulation rings).
The condition is fully replied on the simulator.
The comparative analysis is focalized on the dynamic trends and on the final steady state values of the main thermodynamic process variables of the steam coming out from the superheating section of the high and intermediate pressure areas of the recovery system. These process variables are the most significant ones as far as concern the main phenomena which take place during the manoeuvre:
- The first evaporation (phase change) inside the circulation loop evaporators during the metal heating over 100°C (evaporation temperature at atmospheric pressure)
- The subsequent circulation rings pressurization with the consequent steam production.
- The steam superheating at the recovery system outlet as function of the available energy in the gas turbine flue gas during the load increasing.
- The status of the recovery system during the manoeuvre depends only on the condenser by-pass valves position.
- The position of the high and intermediate pressure section by-pass valves is manually set on 80%.

The results are presented in the graphics of the fig. 2-3-4. Each figure includes two graphics: the upper one refers to the plant recorded values while the lower one refers to the corresponding simulator values.
A brief description of the variables and their comparison is reported:

FIG. 2

0MBR10CT001	Outlet gas temperature at the recovery system inlet
OHNA20CT001	Outlet gas temperature at the recovery system outlet.

The initial different trend depends on the different thermal state of the exchangers metal wall at the beginning of the manoeuvre.

FIG. 3

0LBA10CT001	High pressure steam temperature at the final superheater outlet
OLBA10CP001	High pressure steam pressure at the final superheater outlet
0LBA10CF001	High pressure steam mass flow rate at the final superheater outlet

The initial flow rate difference depends on the lack of information about the valve drains.

FIG. 4

0LBC80CT001	Intermediate pressure exhaust steam temperature at the reheater inlet (high pressure by-pass discharge)

The difference depends on the different operating procedure of the high pressure by-pass desuperheating system.

OHAH81CT001	Intermediate pressure steam temperature at the superheater outlet. It is the temperature of the steam coming from the IP circulating loop, before the mixing the IP exhaust steam coming from the HP bypass discharge.
OLBB90CT001	Intermediate pressure steam temperature at the reheater outlet
OLBB90CP001	Intermediate pressure steam pressure at the reheater outlet

5. CONCLUSION

The Puertollano IGCC Dynamic Training Simulator, fully completed in September 1997, appears to be one of the very few products developed for this kind of technology on industrial size at international level.

The Simulator, after an adequate tuning on the running plant, due to the fact that it has been developed starting from first principles and due to the characteristics of the LEGO Environment, could be easily used as starting point to develop specific simulators for other plant configurations and technologies involving the gasification coupled with the combined cycle or for new advanced power generation systems.

BIBLIOGRAPHY

1. G. Garbossa, R. Mosca, S. Spelta, R. Cori, P. Cento, "A model of Alto Lazio boiling water reactor using the Lego code: Nuclear Steam Supply System Simulation", Seventh Power Plant Dynamics, Control and Testing Symposium, Knoxville, Tenn. (USA), May 15-17, 1989.
2. G. De Michele, L. Ghiribelli, S. Pasini, A. Tozzi, "A 3-D Code for Predicting Radiative and Convective Heat Transfer in Boilers", ASME 26th National Heat Transfer Conference, Philadelphia, Pennsylvania, August 6-9, 1989.
3. G. De Michele, S. Pasini, A. Tozzi, "Simulation of Heat Transfer and Combustion in Gas and Oil-fired Furnaces", The Combustion Institute, Sezione Italiana e Sovietica, Pisa, Italia, Novembre 5-7, 1990.
4. G. Garbossa, S. Spelta, P. Groppelli, U. Zoni, A. Rossi, "Real-time models for an engineering simulator of a 660 Mwe fossil-fired power plant", Western Multiconference on Computer Simulation - Modelling on Micros & Workstation Technical Conference, Anaheim California, January 23-25, 1991.
5. G. De Michele, S. Pasini, A. Tozzi, "Simulation of Combustion and Pollutant Formation in Gas- and Oil-fired Furnaces", 2nd European Conference on Industrial Furnaces and Boilers, Algarve, Portugal, April 2-5, 1991.
6. G. De Michele, S. Malloggi, S. Merlini, S. Pasini, "Simulation of Combustion and Nox Formation in Gas- and Oil-fired Furnaces", SIAM Fourth International Conference on Numerical Combustion, St. Petersburg, Florida. December 2-4, 1991.
7. G. De Michele, S. Pasini, A. Tozzi, "Combustion Optimization in Power Station Boilers by Advanced Modelling Techniques", Int. Workshop on Mathematical Methods in Combustion, Como, Italia, May 18-22, 1992.
8. G. Garbossa, N. Irrera, P. Pagani, F. Pretolani, S. Spelta, "Real-time engineering simulator of a 320 MWE fossil-fired drum boiler plant for design and training activities", 8th Power Plant Dynamics Control & Testing Symposium - Knoxville, Tennessee, May 27-29, 1992.
9. G. De Michele, S. Pasini, "Simulation of the Impact of Combustion Modifications on Heat Transfer and Nox Formation in Oil-fired Furnaces", Int. Joint Power Generation Conference & Exposition, Atlanta, Georgia, October 18-22, 1992.
10. L. Anzano, R. Cori, G. Guagliardi, R. Mengucci, F. Pretolani, M. Ruscio, S. Spelta. "ENEL combined cycle power plant: design verification and optimization using an engineering simulator" - International Symposium on Improved Technology for Fossil Power Plants New and Retrofit Application - Washington - March 1-3, 1993.
11. G. De Michele, S. Pasini, "Reburning Scaling Via Mathematical Modelling", 7th Topic Oriented Technical Meeting. IFRF/GRI/AFRC, Chicago, Illinois, April 27-29, 1993.
12. R. Cori, G. Garbossa, F. Pretolani, S. Spelta, "Il ruolo degli strumenti di simulazione nella verifica del progetto di impianti per la produzione di energia elettrica", ANIPLA: Giornata di Studio su: "Ingegneria dell"automazione" - Milano 16Marzo 1995.
13. P. Colombo, R. Cori, F. Pretolani, "Power plant manoeuvres optimization and standardization using a set point generator" - 9th Power Plant Dynamic, Control & Testing Symposium, Knoxville, Tennessee - May 1995.

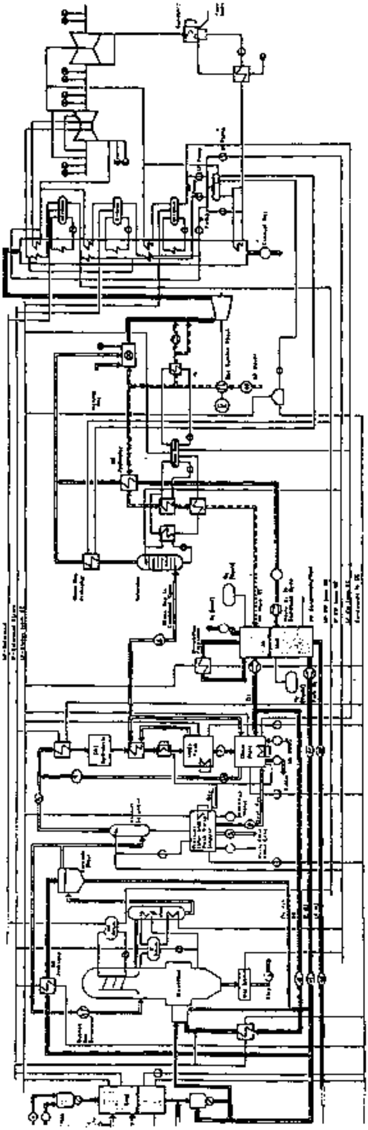


FIG. I

Puertollano IGCC Plant
Process Scheme

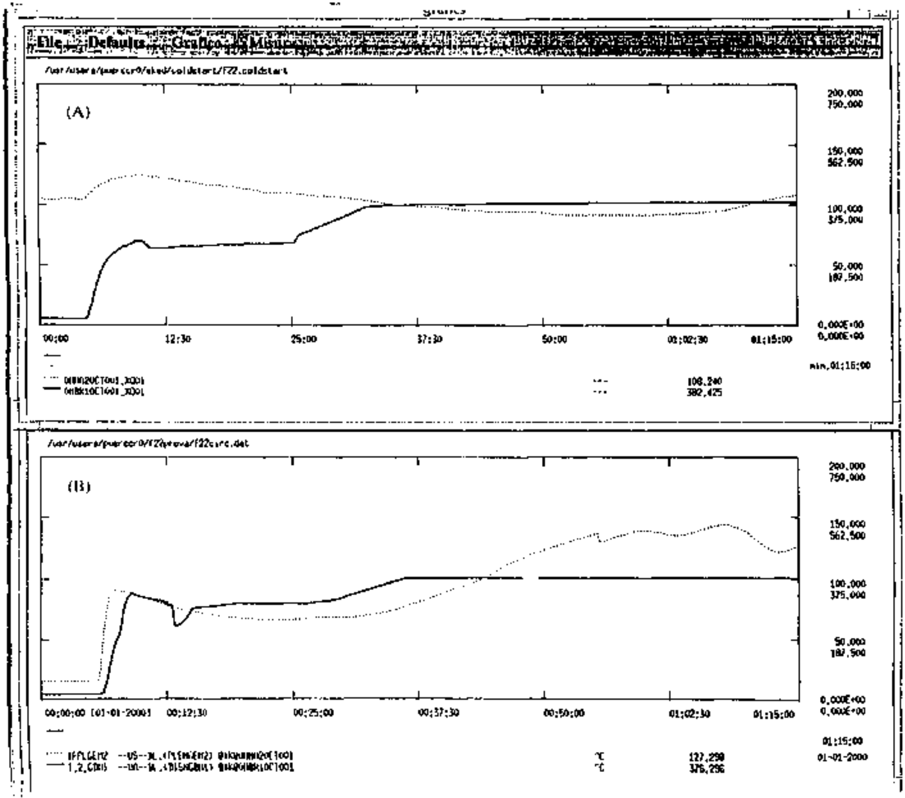


FIG. 2 - Outlet gas temperature at the recovery system inlet/outlet
Comparison of the plant recorded values (A) and the corresponding simulator values (B)

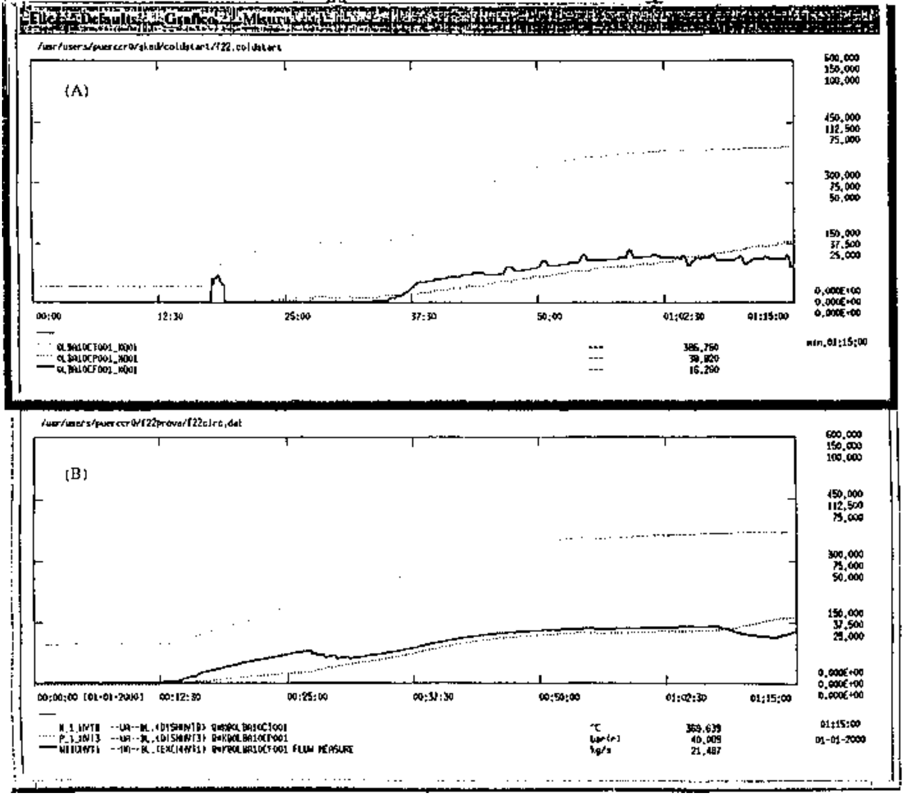
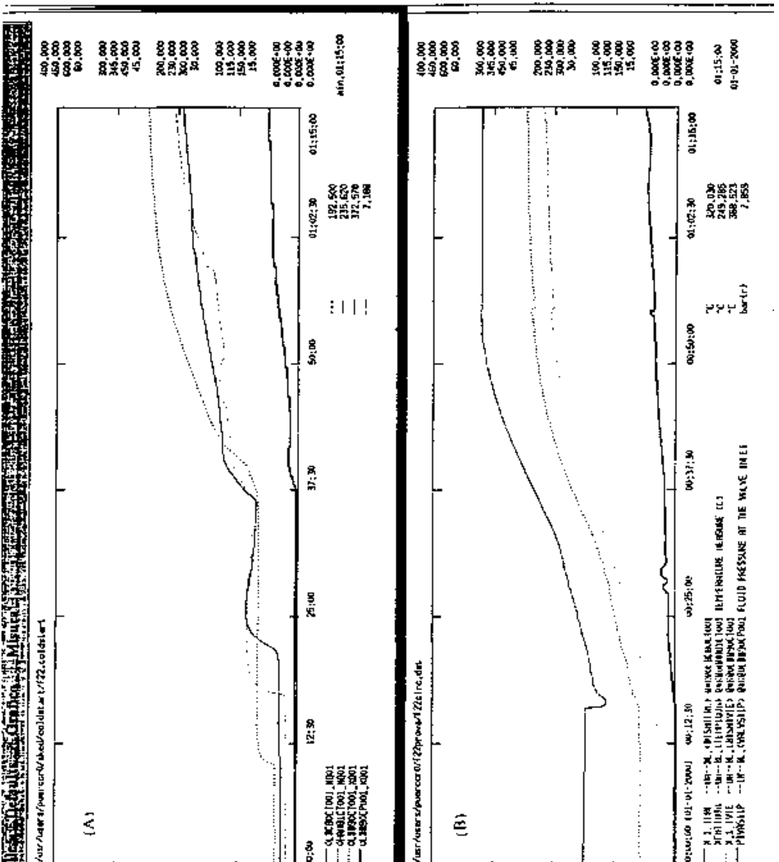


FIG. 3 - Temperature, pressure and mass flow rate of the high pressure steam at the final superheater outlet

Comparison of the plant recorded values (A) and the corresponding simulator values (B)

FIG. 4 - Intermediate pressure exhaust steam temperature at the reheater inlet and outlet
Comparison of the plant recorded values (A) and the corresponding simulator values (B)



FUEL DESULFURIZATION AND SULFUR **RECOVERY FOR IGCC POWER PLANTS**

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Summary

Fuel Desulfurization of an IGCC plant is usually performed in a wet scrubbing unit followed by a sulfur recovery unit to convert the H_2S into elemental sulfur. For the selection of an appropriate solvent high H_2S selectivity, low $C0_2$ coabsorption and high chemical stability are the main criteria. Up to date no solvent has proven to be the optimum for all cases. Due to the wide range of crude gas conditions resulting from different feedstocks and gasification technologies the solvent has to be selected on a case by case basis.

Linde joined in the construction and operation of two pilot plants. The crude gases were very different in composition and represent the whole range of usally encountered H_2S and $C0_2$ concentrations. The first plant, located in Japan, was operated with the new solvent DMPU while for the second plant, located in Germany, the solvent Flexsorb SE PLUS was selected. The pilot plant in Germany additionally comprises a CLINSULF ® sulfur recover,' unit.

The test results were very successful, verifying theoretical calculations and laboratory test results. H_2S concentrations below 10 ppm(v) are achieved with very low $C0_2$ coabsorption of 10 - 15 %. No long term effects like solvent degradation, corrosion or high solvent losses were observed. The results and experiences allow for a reliable scale-up to a commercial plant.

For sulfur recovery the simple CLINSULF ® process demonstrated its high performance. The liquid sulfur product is of excellent quality. The combination of sourgas scrubbing and CLINSULF ® sulfur recovery with recycling the hydrogenated CLINSULF ® tailgas upstream the scrubbing unit results in nearly 100 % fuelgas desulfurization.

1. Scrubbing Units for Sulfur Removal

1.1 Introduction

The feedstock of an IGCC plant contains sulfur in various chemical compounds. During combustion this sulfur would be converted into $S0_2$- Environmental legislation restricts emission of SO_2 to atmosphere. Furthermore SO_2 and sulfuric acid show a very high corrosion potential towards downstream equipment. Therefore, each IGCC plant requires a sulfur removal unit (SRU).

Up to date the usual and only reliable way of sulfur removal is a scrubbing process at ambient conditions. Hot gas desulfurization is not proven on a commercial scale yet. Fig. 1 shows the integration of the desulfurization unit within the IGCC flow diagram. It is located in the crude gas downstream the gas cooling section.

A basic requirement to the scrubbing process is its high selectivity towards sulfur compounds. Coabsorption of CO_2 causes a loss in power plant overall efficiency and an undesired increase in the size of the SRU.

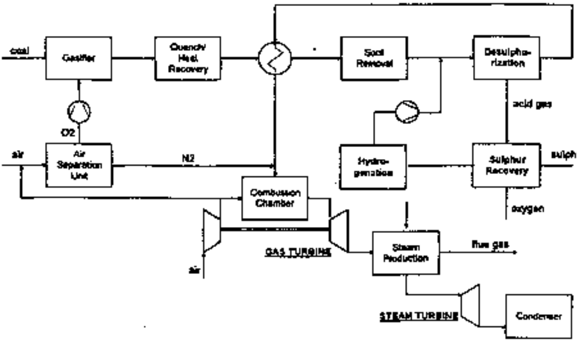

Figure 1 Block flow diagram of an IGCC plant

There are two types of solvents available for SRU. Physical acting solvents make use of the different solubility coefficients of H_2S and CO_2 to achieve selectivity. Chemical acting solvents make use of kinetic influences, whereby additives lower the reaction rate between CO_2 and the solvent. Up to now none of the two basic solvent types has proved to be the optimum for all cases. Each design has to consider the individual feedgas composition, pressure and clean gas requirements. A more detailed discussion of SRU technologies for Combined Cycle Plants was presented in [1].

Fig. 2 qualitatively illustrates the relation between the overall cost of the SRU and the H_2S content in the feedgas to the SRU for both t\pes of solvents. The physical scrubbing process is more favourable for higher H_2S concentrations. The break-even point also depends on other parameters like the CO_2/H_2S ratio or their partial pressure in the feedgas. For example. Fig. 3 demonstrates the solvent pick-up capacity as a function of the acid gas partial pressure and explains the relative cost illustrated in Fig 2.

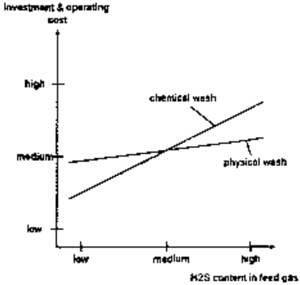

Figure 2 Overall cost comparison

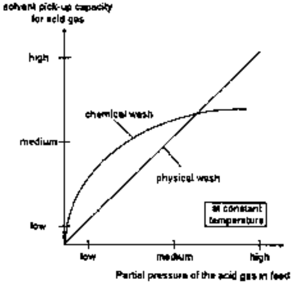

Figure 3 Solvent pick-up capacity

The H2S fraction leaving the SRU is fed to the sulfur recovery unit. In order to meet environmental specifications the tailgas from the sulfur recovery unit requires further treatment before release to atmosphere. Recycling it upstream the wash unit results in a "no emission - configuration".

Today the Rectisol ® process is the world standard for acid gas removal in gasification plants for coal and residues as feedstocks. It is proven for decades in large industrial plants with feedgas capacities up to 1.6 million Nm^3/h. Due to the solvent methanol with its inherent high coabsorption of CO_2 and its low absorption temperature below - 20 °C it is less favourable for IGCC plants, as long as overall plant efficiency aspects are given preference to.

Currently a lot of development work is done to find out the most economic solvent for desulfurization. Linde is involved in the construction and operation of two pilot plants.

The first was located in Japan downstream a Texaco coal gasification unit with a H_2S rich crude gas. Together with the Japanese partner UBE Industries extensive testing was performed. The physical acting solvent DiMethylPropyleneUrea (DMPU) was selected by Linde based on theoretical investigations and laboratory tests.

The second pilot plant is located in Germany at the HTW demonstration plant (High Temperature Winkler gasification) of Rheinbraun, which is an associated company of RWE Energie. The feedgas to the pilot plant is characterized by a very low H_2S content and a rather high CO_2/H_2S ratio. Therefore the chemical acting solvent Flexsorb SE PLUS, licensed by Exxon, was selected. This pilot plant also includes a CLINSULF ® sulfur recovery unit.

1.2 Pilot Plant in Japan scrubbing high sulfur containing gas

UBE is operating a 1000 TPD ammonia plant using the Texaco gasification process with high sulfur containing coal as feedstock. Fig. 4 shows a block diagram of the plant and the integration of the pilot plant. Feedgas to the pilot plant is withdrawn between quench and CO shift units, the desulfurized stream is fed back to an intermediate stage of the CO shift. The H2S fraction produced by the pilot plant is routed to the existing Claus plant.

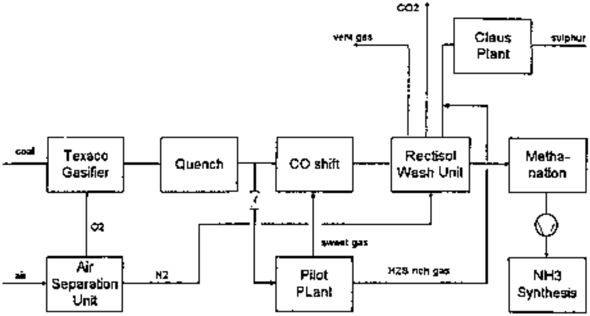

Figure 4 Integration of the pilot plant into the UBE ammonia plant

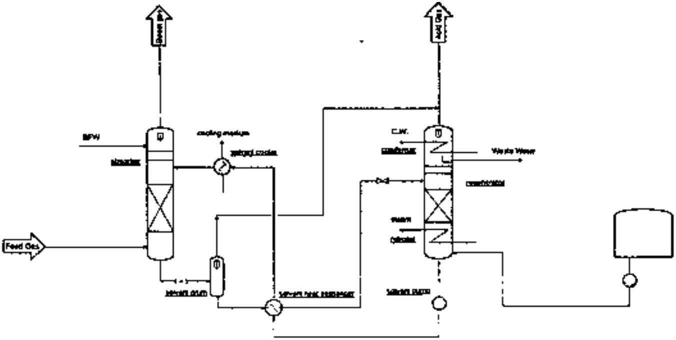

Figure 5 Process flow diagram of a standard wash unit

The process flow diagram of the pilot plant is similar to that of a standard wash process given in Fig. 5. Fig. AI shows a photo of the plant.

The pilot plant was designed to handle 500 Nm3/h of crude gas with a pressure of 38 bar. Average CO_2 and H_2S content had been 20 resp. 1.7 mol-%. Provisions had been made for multiple solvent feed points to the absorber, controlled inlet temperature of crude gas (10-40 °C) and controlled solvent flow rate (0,7 - 1,8 m3/h). For comparison the well known solvent NMP (NormalMethviPyrrolidon) was operated after the DMPU test period.

The plant was on stream from July 1993 to November 1994 without any considerable interruptions. This period was long enough to recognize long term effects. Three campagnes taking each approx. 2 months have been carried out to pick-up a lot of measurement data for various operating conditions.

At all process conditions the H_2S content of the treated gas was below 10 ppm(v). Often 2-6 ppm(v) have been analyzed. The undesired coabsorption of $C0_2$ depends on the operating conditions. Fig. 6 reveals that coabsorption figures down to 10 % of the $C0_2$ in the feedgas that were obtained with DMPU.

Special attention was paid to COS removal. By adding an activator to the DMPU and slightly modifying the process configuration COS removal could be increased from 50 % to more than 90 %. Reduction of COS in the sweetgas to values below 10 ppm(v) was achieved. Still lower residual COS content is possible by increasing the solvent flowrate. However, the drawback is an increase in $C0_2$ coabsorption.

The NMP operation period showed similar results. At identical process conditions the required solvent circulation rate was comparable. The main advantage of DMPU is its significantly higher selectivity. In Fig. 6 the C02 removal rates are plotted for both solvents as a function of the solvent-/ feed gas flow ratio, which was adjusted to drop the H2S content in the sweet gas below 10 vppm.

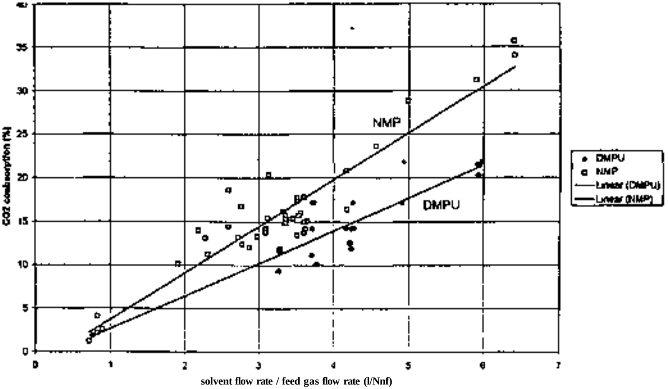

Figure 6 Selectivity comparison of DMPU and NMP solvents

During the approx. 18 months of operation no solvent change has been made except the change from DMPU to NMP. No excessive solvent losses have been observed. The colour of the DMPU inventory remained nearly unchanged showing the excellent stability of the solvent. Plugging was never observed. After careful investigation of the construction materials no corrosion damages could be identified.

Regeneration heat input was kept high enough to achieve sufficient regeneration of the solvent. Determination of the minimum required regeneration energy was not an object of the test program. The insulation losses, even in the absorber, are rather high and depend strongly on the ambient temperature. Therefore recalculation of the enthalpy balance is of low accuracy. Extraction of a reliable value for the required minimum heat input is not possible.

1.3 Pilot Plant in Germany scrubbing low sulfur containing gas

Rheinbraun is operating a HTW demonstration plant using a low sulfur lignite as feedstock. The crude gas leaving the gasifier contains only a few hundred ppm of H_2S. In order to gain experience and to have the most favourable technology RWE Energie, which is an associated company of Rheinbraun, decided to build a pilot plant for sulfur removal and sulfur recovery. Linde joined in the construction of the pilot plant and in evaluation of the test data.

A block diagram of the pilot plant is illustrated in Fig. 7. A small side stream of the prepurified and cooled crude gas downstream the gasifier is passed through a COS hydrolysis and HCN conversion reactor. In order to simulate the higher pressure of an IGCC plant the crude gas feeding the pilot plant is compressed to 23 bar. Downstream a water wash system for ammonia removal the gas is desulfurized in a scrubbing system. The sweetgas is sent back to the HTW demonstration plant while the H_2S fraction is routed to a CLINSULF® pilot plant to recover elemental sulfur. The hydrogenated CLINSULF® tailgas is recycled upstream the feed gas compressor or fed back to the HTW plant.

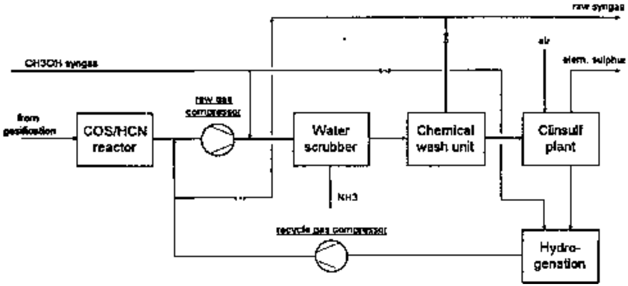

Figure 7 Block diagram of the pilot plant in Germany

The design of the pilot plant allows for testing a wide range of process conditions independent of the actual gas composition of the HTW crude gas. The feedgas flowrate is adjustable between 1500 and 3500 Nm3/h with a $C0_2$ content ranging from 10 to 22 mol-% and a H_2S concentration ranging from 400 to 2500 ppm. The solvent flowrate of maximal 2.5 m^3/h may be distributed between several feedpoints at absorber inlet.

The test period started in March 1995 and was finished in October 1996. Extensive testing with the chemical acting solvent Flesorb SE PLUS licensed by Exxon was performed. Further test series with different solvents will be executed.

Preliminary results show that a H_2S purity in the treated gas below 10 ppm(v) is easily achieved at a reasonable solvent to feedgas flow ratio. The $C0_2$ coabsorption varied between 10 - 20 % with most data in the range from 10 - 15 %. Values below 10 % were observed at special process conditions. The process parameter influencing the $C0_2$ coabsorption are the $H_2S/C0_2$ ratio in the feed, the absorber temperature and the solvent feedpoints to the absorber. Tab. 1 shows the performance data determined for a reference point.

		Feed Gas	Sweet Gas	Acid Gas	Solvent to Scrubber
C02	mole%	10.50	9.35	94,1	
H2S	mole%	0.08	7 vppm	5,9	
Others	mole%	89.42	90.65		
Flow rate	Nm^3/hr	2800	2762	38	
Flow rate	1/hr				1350
Temperature	°C	30	40	35	40
Pressure	bara	23		2	23

Table 1 Performance data of the Pilot plant operated with Flexsorb SE PLUS solvent

Despite the very high selectivity of the Flexsorb solvent the H_2S concentration in the acid gas of an one stage scrubbing unit is too low for treating in a conventional Claus plant. This favours the application of the CLINSULF® process for sulfur recovery.

Due to very low COS concentrations no results about COS absorption or hydrolysis in the wash system could be obtained.

As in the pilot plant in Japan no attention was paid to determine a reliable figure for the minimum required regeneration energy. The energy input is strongly influenced by the insulation losses making it difficult or even impossible to develop such a figure.

The whole Flexsorb SE PLUS test period was performed with the initial solvent filling. Solvent losses were in a moderate range. No alteration of the solvent has been observed showing its good stability. No corrosion damages occured.

2. CLINSULF®-Process for Sulfur Recovery in IGCC Plants

2.1 Basic Features

Today almost all operating sulfur recovery plants are based on the modified Claus process. This process comprises a thermal step followed by two or more catalytical steps. The elemental sulfur is produced via the well known Clans reaction between H_2S and SO_2. It is condensed and separated from the gas stream after each process step.

The thermal and catalytical process steps differ in their temperature levels. The thermal stage operates at a temperature range of 950 - 1600 °C using a furnace to oxidize the H_2S to sulfur. The minimum temperature represents the limit for a stable furnace operation while beyond 1600 °C expensive materials have to be applied.

The catalytic stages usually operate at temperatures in the range of 180 - 350 ° cel. At this temperatures the reaction kinetics have to be accelerated by using catalysts to achieve sufficient efficiencies. The lower operation temperature limit is set by the sulfur dewpoint. In order to avoid sulfur condensation on the catalyst bed, followed by catalyst deactivation, a safe margin from the sulfur dewpoint has to be established.

The thermodynamics of the Claus process reveal that the sulfur conversion rate in the thermal stage increases with increasing temperature while in the catalytic stages the opposit is true. By lowering the process temperature of the catalytic reactor the sulfur conversion rate becomes higher.

With this in mind Linde developed the CLINSULF® process. It is characterized by the application of an internally cooled catalytic reactor. A spiral wound heat exchanger is submerged in the catalyst bed transferring the reaction heat of the exotherm reactions to the cooling medium. Thus the temperatures within the catalyst bed can be kept low and due to the thermodynamic equilibrium high conversion rates within one reactor are achieved.

There are three main types of the CLINSULF® process. Each type has its typical application range characterized by the H_2S concentration in the feedgas and by the process efficiencies expressed in terms of the sulfur recovery rate. Tab. 2 lists the three CLINSULF® process types.

According to the prevailing project conditions each of the above mentioned CLINSULF® types may come into application for IGCC projects.

In the following only the CLINSULF DO ® type plant as installed on a pilot scale in Germany will be described. DO stands for direct oxidation, which means that the oxidation of H_2S to sulfur takes place in the catalyst bed by adding the reaction air or oxygen in "front of the catalytic reactor. No thermal stage is required. This type of process is particularly suitable for coal gasification plants resulting in a low H2S content in the feedgas to the sulfur recovery unit

Type	H_2S in Feedgas	Sulfur Recovery Rate	Main Feature
	vol.-%	%	
CLINSULF DO	1-20	80-95	Without thermal stage
CLINSULF	20 - 100	90-98	Including thermal stage
CLINSULF SDP	20 - 100	98 - 99.5	Sub-dewpoint-plant

Table 2 CLINSULF® process types with basic features

Details on a Clinsulf®-DO plant operated within a Shell Gasification Plant in South Korea are presented in [2]. For a performance report an a CIinsulf®-SDP plant operated in Nynas Refinery in Sweden, reference is made to [3].

2.2 CLINSULF DO® Pilot Plant at the Rheinbraun HTW Demonstration Plant

The acid gas feed to the CLINSULF® pilot plant is a typical loan acid gas. The H_2S concentration is well below 10 vol-% allowing no stable operation of a Clans furnace. Therefore this gas cannot be treated in a conventional Claus plant.

Fig. 8 shows a process sketch of the pilot plant. The acid gas feed leaving the scrubbing section is preheated in El to the necessary reactor inlet temperature of about 200 °C. The required amount of reaction air is preheated in E2 and mixed with the acid gas entering the CLINSULF® reactor Rl. This reactor is equipped with a spiral wound heat exchanger and filled with a commercially available Claus catalyst. The sulfur formed in the reactor Rl is condensed in heat exchanger E3 and separated from the gas stream in D1.

In order to allow for recycling the tailgas to the scrubbing unit the remaining sulfur compounds, i.e. sulfur vapour and SO_2, have to be hydrogenated in the catalytic reactor R2 filled with a commercially available Co-Mo catalyst. The hot gas leaving this reactor is cooled in the quench column Tl before feeding the recycle compressor.

In the pilot plant a hot oil system provides the necessary heating and cooling services including the CLINSULF® reactor cooling. For a commercial plant a steam system is the preferred configuration. The CLINSULF® reactor will be cooled by generation of IP steam.

The pilot plant is designed to treat an acid gas stream of 100 Nm3/h with an H_2S concentration ranging from 0.5 to 10 vol.-%.

2.3 Results

The H_2S fraction leaving the scrubbing process contains traces of NH_3, HCN, CO, H2 and CH_4. Within the test program benzene was adjusted ranging from about 100 to 7000 ppm with an average of 500 ppm. Naphtalene was saturated at acid gas inlet conditions (30 ° cel, 2.0 bar).

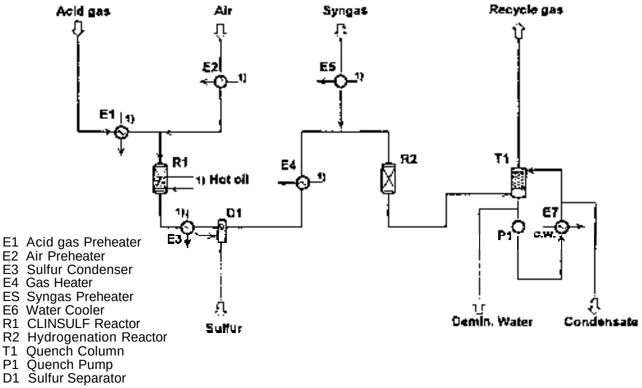

Figure 8 CLINSULF DO® pilot plant in Germany

In order to show the suitability of the CLINSULF DO® process in this application initially a laboratory scale plant with an acid gas feed of 2 Nm3/h was run for about 1000 operating hours. The results were promising. H_2S conversion rates of 90 % were achieved and no catalyst deactivation was observed. The sulfur product was bright yellow with a purity of > 99.95 wt-%. Even at catalyst temperatures as high as 400 ° eel and high benzene content no discoloured sulfur was observed. Cracking of benzene did not occur.

SO_2 was never dectected downstream of the hydrogenation reactor , proving the reliable performance of this reactor.

Investigation of the catalyst after 1000 operating hours, performed by the catalyst supplier and own measurements, revealed a small decrease in catalyst activity at the top of the catalyst layer. The reason for this desactivation may be entrainment of scubbing agent which was observed from time to time. Activity in deeper layers did not decrease. Mechanical catalyst data remained nearly unchanged.

With this positive results a pilot plant was contracted to treat the whole acid gas feed delivered by the upstream scrubbing process (100 Nm3/h). This plant was started up in July 1996. First results confirmed the data of the laboratory plant. Sulfur conversion rates up to 94 % were achieved with the single CLINSULF® reactor.

Targets for the further operation of the pilot plant are the verification and improvement of the process design for a commercial scale plant and the proof of the long term perfonnance of the CLINSULF® and hydrogenation catalysts in this specific application.

} References

[1] Horst WeiB: 'Fuel Desulphurization for GuD Turbines (Combined Cycle)', Linde Reports on Science and Technology 44, 1988

[2] Michael Heisel, F. J. Marold, M. Gvvinner: 'Clinsulf ®-DO for Sulphur Recovery from H2S containing Gas',Linde Reports on Science and Technology 53, 1994

[3] J. Kunkel, P. M. Heisel: 'Clinsulf ®-SDP makes its debut', Sulphur, 243, 1996

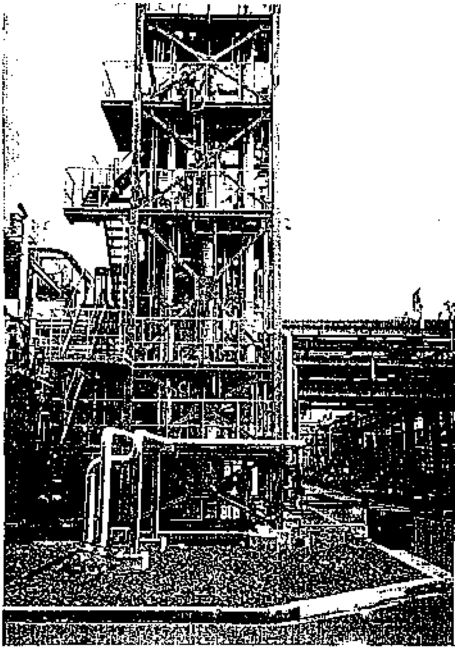

Figure A1 Pilot Plant at the UBE ammonia plant in Japan

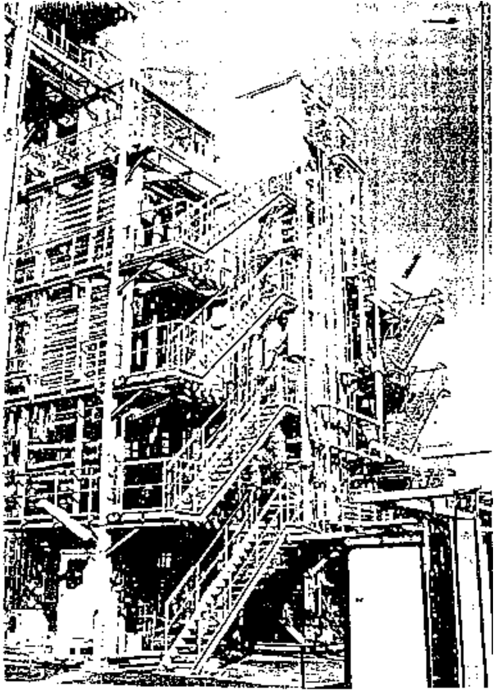

Figure A2 Pilot Plant located in Germany

PURIFICATION AND RECOVERY OPTIONS FOR GASIFICATION

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Abstract

In response to needs created by the gasification industry for power production or chemicals production, UOP* offers three technologies: the Selexol* process for acid gas removal and the POLYSEP* membrane and POLYBED* PSA systems for various separation and purification schemes. This paper discusses the acid gas removal process as well as separation schemes to produce hydrogen and synthesis gas for ammonia, methanol, and oxo-alcohols from gasifier effluent.

Gasification for power generation requires desulfurization. The Selexol process selectively removes hydrogen sulfide and carbonyl sulfide from the synthesis gas by physical absorption. The bulk of the carbon dioxide is left in the gas stream for power generation. Energy consumption is minimized.

Gasification for the production of chemicals and the production of hydrogen requires not only the desulfurization step, but also downstream separation technologies that maximize the production of the required components. The POLYSEP membrane system uses differences in relative permeation rates among stream components to effect a separation. The POLYBED PSA system uses pressure-swing adsorption to remove the impurities from a hydrogen-containing stream and produce a high-purity hydrogen product.

These UOP purification and recovery options have been selected in recent gasification projects to optimize the overall flow scheme. This paper includes examples of the flow-scheme optimization for specific applications.

Gas Separation in Gasification Applications

The gasification of heavy residual oil, petroleum coke, or coal feedstocks to generate a H_2/CO synthesis gas produces a clean fuel for firing in a gas turbine. Gasification is a well-established technology, has broad flexibility of feedstocks and operation, and is the most environmentally friendly route for handling these feedstocks for power production. Electricity can be produced using proven gas-turbine combined cycle technology. For optimal economics, the combined cycle technology is integrated with the gasification system and known as Integrated Gasification Combined Cycle (IGCC).

The key gas separation necessary for IGCC is the selective removal of the H,S and COS from the synthesis gas. Figure 1 illustrates an IGCC flow diagram. For sulfur removal, COS hydrolysis first converts *COS* to H_2S and CO_2 and then a Selexol unit selectively removes H_2S from the synthesis gas. The resulting H_2S-enriched acid gas from the Selexoi unit is suitable for Claus plant elemental sulfur production. The Claus tail gas is recycled to the Selexoi unit for processing, eliminating separate tail gas treating plus incineration. The low sulfur synthesis gas is suitable for the gas turbine.

In addition to electricity and steam production for plant use, refineries can also coproduce H_2 for hydroprocessing needs.[2] This production is accomplished by the addition of POLYSEP membrane and POLYBED PSA units. A portion of the Selexoi synthesis gas product is H_2 enriched by the membrane unit and then purified in the PSA unit to produce a high-purity H_2 stream.

If H_2 production rather than power is the objective for a refinery, then high-purity H_2 is required.[3] The Selexoi unit removes H_2S and, as an option, can also be used for partial $C0_2$ removal. The resulting product synthesis gas is processed to high-purity H_2 in a PSA unit and is now suitable for refining use.

If NH_3 is desired for fertilizer production, the high-purity H_2 is combined with N_2 from the air separation plant for NH_3 synthesis. Subsequently, a portion or all of the NH_3 from the reactor is routed to the urea plant along with the stoichiometric amount of sulfur-free $C0_2$ produced from the Selexol unit for urea production.

Finally, if chemicals production either in combination with power production or as a stand-alone facility is the objective, then the H_2/CO synthesis gas from gasification can be used to make a wide range of chemicals. Although methanol, ammonia and oxo-alcohols top the list, additional chemicals that can be manufactured from synthesis gas include isobutylene, MTBE, acetic anhydride, TAME, acetic acid, dimethyl terephthalate, acetaldehyde and methylamine. [4] For the gas separation steps, the Selexol process for sulfur removal with a POLYSEP membrane and a POLYBED PSA system for H_2/CO ratio adjustment and high-purity H_2 generation are key.

Figure 2 gives an overview of the potential "products" of gasification, and the key gas separation steps using Selexol, POLYSEP membrane, and POLYBED PSA units to prepare the synthesis gas for appropriate downstream processing.

UOP has been awarded contracts for synthesis gas purification at three IGCC projects in Europe since 1994. In two of these cases, the objective is power production only, but in the third case, both power and high-purity H_2 production are required. UOP is also involved in a number of active proposals for other gasification applications worldwide, including chemicals production.

Advantages of Selexol for Acid Gas Removal

Solvent

The Selexol process is a well-proven, stable acid gas removal system based on the use of a dimethyl ether of polyethylene glycol as a physical solvent. The chemical formula is

$$CH_3\text{-O-}(CH_2\text{-}CH_2\text{-O})_N\ \text{-}CH_3 \text{ where } N = 2 \quad 9$$

The system is based on physical solubility, and the driving force is the high solubility of H_2S and $C0_2$

and other acid gases compared to other light gases. No chemical reactions (i.e. acid-base reactions) occur. Higher partial pressures lead to higher solubilities of all components, but the attractiveness of the Selexol system is a favorable solubility for the acid gases versus other light gases. The H_2 and CO have low solubilities in the solvent; for example, $C0_2$ is 75 times more soluble than H_2 and H_2S is 670 times more soluble than H_2.

The Selexol solvent is ideally suited for the selective removal of H_2S. It has a high solubility for acid gases, and the solubility of H_2S is nine times greater than that for $C0_2$. The result of this relationship is that H_2S can be absorbed while the bulk of the $C0_2$ passes through the solvent and into the treated gas. In addition, the Selexol solvent has a relatively high level of solubility for COS, typically removing 45 to 75% or more of the COS in the feed, depending on the level of $C0_2$ removal. A list of the relative solubilities for Selexol is given in Table 1.

The Selexol solvent has many advantages for gasification applications:

- Low vapor pressure for minimal solvent losses.
- High chemical and thermal stability (no reclaiming or purge)
- Nontoxic for environmental compatibility and worker safety
- Noncorrosive for mainly carbon steel construction
- Nonfoaming for operational stability
- Refrigeration not required for economical, selective H_2S removal

  High $H_2S/C0_2$ solubility ratio for enhanced selective H_2S removal:

    Acid gas sufficiently enriched in H_2S (> 35 vol-%) for Claus plant

    Low $C0_2$ removal (-20%) for greater mass flow to the downstream turbine and improved cycle efficiency.

  Compatibility with gasifier feed gas contaminants:

    High solubility for HCN and NH_3 allows removal without solvent degradation.

    Good solubility for COS permits high COS removal for selective H_2S removal.

    High solubility for nickel and iron carbonyls allows for their removal from the synthesis gas. This could be important to protect blades in downstream turbine operation.

High solubility permits processing of the reduced Claus sulfur plant tail gas in the Selexol unit, thus eliminating the tail gas treater and incinerator.

Process

The process consists of countercurrently contacting the feed gas with lean solvent in an absorber at high pressure and lower temperatures so that H_2S and COS and a desired level of $C0_2$ are absorbed into the solution. Regeneration of the rich solvent is accomplished through a combination of one or more flashes at reduced pressures followed by thermal regeneration with steam stripping at elevated temperatures and low pressure. A lean solvent-rich solvent heat exchange is standard for heat conservation. The regenerator overhead vapors (acid gas and steam) are routed to a condenser plus knockout drum, and the condensed water is returned to the unit to maintain water balance. The high- pressure flash gas vapors are compressed and returned to the absorber for greater H_2 and CO recovery and to provide H_2S-enrichment of the acid gas for the Claus plant. Any lower-pressure flash vapors are rich in $C0_2$ and are either vented or used for urea production, for example, depending on the application.

The complexity of the process flow scheme directionally increases as a result of the following product gas and feed gas specifications:

- Higher CO_2/H_2S ratio in the feed gas
- Higher $H_2S/C0_2$ ratio in the acid gas to elemental sulfur production
- Lower H_2S and COS concentration in the product synthesis gas

Simultaneous production of an H_2S-enriched acid gas to Claus and $C0_2$ to urea or vent.

A typical Selexol flow diagram for selective H_2S removal in IGCC applications is shown in Figure 3.

Experience

The Selexol process has been used commercially for 30 years and has provided reliable and stable operations. Relevant experiences for gasification are as follows:

- About 50 Selexol units have been successfully commissioned for steam reforming, partial oxidation, natural gas, and landfill gas. Of these, 10 have been for heavy oil or coal gasifiers.

- The 100 MW Texaco /Cool Water (California) 1,000 t/d coal gasifier plant for IGCC demonstration was operated continously for about five years in the 1980s. The Selexol unit performed extremely well and had a high on-stream factor. The process delivered H_2S-enriched acid gas to Claus while removing 20 to 25% of the $C0_2$ and treating a high CO_2/H_2S ratio feed gas. The basic process flow scheme is given in Figure 3.

- The TVA/Muscle Shoals (Alabama) 200 t/d coal gasifier demonstration plant was operated continuously for about five years in the early 1980s. It employed a Texaco gasifier, a COS hydrolysis unit, and a Selexol unit to convert coal to clean synthesis gas and $C0_2$ as an alternative feed to an existing ammonia-urea plant. The COS hydrolysis and Selexol units were stable and had a high on-stream factor. The Selexol unit delivered an H_2S-enriched acid gas to elemental sulfur production, a pure (< 1 vppm total sulfur) synthesis gas to NH_3 synthesis, and removed part of the $C0_2$ to provide high-purity $C0_2$ for urea production.

Carbonyl Sulfide Hydrolysis

Although the Selexol solvent has a high solubility for COS, it is not high enough to rely solely on the Selexol system for the removal of all the COS in cases for selective H_2S-COS removal. The intermediate solubility of COS compared to H,S and $C0_2$ (Table 1) leads to excessive $C0_2$ coabsorption when removing higher concentrations of COS.

To ensure that COS is removed, a COS hydrolysis unit is added before the Selexol system for IGCC service. In the fixed-bed catalytic COS hydrolysis system, COS is reacted with steam to form H_2S and $C0_2$. The level of COS in the feed gas is generally in the 300 to 1000 vppm range and is reduced in the COS hydrolysis system to the 10 to 25 vppm range. The COS hydrolysis catalyst technology is established and demonstrated in commercial operations.

In non-IGCC service, such as gasification for H_2 or NH_3 production, the presence of a CO shift converter obviates the need for a separate COS hydrolysis unit. The CO shift and COS hydrolysis are accomplished simultaneously in the shift converter.

POLYSEP Membrane: Concentration of H_2 and Syngas Ratio Adjustment

Membrane systems have been proven in a variety of light gas separations and find wide application in the chemical and refinery industries. More than 30 membrane systems have been supplied by UOP and brought on-stream since 1985. Applications include the adjustment of H_2/CO ratio, upgrading of methanol purge gas, purification of H_2 from refinery offgases, and concentration and purification of CO. In gasification applications, membrane technology can be used to concentrate H_2 from the synthesis gas or to adjust the H_2 /CO ratio.

The POLYSEP membrane technology is based on a polymer packaged as a hollow fiber. A differential pressure is established across the membrane. All gases permeate from the high-pressure (feed) side of the membrane to the low-pressure (permeate) side, and the difference in the permeation rates of gases provides the separation. Molecules that permeate quickly, such as H_2 He, $C0_2$ and H_2S, can be separated from molecules that permeate more slowly, such as CO, CH_4, and N_2. The separation concept is illustrated in Figure 4 for a case where hydrogen is being concentrated into the permeate product.

A photograph of a POLYSEP membrane system is shown in Figure 5. The membranes are packaged as modules and supplied in a skid-mounted system. The use of hollow-fiber membranes allows a large surface area to be packed in a given number of modules, enhancing recovery of the desired product while minimizing cost. All equipment is carbon steel and the control components are straightforward. Utility consumption is minor for POLYSEP membrane units.

POLYBED PSA: H_2 Purification

POLYBED PSA units are used in a wide range of applications and include the purification of steam reformer effluent, refinery offgas streams, gasification effluent, methanol purge streams, ammonia synthesis and ammonia plant purge gases, and numerous other applications. More than 500 POLYBED PSA units have been supplied since 1966 for the production of high-purity hydrogen. UOP has also supplied integrated PSA and membrane systems for the production of chemicals such as oxo-alcohols from synthesis gas.

The POLYBED PSA systems operate by adsorbing light gases, such as CO, CO_2, and CH_4 from H_2-containing feed streams onto a fixed bed of adsorbents. Adsorption of the impurities occurs at a relatively high pressure. Hydrogen is adsorbed in only small amounts and can, therefore, be recovered at high pressure and purity after passing through the bed. Regeneration is accomplished by reducing the pressure on the adsorbent to desorb the impurities into the tail gas.

The POLYBED PSA process operates on a cyclic basis, with individual adsorber vessels cycled between adsorption and desorption steps. Multiple adsorbers are used to provide constant feed, product and tail gas flows. Adsorbents are selected for each application based on the type of impurities present in the feed stream.

Two of the advantages of the PSA process are its ability to remove impurities to any level (for example, levels in parts per million, if desired) and to produce a high-purity hydrogen product. Typical purities for PSA hydrogen products range from 99 to 99.999 vol-%. High hydrogen purity is often of benefit to downstream processes, and because of this, most PSA units are designed to produce these high purities.

The utility requirements to operate the POLYBED PSA system are minor and consist of a small amount of electric power and instrument air to actuate the control valves associated with the system. The overall operating cost of the system is a result of the pressure swing for regeneration, which produces a tail gas depleted of H_2 at low pressure. In most cases, this tail gas is used as fuel, which may or may not require compression.

A photograph of a large POLYBED PSA system is shown in Figure 6. Skid-mounted construction is used for the PSA units, which simplifies installation and commissioning.

IGCC Facilities for the Production of Power and H_2 -

Interest in the use of gasification technology to produce power has grown dramatically. In most cases, relatively heavy feeds are gasified and range from coal and coke to heavy refinery streams. After heat recovery and soot removal, a high-pressure stream of about 35 to 50 vol-% H, with similar concentrations of CO results. Although the fundamental purpose of the IGCC facility is the production of power, a fairly large quantity of high-purity H_2 can be extracted. The amount of H_2 produced is typically in the range of 20,000 NmVhr to 50,000 Nm^3/hr (22 to 56 MM SCFD). Also, a ratio-adjusted H_2/CO synthesis gas can be produced from this stream without compromising the capabilities of power production or reliability.

The flow scheme for power and H_2 is depicted in Figure 7. The gasifier effluent is sent to a COS hydrolysis system followed by a Selexol unit for the removal of H_2S. The scrubbed gas is fed at available pressure to a POLYSEP membrane unit that concentrates H_2 in the permeate. The permeate stream is routed to a POLYBED PSA system, and high-purity H_2 is produced. The membrane nonpermeate, which contains the bulk of the CO, CH_4 and $C0_2$ at high pressure, is routed to a gas turbine for power generation.

The fundamental decision affecting the design of the downstream separation technologies is the operating pressure of the gasifier. High-pressure operation in the gasifier permits easier and lower cost separation as a result of higher acid gas partial pressure in the Selexol solvent and easier and more-selective permeation in the membrane unit. Thus, a high-pressure gasifier minimizes the size, cost, and compression requirements for the downstream purification units. Most IGCC designs are based on one of two gasifier effluent pressures: a low-pressure design in the range of 24 to 28 barg (350 to 400 psig) or higher-pressure designs at 48 to 69 barg (700 to 1000 psig).

The Selexol design in IGCC units for power production selectively removes H_2S plus COS to typical levels of 30 to 50 ppm (or less) in the treated gas. This residual sulfur specification was sufficient to meet overall sulfur emission regulations for the three IGCC units awarded to UOP in Europe.

For a typical case as presented in Table 2, the gasifier design is based on heavy oil and is integrated with an oil refinery for power and H_2 production. After heat recovery and soot removal, the gasifier effluent undergoes COS hydrolysis and cooling to 38°C (100°F). The gas then enters the Selexol unit at 51 barg (740 psig) and 38°C (100°F). The desired performance is minimum $C0_2$ removal and essentially complete removal of carbonyls and sulfur. In this single-stage absorber, the sulfur components are selectively removed. Treated gas is suitable for downstream operations.

The acid gas removed by the Selexol unit is sufficiently high in H_2S concentration to be used as feedstock to a Claus plant. Although the specific designs can vary the H_2S level, it is typically in the 35 to 60 vol-% range. One further option that Selexol allows is a recycle of the tail gas from the Claus plant back to the Selexol feed. The resulting closed loop system eliminates the Claus tail gas treating unit and the acid gas incinerator but requires the addition of a recycle compressor.

The treated gas from the Selexol unit contains H_2 and CO in roughly equal proportions along with a low level of $C0_2$. Small amounts of sulfur, water, and light inert gases are also present along with trace components. This gas is sent to the POLYSEP membrane unit at the available pressure of 50.3 barg (730 psig).

The membrane unit enriches the crude H_2 from the 35 to 50 vol-% range to a H_2 concentration of 70 to 90%. The amount of H_2 recovered in this permeate stream can be adjusted depending on the membrane system design and the desired amount of H_2 production, but, in most cases, less than 75% of the H_2 is permeated. Higher amounts of H_2 can be removed, but as larger amounts of H_2 are removed, additional H_2 has less driving force to permeate. The result is a larger, more-expensive membrane that also produces H_2 of lower purity.

The permeate pressure of the membrane has a substantial impact on the design. This pressure is generally chosen to allow the H_2 product pressure to match the refinery H_2 header pressure, typically 17 to 24 barg (250 to 350 psig). A pressure of 24 barg (350 psig) requires more than twice the membrane area as compared to a permeate pressure of 17 barg (250 psig), and so the cost of the membrane system is increased. Operation with a lower permeate pressure improves the performance of the membrane and the result is a higher H_2 purity.

The permeate stream from the membrane unit is sent to a POLYBED PSA system. The PSA unit operates at a higher H_2 recovery rate and lower cost at 24 barg (350 psig) than 17 barg (250 psig), although operation over this entire range is acceptable. A major design consideration is the PSA tail gas pressure, which has a significant influence on hydrogen recovery. Because the H_2 production capacity for the IGCC facility is fixed, a lower recovery rate in the PSA unit requires a higher feed rate to a larger, more-expensive PSA unit.

Although the overall cost of H_2 production is a trade-off between the various parameters described, important criteria in an optimization are:

1. *Gasifier operating pressure.*

2. *H_2 quantity required:* As the membrane and PSA units operate by extracting H_2 from a gasifier volume that is fixed, a higher quantity of H_2 product requires larger membrane and PSA units.

3. *H_2 use pressure:* Low H_2 pressure lowers the permeate pressure for the membrane unit and minimizes the size of the membrane.

4. *PSA tail gas:* Lower tail gas pressures minimize losses of H_2 and CO to fuel and minimizes membrane and PSA cost.

The use of membrane and PSA units to produce high-purity H_2 in IGCC applications results in a system with advantages over a PSA separation unit alone. The membrane system takes advantage of the available pressure to enrich H_2 to a desirable concentration, and this enriched H_2 allows for a relatively small PSA unit with minimal loss of light gases into the PSA tail gas as compared to the case of a PSA operating alone.

Reliability Considerations for IGCC

Electric power production requires extremely high levels of reliability based on the needs of consumers and plants that rely on the electric supply. In the design of Selexol systems for high reliability, attention is given to a flexible design to handle the inevitable variation in the Selexol feed sulfur levels as a result of changes in the raw feedstock as well as the ability to operate at desired turndown rates. The Selexol units at the Texaco coal-based gasifiers of Texaco/Cool Water and TVA/Muscle Shoals demonstrated extremely high on-stream factors in single-train designs.

In the design of the COS hydrolysis unit, guard beds are used to greatly extend the main bed life and increase reliability. For large gasifiers, parallel guard and main bed combinations are employed.

POLYSEP membrane systems contain no moving parts and have proven themselves to be highly reliable. POLYBED PSA units also operate with on-stream factors that approach 100% through the use of reliable components, selectively chosen redundancy, and the ability to continue operation in the event of a valve or instrument malfunction.

Gasification for Hydrogen Production

The gasification flow scheme for H_2 production is shown in Figure 8. The gasifier effluent, which contains H_2 and CO, is directly quenched with water and passed to a CO water-shift reactor, where CO reacts with water to produce additional H_2 and CO_2. Because of the CO shift reactor, a separate COS hydrolysis unit is no longer needed. The downstream Selexol unit selectively removes the H_2S and residual COS to low levels and provides an H_2S-enriched acid gas stream for the Claus plant. This selectivity is critical because the syngas after CO shift has a high CO_2/H_2S ratio and the solvent system has to be highly selective to accomplish this separation.

The effluent stream containing H_2 CO, and CO_2 would in most cases be sent directly to a PSA unit, where the impurities are adsorbed and high-purity H_2 is produced. The PSA tail gas is available at low pressure as a source of fuel. Because of the low heating value of this PSA tail gas stream as well as its low pressure, typically 0.3 to 1.0 barg (5 to 15 psig), a waste heat boiler is normally designed to combust this tail gas stream for the production of steam.

The selection of a gasifier for H_2 production in place of a steam reformer facility is based on an analysis of a number of technical variables. However, the key issue is availability of heavy residual oil or petroleum coke with no other viable use for these materials.

The advantages of a gasifier for the production of H_2 include little or no pretreatment of the feedstock and the availability of a high-pressure hydrogen product of up to about 69 barg (1000 psig). The gasification technology does not generate any nitrogen oxide, carbon dioxide or sulfur oxide emissions, so air pollution control equipment is not required on the stack.[7] Even though steam reforming has been, and is expected to continue to be, the dominant technology applied to H_2 production, site-specific economics will allow gasification to compete.

Gasification for the Production of Ammonia/Urea

The attractiveness of gasification for NH_3 production is due to its ability to generate synthesis gas from a wide variety of feedstocks, including coal and heavy hydrocarbons.

The traditional design of a gasifier for the production of NH_3 synthesis gas is oxygen-based. The gasifier effluent uses a direct quench followed by scrubbing for soot removal and the CO shift reaction. The acid gases are removed to low levels using absorption technology and final purification to remove argon, CO and CH_4 is made with a liquid N_2 wash cold box.

As an alternative to the traditional route, a POLYBED PSA unit can be used to remove the CO_2 CH_4 and CO in a single step. The bulk of the argon can also be removed. The product from the PSA unit is high-purity H_z which has substantial benefits in the downstream NH_3 converter. These benefits include a dramatically reduced purge requirement; a higher conversion rate in the reactor, less reactant losses; and the associated reduction in operating pressures, compression requirements, and capital cost. The Selexol-PSA combination is also considerably less expensive and easier to operate than the traditional route of solvent desulfurization and N_2 wash.

One consideration when using a PSA unit is a loss of approximately 10% of the H_2 contained in the PSA feed gas to fuel. To minimize H_2 losses, a POLYBED PSA alternative can be applied. In this system, N_2 produced in the air separation unit is introduced into the PSA at low pressure to regenerate the adsorbent. The main impact of this N_2 purge is to reduce the H_2 losses to fuel to approximately 5%. An additional advantage results from the low-pressure N_2 being pressurized via the PSA process steps to PSA product pressure. This "free" compression reduces the flow rate required from the N_2 compressor. This technology has been demonstrated in a commercial plant. A flow scheme of a gasifier for ammonia synthesis gas production using a N_2-purged PSA system is shown in Figure 9.

Production of Methanol via Gasification

Methanol is one of the largest commercially produced chemicals and plays an important role as a chemical intermediate. It is produced by the reaction of H_2 CO, and CO_2 ideally based on a stoichiometric ratio of the reactants as given by the following relationship:

$$(H_2-CO_2)/(CO + CO_2) = 2$$

Limiting the CO_2 content of the methanol synthesis gas may also be desirable.

Gasification based on heavy oils, coal, or coke can provide a desirable feed source for methanol synthesis gas production. The COS hydrolysis and Selexol processes are required to reduce the residual H_2S and COS levels in the synthesis gas to low levels. Gasification of heavy ends results in an insufficient quantity of H_2 and a ratio adjustment is commonly practiced to more closely approach the stoichiometric ratio of reactants. This adjustment can entail a variety of means, including shift conversion of the CO or the use of membrane and PSA units.

The membrane-PSA system can be attractive for methanol synthesis gas production when the main aim of the gasifier facility is the production of power. A membrane permeates H_2 from the high-pressure gasifier effluent, resulting in a stream enriched in H_2 and leaving a nonpermeate stream that is depleted of CO_2. A PSA unit is used to produce high-purity H_2 from the membrane permeate. Mixing the H_2 with the nonpermeate allows the production of methanol synthesis gas of any desired ratio of the reactants, regardless of the gasifier effluent composition. A flow diagram using membrane and PSA units for methanol syngas ratio adjustment is shown in Figure 10.

An advantage of a gasifier for the production of methanol synthesis gas is that it can operate at the pressure required for the methanol reaction. Only the H_2 product from the PSA system requires compression to the methanol synthesis reactor p -essure.

The production of methanol synthesis gas through gasification as a stand-alone facility is generally not attractive when compared to facilities based on steam reformers. However, when associated with the production of power or when associated with the production of purified CO, gasification can be an attractive source for synthesis gas.

Oxo-Alcohol Production via Gasification

Oxo-alcohols are an important chemical and chemical intermediary. In a commonly-used commercial process, oxo-alcohols are produced in a two-step process. The first step requires a one-to-one ratio of H_2 and CO for the production of an aldehyde intermediary. In the second step, this aldehyde is hydrogenated with high-purity H_2 to produce the final oxo-alcohol product. Synthesis gas treatment in a gasifier facility requires H_2S and $C0_2$ removal, careful control of the H_2/CO reactant ratio, and the purification of H_2.

One example of an oxo-alcohol facility that produces both synthesis gas and H_2 based on a specific heavy-feedstock case is represented in Figure 11. For heavy gasifier feeds, the H_2/CO in the gasifier effluent is close to the 1:1 ratio required. For H_2-rich effluent a membrane unit easily removes the excess H_2 to provide the desired ratio with little loss of CO. Heavy feeds that are lean in H_2 require hydrogen addition to produce the 1:1 ratio synthesis gas. Heavy feeds are also deficient in the H_2 needed for hydrogenation and the oxo-alcohol plant requires imported H_2 or its production. H_2 for lean gasifier effluent and for hydrogenation can be produced from CO and water by the water-gas shift reaction followed by purification of the resulting stream in a PSA unit. Other schemes are also possible and would be evaluated based on the feed and product requirements of a particular project.

For natural-gas-based facilities, the use of integrated PSA and membrane units has been well established in oxo-alcohol production.

Conclusion

Interest in gasification for power production, H_2 production, and chemicals continues to grow. The UOP separation technologies of Selexol for acid gas treatment and the POLYSEP membrane and POLYBED PSA systems for synthesis gas ratio adjustment and H_2 purification are leading technologies for treatment of the gasifier effluent. Specific projects require optimization of the process flow scheme. UOP has been selected to provide technology for three IGCC projects in Europe since 1994. Because of its ability to optimize the purification and separation train, UOP is also involved in several active proposals for other gasification applications, including chemicals production.

UOP, POLYSEP and POLYBED are trademarks and/or service marks of UOP. Selexol is a trademark and/or service mark of Union Carbide.

References

1. G. Parkinson and K. Fouhy, "Gasification: New Life for an Old Technology", *Chemical Engineering Magazine,* Mar. 1996, pp. 37-43.

2. G.L. Farina and M. Fontana, "Cogeneration Refinery for Total Residue Destruction", *Hydrocarbon Processing,* Nov. 1993, pp. 52-56.

3. B.L. Shulman, F.E. Biasca, RE. Johnson and R.L. Dickerson, "New Routes - Processes and Uses - for Minimizing Heavy Fuel Oil Production", NPRA Meeting, San Antonio, TX, Mar. 20-22, 1994.

4. D.L. Heaven, "Gasification Converts a Variety of Problem Feedstocks and Wastes", *Oil and Gas Journal, May 27,* 1996, pp. 49-54.

5. J.W. Sweny, "Gas Treating with a Physical Solvent", AIChE Meeting, Anaheim, CA, Mar. 1984.

6. N.R. Udergaurd and V. Berzins, "Catalytic Conversion of COS for Gas Clean Up", AIChE Meeting, Anaheim, CA, Mar. 1984.

7. W. Francis Fong and M.E. Quintana, "Hytex - A Novel Process for Hydrogen Production", NPRA 89th Annual Meeting, Mar. 1991.

8. KJ. Doshi, R.G. Werner, and M.J. Mitariten, "Integration of Membranes and PSA Systems for the Purification of Hydrogen and Production of Oxoalcohol Syngas", AIChE Symposium on Membrane Separations in Chemical Engineering, Mar. 1989.

Table 1

Selexol Solvent Relative Solubility Of Gases

Compounds	**Solubility Ratio**
H_2 (Least Soluble)	1.0
N_2	1.5
CO	2.2
CH_4	5
$C0_2$	75
COS	175
H_2S	670
CH_3SH	1,700
$S0_2$	7,000
H_20	55,000
HCN (Most Soluble)	95,000

Table 2

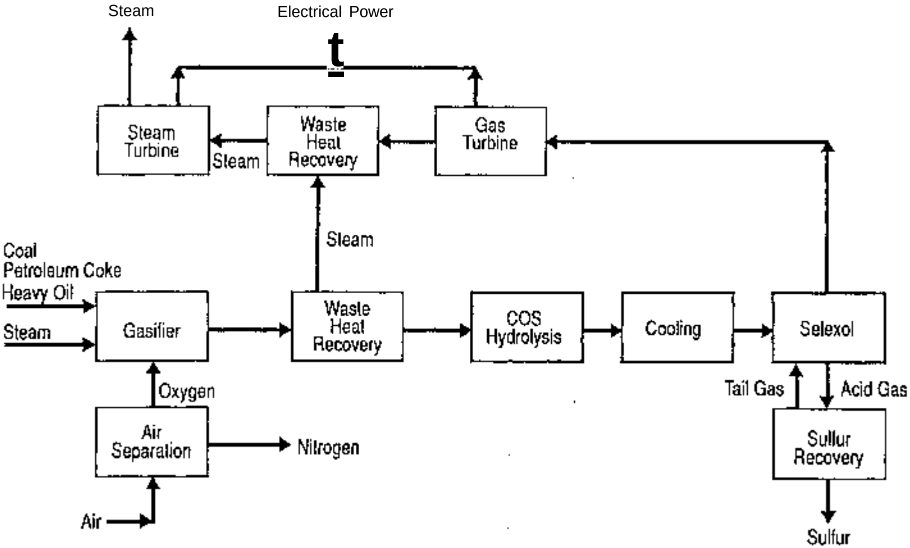
Composition of Key Streams from Purification Section in IGCC Case with H_2 Recovery

	Selexol Feed Gas	Acid Gas to Claus	Membrane Nonpermeate **To Gas Turbine**	High Purity H_2 Product
Flow rate, Nm^3/hr	188,237	8,038	145,588	24,227
MMSCFD	168.6	7.2	130.4	21.7
Pressure, Barg	51.0	1.0	48+	16.5 - 23.5
Psig	740	14	700+	240 - 340
Temperature, °C	37.8	40	45.6	51.1
°F	100	104	114	124
Composition, vol-%				
H_2	43.38	0.22	36.49	99.95+
CO	41.98	0.98	52.35	10 ppm
CO_2	10.30	48.60	8.40	—
H_2S	2.06	46.24	20 ppm	—
CH_4+Ar+N_2	2.15	0.08	2.70	0.05
H_2O	0.12	3.69	0.06	—
COS, vppm	40	427	22 ppm	—

Note: The Selexol performance for this case met the following specifications: 99.9% H_2S removal, 99% carbonyl removal and 21% CO_2 removal. The overall COS removal in the hydrolysis and Selexol units was 98.1%.

Figure

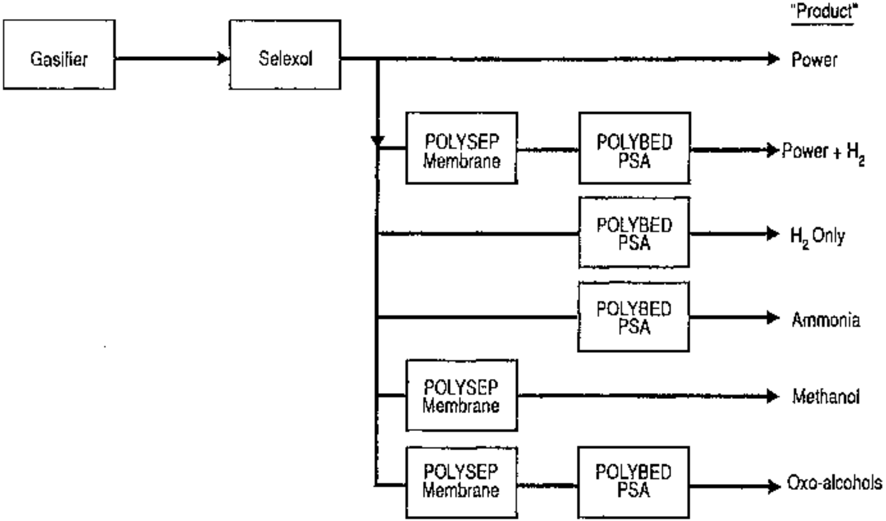
Power Generation (IGCC) Flow Scheme



UOP 2582-1

Figure 2

Purification Options for Alternate Gasification Applications



UOP 2502-2

Figure 3

Basic Selexol Flowscheme for Gasification

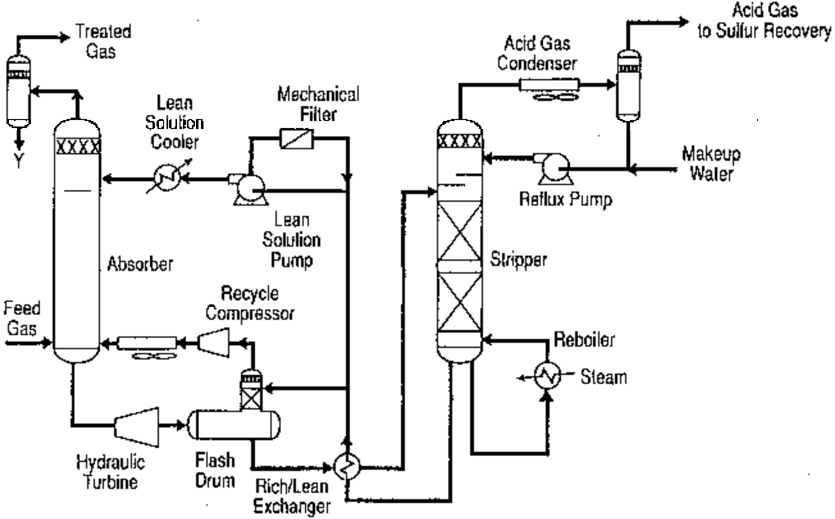
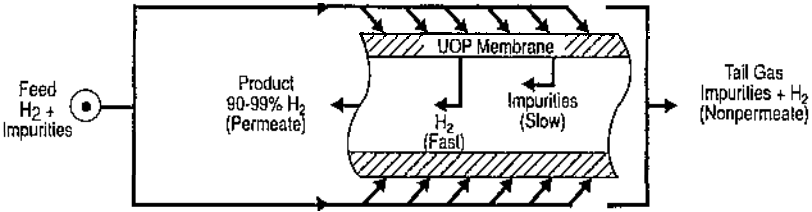


Figure 4

POLYSEP Membrane System

Containment Vessel



Containment Vessel

Figure 5

POLYSEP Membrane Unit

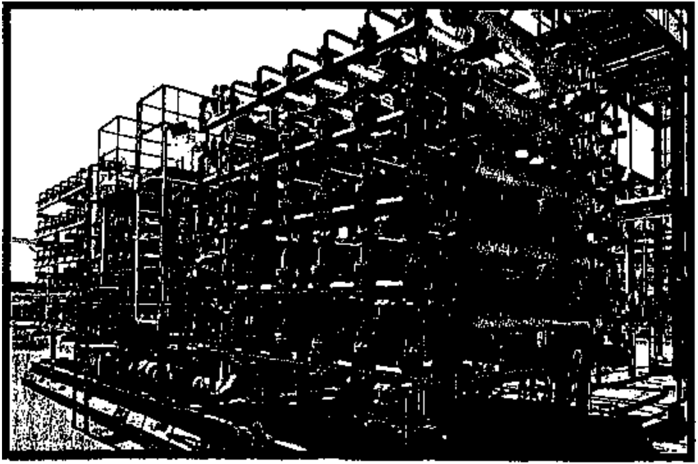


Figure 6

POLYBED PSA Unit

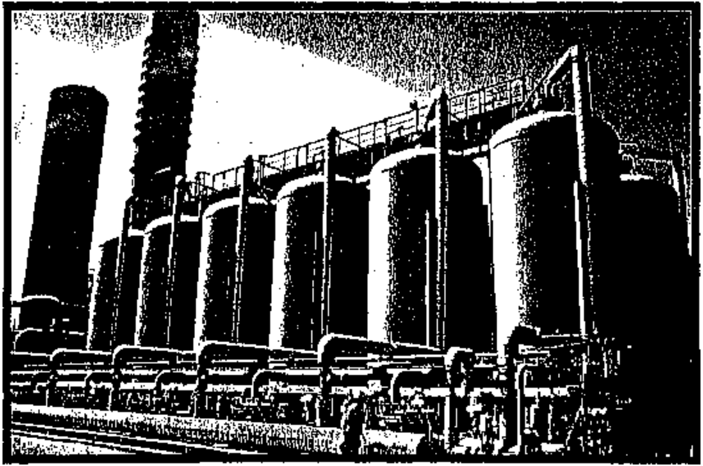


Figure 7

IGCC Scheme with H_2 Recovery

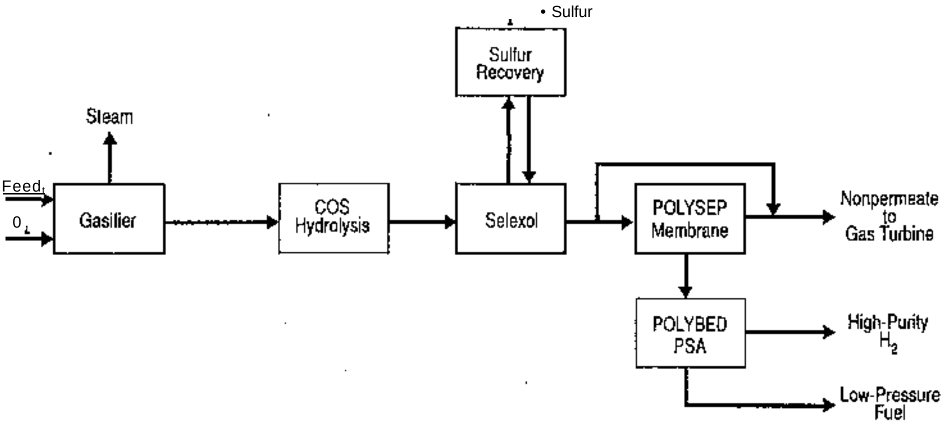


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Figure 8

Gasification Scheme for Hydrogen Production

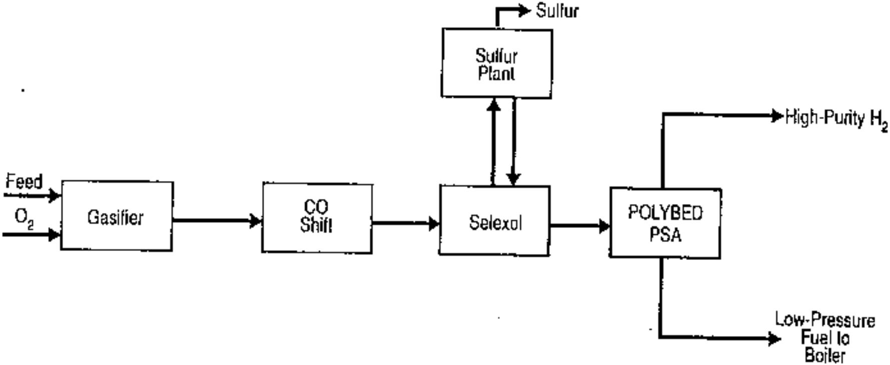

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Figure 9

Nitrogen-Purged PSA Scheme for Ammonia Production

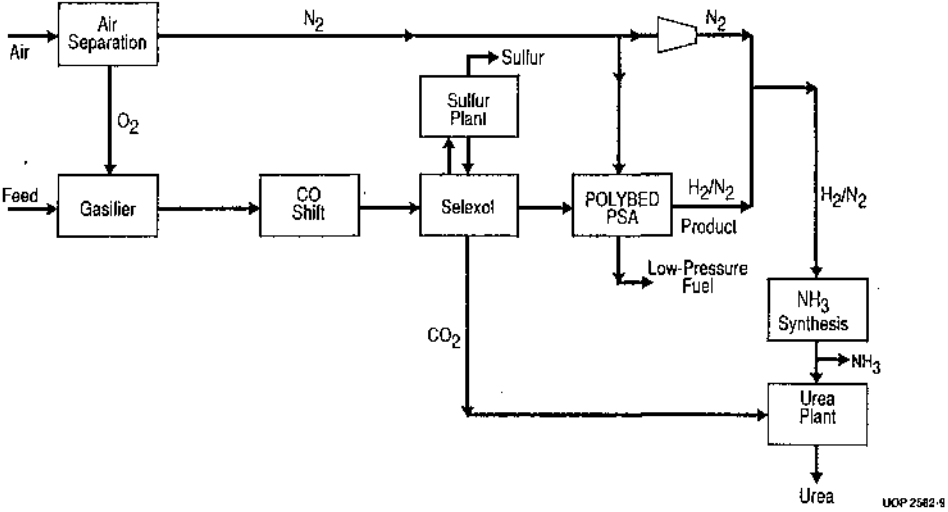


Figure 10

Gasification Scheme for Methanol Production

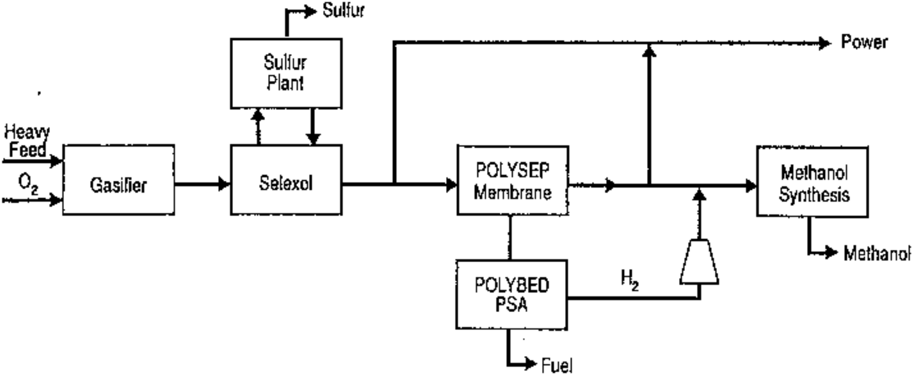
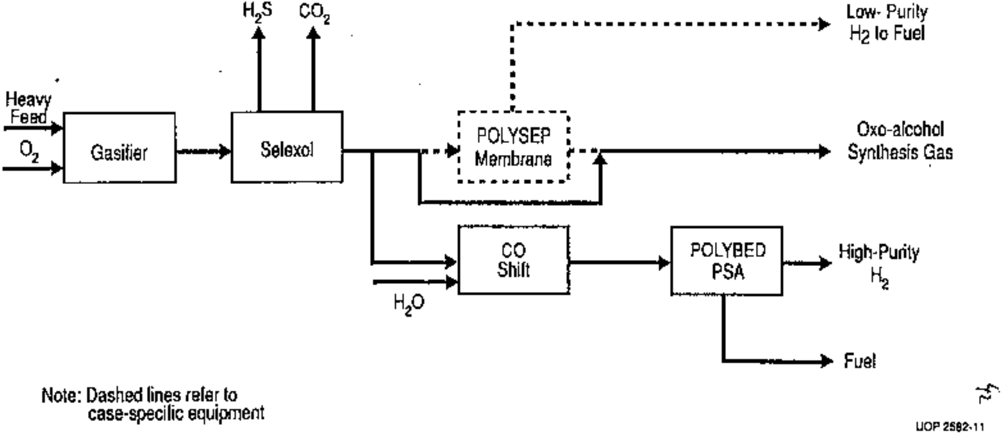


Figure 11

Heavy Feed Gasification for Oxo-alcohol Production



Paper 0056

SELECTION OF THE ACID GAS REMOVAL PROCESS FOR IGCC APPLICATIONS

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Introduction

Typically the acid gas removal section of an IGCC contributes about 10 to 15 % to its capital cost. In addition its selection can influence the operating efficiency but up to 1 % point.

The optimum selection of the AGR process is therefore a complex challenge for designers of IGCC plants. The known and proven absorption technologies are commonly looked upon as a simple utility, whereas with IGCC they play a key role in separating the environmentally sensitive aspects of gasification from the need for clen and efficient power generation and synthesis gas offtake.

The selection is therefore not only based on operating and capital costs, but also depends to a high degree on the feedgas quality and on the required product gas purity, i.e. components to be removed and the destination of the purified gas.

This paper reviews the following topics in association with IGCC applications :

- Comparison of available classes of AGR processes on the basis of a typical 500 MW IGCC project.
- The choice of Rectisol for AGR in the Pernis refinery upgrading project (IGCC plus hydrogen production)
- Integration of an oxygen-blown Claus unit into the acid gas removal system.

Selection of the acid gas removal process

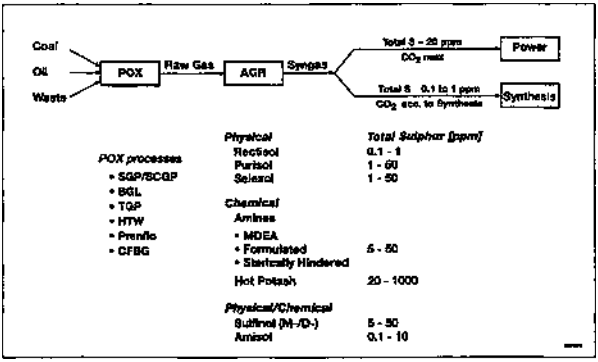
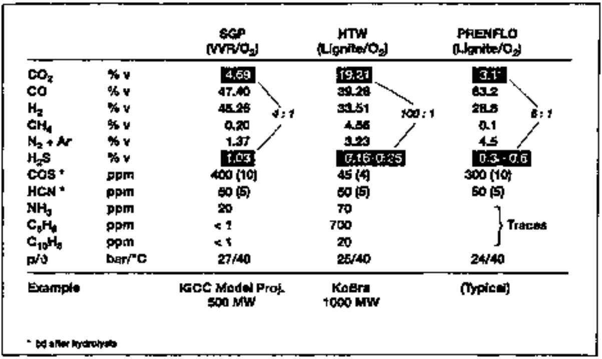


Figure 1 : AGR = Switch-box between Source and Destination

The selection of the optimum acid gas removal process is influenced by many factors. Looking at Figure 1 you can see at first glance that the key issues are the source and the destination of the gas. Both set the points for the appropriate AGR process.



		SGP (VR/O_2)		HTW (Lignite/O_2)		PRENFLO (Lignite/O_2)	
CO_2	% v	4.69		19.21		3.1	
CO	% v	47.40		39.28		63.2	
H_2	% v	45.26	4 : 1	33.51	100 : 1	28.5	6 : 1
CH_4	% v	0.20		4.56		0.1	
N_2 + Ar	% v	1.37		3.23		4.5	
H_2S	% v	1.03		0.16-0.25		0.3 - 0.6	
COS *	ppm	400 (10)		45 (4)		300 (10)	
HCN *	ppm	60 (5)		60 (5)		60 (5)	
NH_3	ppm	20		70		Traces	
C_6H_6	ppm	< 1		700			
$C_{10}H_8$	ppm	< 1		20			
p/ϑ	bar/°C	27/40		25/40		24/40	
Example		IGCC Model Proj. 500 MW		KoBra 1000 MW		(Typical)	

* (x) after hydrolysis

Table 1 : Typical raw gases from Partial Oxidation (POX)

Let's have a look at the source first. If feedstocks like coal, heavy oil or waste are gasified by partial oxidation under greatly differing process conditions in processes like Shell Gasification Process (SGP), British Gas Lurgi Gasifier, Texaco Gasification Process, High Temperature Winkler Gasification, Prenflo or Circulating Fluidized Bed Gasification the resulting raw gases differ greatly in quality. This refers to the absolute concentrations of H_2S and $C0_2$. Even more important in our context is that their ratio varies from some 1:4 up to 1:100. Furthermore, hydrocarbon and nitrogen compound formation differ substantially.

The destination of the cleaned syngas is then the next decisive factor for selecting of the best-suited acid gas removal process.

If going for power production only, there is a relative mild requirement regarding total sulphur in the syngas, e.g. not more than 20 ppm. But maximum selectivity is needed: The more $C0_2$ is left in the purified gas, the better.

When using the syngas for synthesis, deep desulphurisation is required. In most cases a removal of the total sulphur down to 100 ppb is preferred. Selectivity is not generally required, $C0_2$ may be removed in bulk or partially left in the syngas according to the needs of the synthesis.

Partial pressure ratio and final sulphur content are prominent criteria, but not the only ones. The evaluation of an acid gas removal step must include the appropriate sulphur recovery concept (SRU). Therefore, it is necessary to evaluate both together.

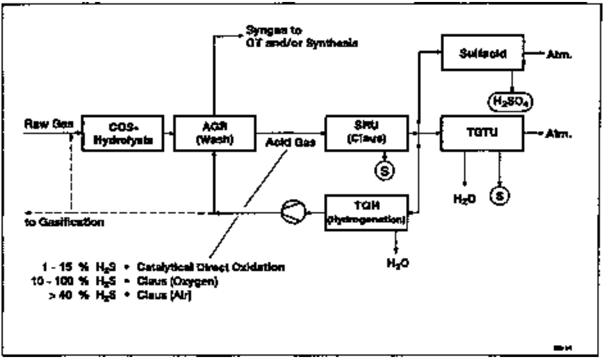


Figure 2 : Options for AGR plus SRU

Figure 2 displays some simplified concepts: Catalytic conditioning, of the rawgas in a COS and HCN hydrolysis step upstream of the wash process followed by processing the acid gas in a Claus unit to produce elemental sulphur. The Claus taiigas typically contains 3 to 6 % of the feed sulphur as H_2S, $S0_2$, COS, CS_2 and S_x.

Therefore a further, mandatory step is either final taiigas desulphurisation and release to atmosphere or catalytic processing to sulphuric acid, e.g. by the SULFACID process. Another alternative is the hydrogenation and hydrolysis of the sulphur species in the taiigas to H_2S and the recycling of those treated gases back to the rawgas or directly to the wash process.

The latter option has the advantage of avoiding an additional emission stream within the total plant and a complete release of the $C0_2$ with the syngas. This results, of course, in an increased hydraulic load on the wash process and therefore requires higher H_2S selectivity.

To explain the above in some more detail, a few results of a power plant study are presented here:

An IGCC Model Project was calculated in depth to obtain a sound basis for adaption to customer needs and to the different scenarios of local markets.

The object of the study is a standalone IGCC power plant based on SGP with partial oxidation of vaccum visbroken residue at 30 bar. Net electrical power output is 500 MW. Acid gas removal in the fuelgas was set to less/equal 20 ppm of total sulphur. Two different desulphurisation options, representing the most prominent classes of AGR processes were chosen, each including the fitting SRU process :

- RECTISOL + CLAUS + SULFACID
- COS-Hydrolysis + MDEA + CLAUS + SULFACID

Cost of Acid Gas Removal		**Rectisol**	**MDEA**
Relative CAPEX	**%**	**100**	**43**
Relative OPEX	**%**	**100**	**46**
Cost of AGR plus SRU			
Relative CAPEX	**%**	**100**	**90**
Relative OPEX	**%**	**100**	**79**
Influence of AGR selection on total IGCC			
Relative CAPEX	**%**	**100**	**99.1**
Sulphur Recovery	**%**	**99.9**	**99.8**
Total Efficiency	**% (LHV)**	**45.5**	**45.4**
Cost of Electricity	**Pfg/kWh**	**5.5**	**5.5**

Table 2 : Summarized results of the IGCC Model Project 500 MW / SGP at 30 bar.

Looking at the cost of the actual AGR unit alone, this tips the scales strongly in favour of the chemical process.

But when enlarging battery limits to include the necessary SRU with the appropriate tailgas treating, the picture changes significantly.

Finally looking at the complete IGCC, it is remarkable to recognize only minor differences:

- Investment costs vary in a narrow band of 1 *%.*
- Total efficiencies are all high and in a narrow band , i.e less than 0.5 %.
- Sulphur recovery rates are all excellent, even 99.9 % are achieveable satisfiing the most stringent specification seen so far.
- Electricity production costs are attractive in all cases.

Often there is another aspect of interest: Operating the gasification at elevated pressure could be beneficial to the final product gas use. For example saving compression needs for syngas used at high pressure like methanol syngas or hydrogen in a hydrocracker.

Higher acid gas partial pressures favour physical absorption processes. In case of an IGCC power plant this would allow overcoming higher capital and operating costs of a physical AGR versus a chemical AGR process.

Having a gas expander in the cleaned syngas has the following, but not only benefits:

- Generating extra power (approx. 1 % of total IGCC power)
- Covering the necessary refrigeration needs of the physical absorption process
- Finally adding some 0.8 %-points to the overall IGCC efficiency

)

Highpressura SGP* {60 *bar) with Gasexpander* *Cost of AGR plus SRU		**Rectisol**	**Purisol**	**MDEA**$^{1)}$
CAPEX	**[Mio DM]**	**114**	**131**	**93**
A CAPEX	**[Mio DM]**	**21**	**38**	
OPEX	**[DM/h]**	**-415**	**-253**	**390**
A OPEX	**[DM/h]**	**- 805**	**-643**	
A OPEX	**[Mio DM/a]**	**- 6.44**	**-5.14**	
Pay out time	**[a]**	**3.3**	**7.4**	
IGCC total Eff.	**[%** ***UN]***	**46.1**	**46.0**	**45.4**

[1]) ***no expander possible because of water-saturated cleangaa***

Table 3 : Summarized results of the IGCC 500 MW / SGP 60 bar with gas expander

Table 3 summarizes the results of a study based on the same scenario as described before, but using SGP at 60 bar and a gas expander. In case of the chemical AGR process a gasexpander is not feasable due to ice formation during expansion.

You can see at a glance that the additional capital needed (Delta CAPEX) for a physical AGR process compared to the chemical process, will be paid back by the operational cost credits (Delta OPEX) within 3 to 7 years. Whether these payout times are attractive or not will depend on the philosophy of the investor.

To summarize:

- The selection of a chemical or physical AGR process within an IGCC-complex will be primarily governed by:
    - Feedstock and POX-process
    - Stand-alone plant vs plant tied into an existing, industrial environment
    - Local emission allowances, e. g. trading of environmental permits in the US ("cheese-cover policy")
    - Use of the cleaned syngas
- For a power plant, without synthesis gas offtake, the choice of the AGR is of minor importance, but it is worthwhile considering the option of physical absorption at high pressure.
- For a plant with synthesis gas offtake, deep desulphurisation is needed. This can normally only be achieved without additional process steps when using a physical acid gas removal process like RECTISOL®.

The choice of RECTISOL* for acid gas removal in the Pernis refinery upgrading project (PER^{+})

The next topic deals with an IGCC integrated into an existing refinery where the products, hydrogen and power, are defined and quantified by the upgrading needs of the whole system.

In a refinery heavy residues are frequently generated as by-products. Its quality depends on the refinery conversion concept, but it is always rich in sulphur and metals. The Shell Gasification Process (SGP) is a very versatile process with a long track record of well performing plants for the conversion of even the heaviest refinery residues to clean syngas.

Most of the existing SGP units are designed to produce ammonia or methanol; currently the focus is more on power and/or hydrogen production. The latter is the most attractive gasification product in a refinery environment for consumption in hydrocracking or hydrodesulphurisation units. Often, however, the amount of residue for gasification and the refinery hydrogen requirements leave a surplus of syngas, which can be used for high-efficiency combined cycle power generation and power delivery over the fence. A typical example of such a project is the PER+ refinery upgrading project at the Shell Pernis refinery near Rotterdam.

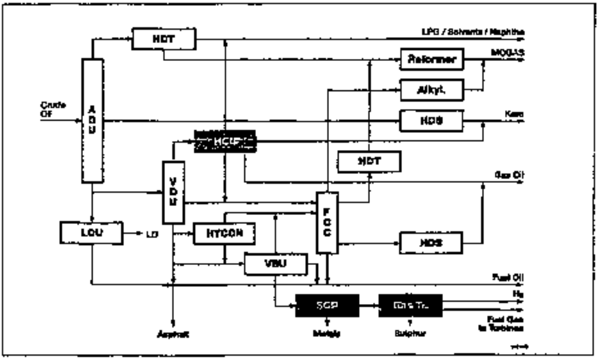


Figure 3 : SGP in the PER* Refinery Environment

Factors governing the selection of new processing facilities were based on a low-cost crude input spectrum, enhanced product quality as well as shifting product demands. A new hydrocracking unit (HCU) will replace the oldest of the two fluid catalytic cracking (FCC) units.

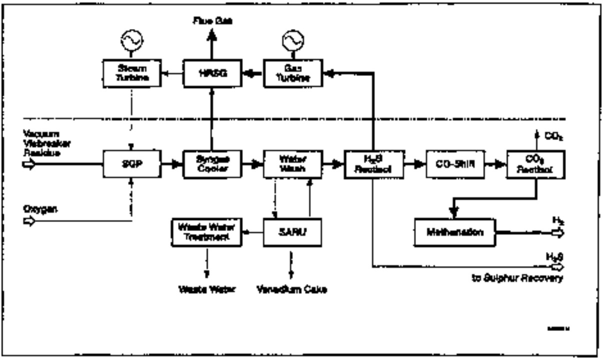


Figure 4 : SGP-based Hydrogen Plant SGHP for PER*

The hydrogen required for the HCU, approximately 250 t/sd, will be produced from syngas originating from gasification of heavy, vacuum-flashed, visbroken residue. Under normal operating conditions the syngas in excess of the requirements for hydrogen production will be used as gas turbine feed. The comoined cycle plant will have an output of 115 MW.

The new plants in the PER+ concept are due for start-up in the second quarter of 1997.

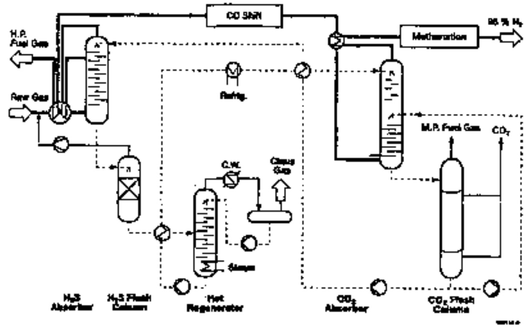


Figure 5 : Selective RECTISOL* Process in SGHP

Figure 5 shows a simplified process flow diagram of the PER+ Fuelgas and Hydrogen plant, i. e. the Selective Rectisol® Process in the so-called SGHP, the Shell Gasification Hydrogen process.

Rectisol* is a physical washing process which uses cold methanol as a solvent. Using cold methanol at -30 °C, both H_2S and $C0_2$ are washed out of the incoming raw gas to a residual total sulphur content of less than 100 ppb, a purity which is sufficient for use with the sensitive catalysts without further processing.

The H.P. fuel gas feed for the combined cycle process is branched off from the H_2S absorber.

The desulphurized gas used for hydrogen production is then shifted outside the Rectisol unit. Next carbon dioxide is removed from the shifted gas in the $C0_2$ absorber. This column is divided into two sections, a bulk $C0_2$ removal section using flash regenerated methanol and a fine $C0_2$ removal section in which hot regenerated methanol is used. The $C0_2$ removal section operates at lower temperatures than the desulphurizing section, typically at about - 60 °C. The permissible $C0_2$ slip is dependent on the application. For hydrogen production based on methanation, typically 100 ppm would be appropriate.

The main reasons for choosing the RECTISOL* process for the PER+ upgrading project are:

- Ability to desulphurize down to less than 100 ppb of total sulphur without additional process steps as required by other physical or chemical acid gas removal processes, e. g. COS/HCN hydrolysis in the raw gas and fixed-bed sulphur polishing of the syngas
- Deep desulphurisation and $C0_2$ removal with a single solvent in a single integrated system
- Inherent flexibility to operate with and without HP-fuel gas offtake.
- High H_2S-selectivity allowing for an enriched acid gas to be treated in a standard Claus plant
- Refinery sulphur emissions will be reduced by one third.

Integration of an oxygen-blown Claus unit into the acid gas removal system

The sulphur recovery unit (SRU) as mentioned already, must be evaluated together with the acid gas removal system (AGR).

The basic part of all sulphur recovery concepts is the Claus unit, which is commonly air-blown. But why not use oxygen in SRU since oxygen is used already for gasification in most IGCC plants. Thus, it is well worth looking at the benefits of operating the Claus unit on oxygen.

Some 10 years ago, when the merits of the novel Lurgi OxyClaus* technology were first presented to Claus plant operators, they reacted with amazement and even disbelief.

In the meantime, oxygen technology has become an established method, both for existing and for new Claus plants.

Using oxygen instead of air considerably reduces the volume of inert gas, primarily nitrogen, to be passed through the system.

This leads to a lower pressure drop in an existing system, thus providing for additional sulphur processing capacity or - in case of new plants - reducing size and capital cost significantly. For example, when adding an IGCC to an existing refinery, revamping the existing air-blown Claus unit with the OxyClaus* Technology, may allow the accommodation of the necessary additional SRU-capacity.

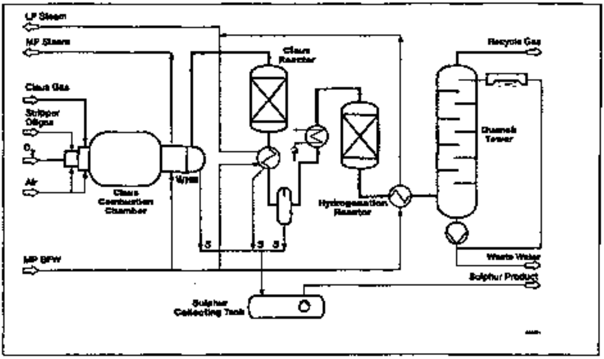


Figure 6 : Claus Unit for IGCC Power Plant

To explain the other benefits of the Lurgi OxyClaus® it is necessary to look at the operating principle of the Lurgi OxyClaus® Burner which is depicted in the next picture.

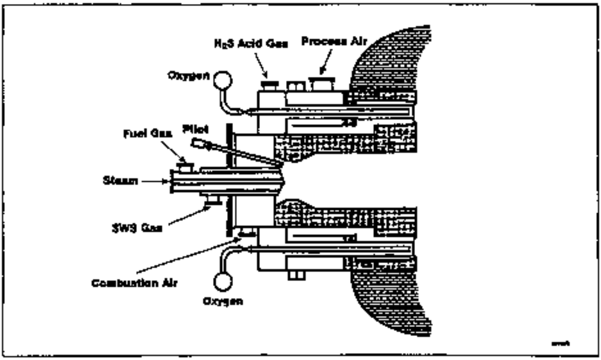


Figure 7 : Lurgi OxyClaus® Burner

The design of this multi-purpose burner provides for operation on air, air plus oxygen or pure oxygen. The control concept allows for flexible processing when handling temporarily high or low amounts of H_2S by automatic change-over from air to oxygen operation and vice versa.

This is a great benefit during load changes of the IGCC plant.

The OxyClaus* Burner has a number of acid gas burners concentrically aranged around a central burner muffle. Each acid gas burner consists of three concentric lances, the oxygen being injected through the inner lance, the acid gas through the central lance and the air through the outer lance.

This results in an extremely hot oxygen flame with a core temperature of about 2000 °C enclosed by a cooler air/acid gas flame, {see photo)

This enables:

- Processing of feed gases with both high and low hydrogen sulfide contents (10-100 %).
- Processing of Claus gases with elevated hydrocarbon content.
- Almost complete combustion of ammonia-containing gases, as encountered in IGCC applications, like sour water stripper gases, off-gas from the carbon slurry flash, etc.

But coming back to the AGR/SRU options in an IGCC (Figure 2), the benefits of an oxygen-blown Claus unit are apparent:

- In case of downstream tailgas treating units, these will be smaller due to the lower process gas load.
- In case of tailgas hydrogenation and recycling, the reduced load will lead to lower operating and capital cost for the hydrogenation, recycle compression and the acid gas removal unit.
- Processing low H_2S acid gas in the Claus unit and, hence reducing the selectivity requirement for the acid gas removal unit, may result in lower capital and operating costs, too.

Conclusion

Selection of an acid gas removal process within an IGCC complex is always a systematic approach taking into consideration numerous aspects:

- Feedstock and type of Partial Oxidation Process (POX)
- Standalone or integrated IGCC
- With synthesis gas offtake or power production only
- Environmental regulations
- Sulphur recovery unit to be evaluated together with the AGR process
- Oxygen in SRU could save money and eases operation.

References

1. C. Higman, Lurgi.Frankfurt,"Gasification of Heavy Refinery Residues", IMP Gasification Seminar, Mexico City, February 24,1994
2. LO.M. Koenders, S. A. Posthuma, P.L. Zuidefeld, SIPM, The Hague.'The Shell Gasification Process for Conversion of Heavy Residues to Hydrogen and Power", EPR11996, Gasification Technologies Conference, San Francisco, October 2 - 4,1996
3. W. Liebner, Lurgi, Frankfurt and N. Hauser, Shell Synthetic Fuels, Houston, "Optimizing/Costing Study for a 500 MW/IGCC Power Plant based on the Shell Gasification Process", EPR11996 Gasification Technologies Conference, San Francisco, October 2 - 4,1996
4. J. Wagner, Lurgi, Frankfurt, "Lurgi OxyClaus* Technology", Western Research Sulphur Recovery Seminar, Istanbul, November 1996.
5. U. Zwiefelhofer, H.-D. Holtmann, Lurgi, Frankfurt, "Gasexpanders in the process gas purification" (in german), VDI-GET conference 'Energy recovery by gasexpansion in industrial use", Magdeburg, March 14-15, 1996
6. M.M. Weiss, H. Schlichting, Lurgi, Frankfurt, "Comparison of gas purification processes for selective removal of H_2S from raw gases in an IGCC" (in german, not published), Annual meeting of the VDI-GVC expert committee 'Thermal separation of gas- and liquid mixtures', Luzern, April 29 - 30,1996

OPTIONS FOR THE USE OF REGENERABLE SULPHUR SORBENTS IN COMBINED CYCLE PROCESSES

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ABSTRACT

The Coal Technology Development Division (CTDD) of British Coal is developing components of a coal gasification based combined cycle power generation system known as the Air Blown Gasification Cycle (ABGC). The ABGC is being developed jointly by an industry led group comprising GEC Alsthom, PowerGen, Mitsui Babcock Energy and British Coal Corporation towards a demonstration of the technology. The cycle offers a significant increase in power generation efficiency compared to conventional coal fired systems with the added advantage that this can be achieved at reduced capital cost and low emissions.

A number of work programmes are being undertaken at CTDD to develop the component parts of the cycle ie high pressure gasifier, char combustor, gas cleaning and gas combustion systems. The information from the various work programmes is being used to design a Prototype Integrated Plant (PIP).

Sulphur control to meet current European legislation is inherent in the process, however tighter emissions limits are anticipated for new power plant in the near future and this will require further sulphur control. Sulphur removal at high temperatures using regenerable sorbents with fluidised bed technology is the preferred route for the ABGC. This option offers a lower cost alternative to conventional techniques which are based on wet gas cleaning systems that have high capital costs, an adverse effect on cycle efficiency and produce a liquid effluent that is expensive to treat.

This paper describes the ABGC with in-bed sulphur retention and methods under development including recent experimental work performed at CTDD using regenerable sorbents for further sulphur emissions reduction. Various options for the method and technology of sorbent regeneration and subsequent processing of the sulphur are described. A status report on other gas cleaning systems and on various hot gas cleaning demonstration plants world wide is also presented.

1. INTRODUCTION

The ABGC is being developed by the UK led Clean Coal Power Generation Group (CCPGG) which comprises GEC Alsthom, PowerGen, Mitsui Babcock Energy and British Coal Corporation. Significant financial support for the ABGC has been provided by the UK Department of Trade and Industry and European sources. The cycle is based on the partial gasification of coal in an air blown spouted fluidised bed at elevated pressures of typically 25 bar and temperatures up to 1050*C. An outline diagram of the cycle is shown in Figure 1. A sorbent such as limestone or dolomite is also injected into the gasifier to retain most of the sulphur which otherwise would be released into the fuel gas. Between 70 % and 80 % of the coal substance in converted to a low calorific value fuel gas. The fuel gas is then cooled to between 400"C and 600*C and filtered in a ceramic candle filter to remove almost all the remaining particulates before being burnt in a gas turbine. The solid residue from"the gasifier, containing a mixture of mineral matter, char and sulphided sorbent residue, leaves in two main streams; as a coarse solid from the base of the gasifier and as fine material elutriated from the gasifier bed and captured in the hot cyclone and the high efficiency ceramic barrier filter. Both solid residue streams are depressurised, cooled and burnt in a Circulating Fluidised Bed Combustor (CFBC) boiler, thereby raising steam which is used to generate further power in the steam turbine.

The ABGC is a highly efficient Clean Coal Technology (CCT) with a projected commercial plant efficiency of 47 % (LHV basis) using sub-critical steam conditions and currently available technologies. By comparison, pulverised coal plants offer less than 40 % efficiency and IGCC systems typically 41-43 %. Advances in steam conditions and the availability of improved materials will increase the efficiencies of all competing CCTs and efficiencies of > 50 % for the ABGC should be achievable in the next few years.

The base cycle of the ABGC meets all current European emissions legislation. However, new more stringent regulations are imminent and components to reduce the emission of sulphur oxides, nitrogen oxides and hydrogen chloride are being developed. The contaminant reduction technologies being developed are based on hot gas cleaning techniques.

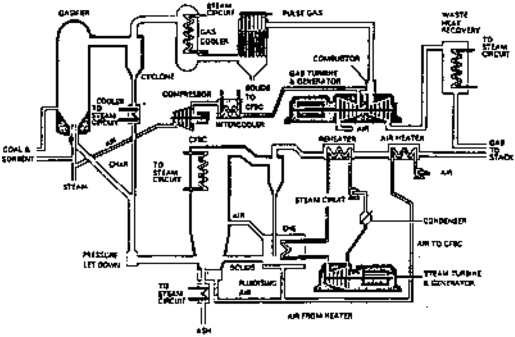

FIGURE 1 Schematic of the Air Blown Gasification Cycle

2. SULPHUR CONTROL TECHNIQUES

Commercially available wet gas cleaning techniques comprise a number of processes including carbonyl sulphide hydrolysis, amine scrubbing, hydrogen sulphide stripping, followed by Claus and Scott type processes with fuel gas humidification and reheat. These items may typically be as large as the rest of the IGCC plant and represent a significant part of the capital and operating cost.

Such techniques were developed initially for use in oil refineries for cleaning syngas for use in the petrochemical industry and for removing hydrogen sulphide from fuel gas produced from high sulphur heavy oil and coke residues for use in gas turbines. These techniques are currently being applied to coal based IGCC systems at sites such as Wabash River and Polk County USA, Puertollano in Spain and Buggenum in the Netherlands.

In recent years hot gas cleaning techniques for the removal of hydrogen sulphide have been developed. The basis of the sulphur removal technology is that the fuel gas is contacted with a sorbent, usually a metal oxide, to form a metal sulphide. The sulphided material is then transferred to a separate process where it is regenerated. The re-oxidised material is returned to the sorber and the now concentrated sulphur containing regeneration off-gas is treated separately. Typically the sulphur may be in the form of H_2S, $S0_2$ or elemental sulphur.

Detailed cost and efficiency studies of gas cleaning options for the ABGC have been carried out 131. Table 1 below shows the costs and efficiency gains of hot gas cleaning for the ABGC compared to the use of conventional low temperature wet gas cleaning techniques. Costs for post turbine gas cleaning based on FGD and SCR are also shown. All achieve similar emission levels.

		Hot Gas Clean-**up**	Flue Gas Cleaning
Efficiency	% LHV	+ 1.7	+ 0.8
Insta'lation Cost	K ECU/MW	-67	+ 39
Electricity Cost	ECU/MWh	- 1.1	+ 0.9

Table 1 Efficiency and Costs of Hot Gas Cleaning cf. Wet Cleaning

As part of the development of a hot gas cleaning process (HGCK) KEMA undertook techno-economic studies of a system for application to a 550 MWe Shell-based IGCC system [4]. Their calculations show a net efficiency increase of 2 % and that the installation cost of their HGCK process is less that half of a conventional wet scrubbing process. The operating costs are reduced by about 30 % leading to a reduction in the cost of electricity of around 3 ECU per MWh.

Rheinbraun carried out a similar study for their various hot gas cleaning concepts [5]. For a 1100 MWe HTW-based IGCC power plant up to 2.3 % net efficiency was gained when using hot gas desulphurisation at 500*C. A specific investment reduction of 130 DM per kW of power production was calculated compared to a conventional low temperature sulphur recovery system with a reduction in the cost of electricity of Pfg 0.3/kW. (approx 1.6 ECU per MWh)

The USDOE also recognises the longer term advantages of hot gas cleaning techniques through its considerable support within the US Clean Coal programme. Clearly, there are significant economic and efficiency benefits to be gained by all CTT's by the application of hot gas cleaning, although the relative benefits are cycle specific.

3. HOT GAS DESULPHURISATION

In developing and applying a regenerable metal oxide sulphur removal system at high temperature there are a number of technical issues to consider. These include the key issues of sorbent choice, sorbent durability, contacting method, regeneration gas and sulphur recovery system. Each of these points is discussed below. A schematic diagram summarising these issues is shown as Figure 2 (courtesy of Rheinbraun)[5].

Sorbent and Contacting Method

The choice of primary metal oxide for the sorbent depends on the temperature of interest and the degree of sulphur removal required. To date formulations based on calcium, zinc, iron, copper, molybdenum, nickel, manganese "and tin have been reported in the literature for use on coal gases in the temperature range 250*C to 900 "C. Additives are often incorporated to stabilise the primary metal oxide and to suppress side-reactions. The current developmental status of many of these sorbents is discussed elsewhere [6J. To date only zinc and iron based materials have been tested in real coal gas systems.

In real coal gas systems not only must the sulphidation reaction be monitored but also side reactions. Methanation and water gas shift reactions can be catalysed by additives to the sorbents resulting in excess bed temperatures and sorbent sintering. Carbon deposition may plug the pores and reaction with hydrogen chloride could produce volatile metal chlorides.

Moving bed, fluidised bed 3nd transport reactor systems are being investigated and suitable sorbent sizes and formulations manufactured for each. Each process has advantages and disadvantages over the others and as yet there is no clear indication as to which is likely to have the greatest long term success.

For economic viability of the process the sorbents must retain their activity as they pass round and round the system from sulphider to regenerator. The sorbents must also retain mechanical stability with minimal attrition losses. Initial calculations on a polishing system for the ABGC based on a twin fluidised bed reactor suggest that if the sorbent performs for 1000 cycles the additional cost of the process on electricity production will be minimal. The calculations assumed that the sorbent picked up around 4 % wt sulphur from the sulphider in each cycle.

The performance of sorbents is often compared on the basis of reduction in capacity over a number of cycles. However, caution should be exercised. For example in a moving bed, a cycle may last several hours with the sorbent picking up 10 % wt S in each cycle, but in a transport reactor the sorbent may collect only a fraction of a percent wt sulphur and a cycle will last only 5-10 seconds.

Regeneration Gas, Sulphur Recovery and Disposal

The sorption process can not be treated in isolation as the regeneration is the most critical step of the whole process. The temperature has to be maintained within limits to ensure good regeneration, to prevent sulphate formation and sintering all of which limit the sorption ability of the sorbent.

The regeneration gas may be air, air\steam (or other inert), flue gas or steam depending on the process chosen and sorbents used. The gasification process must be able to produce the regeneration gas eg using air in nitrogen would not be an option for an air blown system where there is no air separation plant. However, high pressure air is available from the gas turbine compressor and steam from the steam cycle. If air alone is used then heat extraction may be needed from the regeneration bed to moderate the temperature.

The "light off" temperature may also be significant. The sorbents require a minimum temperature to sustain regeneration. Typically, this is around 550*C. Developments are underway to lower the "light off" temperature. This has two advantages, firstly little additional heating of the regeneration gas is required saving on efficiency and secondly higher air concentrations can be used as a higher temperature rise is required in the regeneration process.

To date most of the sorption processes produce concentrated $S0_2$ streams in regeneration. Simple low cost systems are required to collect the sulphur. Options include direct sulphur recovery by reduction with fuel gas over a catalyst to produce elemental sulphur, production of sulphuric acid, flue gas desulphurisation or in the case of the ABGC collection in the CFBC by addition of extra calcium based sorbent. The optimal process in terms of cost and efficiency is cycle specific and may be affected by local issues such as the market for sulphur or sulphuric acid.

Figure 2 Issues For Regenerable Sulphur Sorbent Processes.

Other Approaches

As well as the use of regenerable metal oxides other hot gas cleaning techniques for the removal of hydrogen sulphide are being studied. The use of high temperature ceramic membranes is being considered to selectively remove hydrogen sulphide. At Rheinbraun [5] a process whereby air is added to the fuel gas immediately prior to the filter is being developed. The air oxidises the hydrogen sulphide to sulphur in **a** catalytic process involving the lignite char. The sulphur is then taken out in the filter on the char.

CTDD Experience

At CTDD both fixed and fluidised bed formulations of zinc based sorbents have been tested on pilot scale rigs. A diagram of the test rig is shown in Figure 3. Fluidised bed sorbents from RTI, zinc titanate (ZT-4) and Phillips Petroleum, Z-Sorb*lll, were tested at 10 bar pressure and over a temperature range 450 - 600 *C with fuel gas from a fluidised bed gasifier. H_2S and COS concentrations were reduced to less than 10 vpm, good retention performance was observed and acceptable attrition resistance was measured.

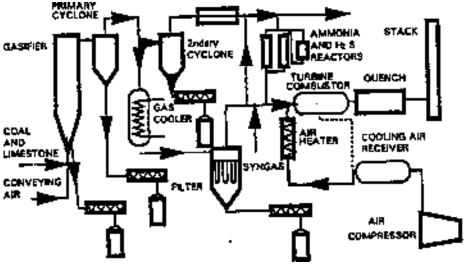

Figure 3 CTDD Gas Cleaning Test Facility

4. OTHER HOT GAS CLEANING COMPONENTS

As well as desulphurisation other hot gas contaminants control technologies are being developed. These are reviewed briefly below and in more detail elsewhere [7].

Hydrogen Chloride

In almost all coal gasification processes the chlorine present in the coal is released into the vapour phase as hydrogen chloride gas. Coal chlorine contents vary widely (100 - 200 ppm wt) although UK coals tend to have a higher chlorine content (up to 8700 ppm wt) than typical US, Australian and European coals. There is little legislation governing the emissions of chlorine from fossil fuel plants. However, tighter legislation is expected and already low emissions levels have been set for plant fired on waste fuels. In addition to potential environmental problems hydrogen chloride removal could be required to protect sulphur and ammonia sorbents, to reduce corrosion of hot gas path components and to suppress the volatility of toxic and corrosive trace elements. For hot gas cleaning applications in the range 300 - 600" C calcium and sodium based sorbents have been identified as thermodynamically favourable.

As for other contaminants, the sorbents could be contacted with the fuel gas using fixed, fluidised, or transport reactors. At CTDD an approach of injecting the sorbent prior to the filter has been adopted [81. Calcium and sodium based materials have been entrained in the fuel gases on an atmospheric gasifier prior to a hot gas filter. Hydrogen chloride reduction of up to 80 % was measured. Krishnan [9] used a laboratory scale fixed bed reactor to test various sodium based sorbents including Nahcolite, Dawsonite, Shortite and Katalco. The sorbents were able to reduce HCl levels from 300 vpm down to less than 1 vpm. This technology remains to be tested at a significant scale on real coal gasification plant. Of particular concern is the possible formation of sticky calcium or sodium chlorides. If formed these could blind the filter or deposit in the duct work causing a blockage.

Ammonia Reduction

NOx emitted from gasification based combined cycles is produced in the gas turbine by three mechanisms. Firstly from oxidation of molecular nitrogen (thermal-NOx) in the oxygen rich zone of the flame. This is normally only a problem where high CV gases and premium fuels are used. For low CV fuels with peak burning temperatures of < 1600*C thermal NOx is not an issue. Secondly, is prompt-NOx formed from atmospheric nitrogen through intermediates as results of the hydrocarbons in the fuel. Finally fuel-NOx formed from bound nitrogen components such as ammonia and hydrogen cyanide.

NOx formation in the gas turbine can be minimised by combustor design. Where thermal NOx is the dominant mechanism low NOx burners are used to reduce the flame temperature. For systems where fuel-NOx is the primary mechanism of NOx formation low NOx burners are used to create a rich-lean air staging in order to promote conversion of fuel nitrogen to molecular nitrogen [101. However, use of these systems is unlikely to be sufficient to meet stringent NOx limits. In order to reduce emissions levels further, removal of fuel nitrogen compounds prior to combustion is required. Where wet gas cleaning techniques are used the ammonia is largely removed by dissolution in the aqueous scrubber. For hot gas cleaning two potentially much simpler processes have been identified.

Decomposition Catalysis

Thermodynamic equilibrium calculations show that for coal gasification processes the ammonia levels in the fuel gas are often above the equilibrium value. In principle a catalyst should be able to decompose the ammonia by the following reaction:-

$$2\ NH_3 \sim N_2 + 3\ H_2$$

A number of catalysts formulated from nickel and iron were tested by at CTDD [11] in a laboratory scale reactor. Good activity was observed over the temperature range 600-800"C in clean gas. However, when the nickel based catalysts were tested in real gas at around 600 *C no activity was measured [12]. The activity of these materials appears to be reduced at these temperatures due to sulphur and chlorine poisoning and carbon deposition. Similar observations were made by Mojtahedi who tested nickel and ruthenium catalysts on real fuel gases [13]. At higher temperatures say > 900*C these problems could be reduced as nickel sulphide formation and carbon deposition are less favourable.

Selective Oxidation

An alternative approach is to oxidise the ammonia to harmless products using a variety of oxidants by a reaction such as:-

$$2NH_3 + 1.5\ 0,,-N_2 + 3H_20$$

The oxidants could be air or NO derived from the flue gases or compressor air. To be of any benefit the reaction has to be highly selective. At KEMA a molybdenum oxide on silica sorbent was tested and gave 90% reduction when 2 % oxygen was added at 450 *C [14]. In Finland ammonia gasification of biomass leads to ammonia levels in excess of 4000 vpm [15]. At VTT nitrogen oxide and air were added to this fuel gas over an aluminum oxide bed in laboratory scale tests. At temperatures below 450 *C over 80 % reduction of ammonia was reported.

The various components of the hot gas cleaning process continue to be developed at CTDD. There a number of issues relating to each technology and how they may be integrated yet to the resolved. Within the UK there are plans to build an 87 MWe Prototype Integrated Plant (PIP) to demonstrate the ABGC at commercial scale. As part of this, a sidestream will be built and operated to develop further the hot gas cleaning technologies. The key components of the sidestream are shown below in Figure 4.

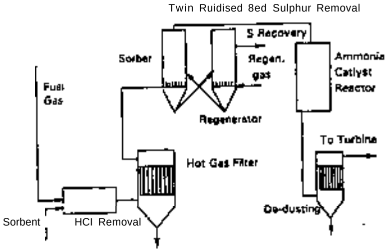

Figure 4 Hot Gas Cleaning Processes

INDUSTRIAL SCALE SULPHUR REMOVAL

Worldwide there are many power companies, research organisations and academic institutions developing dry regenerate sorbent sulphur removal techniques for application above 300 *C in order to reduce the costs of IGCC plant and to improve cycle efficiency.

Enviropower (now Carbona Inc.) has tested several zinc titanate based sorbents on a slipstream of a pilot scale air blown gasifier 116]. The sulphur reactor consisted of two fluidised bed reactors, one acting as the sorber the second as the regenerator. The system was tested continuously for periods of 5-6 days at pressures of up to 20 bar. High sulphidation efficiencies were observed as well as continuous $S0_2$ production. Carbona Inc. are continuing to develop and commercialise the process.

In Japan various dry gas cleaning systems have been tested on slipstreams of the Mitsubishi air blown, two-stage entrained bed coal gasification plant [17]. The original dry cleaning concept was a twin fluidised bed reactor system for sulphur removal followed by a granular bed filter to remove the entrained dust. More recently a ceramic candle filter followed by a fixed bed honeycomb system based on a supported iron oxide sorbent has been tested. The system was shown to operate reliably reducing sulphur levels from around 600 vpm to less than 50 vpm H_2S + COS. Good durability of the sorbent was reported. A clean up system for a conceptual 2000 t/day gasifier is currently being designed.

KEMA have completed bench scale development of a component of a conceptual hot gas clean up process (HGCK) [4]. The system could be used on any gasification plant and comprises of a cooler to reduce the fuel gas to around 380 *C followed by fly ash removal (eg candle filter or rotating particle separator). The gas is then cooled to 230 "C and water scrubbed to remove the HCI, alkalies and trace metals before being reheated and passed to the desulphurisation unit. The continuous sulphur removal process consists of two bubbling fluidised beds, one acting as sorbent, the other a regenerator. The sorbent developed and patented by KEMA is transported continuously between the two beds. Sorption is carried out at between 350 *C and 450 *C reducing H_2S to below 20 vpm. Regeneration is carried out at higher temperatures producing elemental sulphur directly that can be condensed and collected. HCN and NH_3 are removed at the same time as the sulphur gases. KEMA has recently signed and agreement with Foster Wheeler to develop the process further and to demonstrate the technology on a 1 MWth slipstream of an IGCC system.

At Wilsonville, US, Southern Company is developing a second generation PFBC system 118]. The plant will eventually encompass a coal gasifier. A granular bed filter combining simultaneous trace metal, NH_3, HCl and H_2S capture is being developed to provide the fuel gas cleaning.

Two hot gas desulphurisation concepts are being demonstrated as part of the USDOE clean coal programme at Tampa Electric's Polk County and at Sierra Pacific's Pinon Pine plant. Both plants will test zinc based sorbents supplied by Phillips Petroleum.

The unit at Polk County is a moving bed system developed by GE and will operate as a slipstream on the Texaco gasification process. The facility has been built and is currently being prepared for operation. At Pinon Pine a full stream transport reactor is to be tested on the KRW gasifier. The reactor is in an advanced stage of construction and will operate later this summer.

6. CONCLUSIONS

Hot gas cleaning technologies offer significant cost and efficiency savings compared to conventional low temperature wet gas cleaning techniques. Calculations by various organisations suggest efficiency improvements of around 2 % net, lower investment costs and a reduction in the cost of electricity of between 1 - 3 ECU per MWh.

Hot gas desulphurisation systems based on the use of regenerate metal oxides have been shown to match the sulphur retention performance of amine based liquid scrubbers. Long term testing of various processes is currently underway to prove the reliability of the technologies.

REFERENCES

1. Sage P W , Development of the Air Blown Gasifier, Paper to ETSU Coal Contractors Meeting, Nottingham, UK, November 1994.
2. Gayan P, Sage P W, Schofield P A, Minimisation of Calcium Sulphide in Gasification Residues By Combustion n a Circulating Fluidised Bed, to be presented at the 14th Int. Conf. on FBC, Vancouver, May 1997.
3. Cahill P, Dutton M, Carr C, Davison J, Integrated Fuel Gas Cleaning for Advanced Combined Cycle Processes, JOULE II Project, Contract No. JOU2/CT93/043, Final Report December 1996.
4. Meijer R, Janssen F, Farina G, Hellendoorn J, KEMA's Hot Gas Clean-up Process, Paper Presented at Gas Cleaning at High Temperatures Conference, Karlsruhe, Sept 1996.
5. Meyer B, Walter H, Schiffer H-P, Adlhoch W, Sulphur Removal from Coal Based Fuel Gas, ibid.
6. Fantom I R, Radford S, Performance of Regenerable Sulphur Sorbents on Coal Derived Fuel Gas, ibid.
7. Fantom I, Sage P, Cahill P, Hot Gas Cleaning - An Overview, Desulphurisation of Hot Coal Gas with Regenerable Metal Oxide Sorbents : New Developments, NATO ASI Conference, Kusadasi, July 1996.
8. Duo W, et al., Kinetics of HCl Reactions with Calcium and Sodium Based Sorbents for IGCC Fuel Gas Cleaning, Chem.Eng.Sci., 51(11), p2541, 1995.
9. Krishan G N et al., Removal of Hydrogen Chloride from High Temperature Coal Gases, Journal of American Chemical Society, 33, 1, 1988.
10. Kelsall G J, Cannon M F, Combustion of Low Heating Value Gas in a Turbine, Power Production from biomass II, Espoo, Finland, March 1995.
11. Gavin D G et al., Minimisation of Nitrogenous Pollutants during Gasification and Combustion of Coal, ECSC Project No 7220-EC\010, Final Report, April 1995.
12. Fantom I, Removal of Ammonia in Coal Gasification Systems, UKDTI Contract No.C\00089, August 1996.
13. Mojtahedi W et al., Catalytic Decomposition of Ammonia in Fuel Gas Produced in Pilot Scale Pressurised Fluidised Bed Gasifier, Fuel Processing Technology 45, 1995.
14. Bolt N, Meijer R, et al, Haalbaarheidsbeoordeling Hoge Temperatur Gasreiniging KV-STEG, KEMA, 1993.
15. Leppalahti J, Koljonen T, Hupa, H, Kilpinen P, Application of Selective Catalytic Oxidation (SCO) to NH_3 Removal in Gasification, Paper to be presented at Gas Cleaning at High Temperatures Conference, Karlsruhe, Sept 1996.
16. Konttinen J, Mojtahedi W, Abbassian J, Coal Desulphurisation at the 15 MWth Pressurised Gasification Pilot Plant, High Temperature Gas Cleaning, Proceedings of Conference, Karlsruhe, Sept 1996.
17. Kaneko S, et al. Japan Tests Air Blown IGCC, Modern Power Systems, p19, April 1996.
18. Loganbach J, Moorehead E, Styles G, Vimalchand P, Transport Reactor Development Gives a Boost to US Coal Fired Plant, Modern Power Systems, p 37 November 1996.

CHEMICAL SOLVENT-BASED PROCESSES FOR ACID GAS REMOVAL IN GASIFICATION APPLICATIONS

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Introduction

Chemical solvent-based processes are well suited to the removal of acid gases from gasifier product streams. A combination of solvent choice and equipment design can be used to meet specific product and/or emission requirements. Examples will be reviewed of how specialty solvents can be used to meet different objectives, such as maximum selectivity for hydrogen sulfide (H_2S) over carbon dioxide (CO_2), enhanced carbonyl sulfide removal, or efficient total removal of CO_2.

Gasifier streams also present unique challenges to the use of chemical solvent-based processes. Depending upon the gasifier feed stock, the solvent may become contaminated with a variety of species that impose an added burden on the operability of the treating process. Contaminants can be carried in with the gas and/or formed in-situ. A practical strategy for dealing with hydrogen cyanide, carboxylic acids, metal carbonyls and particulates will be discussed.

Acid Gas Cleanup Technologies In Gasification Applications

All gasification processes include an acid gas cleanup step, whatever the raw feed stock used and whatever the ultimate use of the synthesis gas produced. Although several trials of hot (dry) gas cleanup have been conducted, all commercial acid gas cleanup today is carried out via cold (wet) systems. These fall into two broad classes: physical solvents and chemical solvents (and occasionally hybrids of the two). Both are proven technologies with many years of operating experience. Within these two broad categories there are many different products available [1,2].

Physical solvents, as the name implies, rely upon variations in the physical solubility of gases to effect separation. High solubilities of the contaminants are required for physical solvents to perform efficiently, and high partial pressures of those species provide the driving force for absorption. Union Carbide offers SELEXOL®, a proven physical solvent, via a licensed process with UOP. This was first used in gasification applications in the 1980's at Texaco/Coolwater and TVA/Muscle Shoals. More recently it has been selected for the Sarlux and api Energia projects. It's use in gasification applications has been described elsewhere [3] and will not be covered further here.

Chemical solvents are virtually all amine-based and remove H_2S and CO_2 via an acid-base reaction. Building on years of experience in natural gas, refinery, and synthesis gas plants, specially formulated chemical solvents have been developed to meet the various requirements of gasification plants. While experience in other applications has been invaluable in developing products for the gasification market, there have been several new challenges to overcome. This paper addresses the potential problems that can be encountered when using chemical solvents and offers practical solutions.

The decision over which acid gas cleanup technology to use is influenced by many factors, including but not limited to:

- integration of synthesis gas cleanup with existing processes
- acid gas partial pressure
- selective versus total acid gas removal
- capital cost
- operating cost
- nature of feed gas contaminants

Each factor will be described briefly. In-depth discussion is presented concerning the nature of contaminants and their impact on process selection and unit operation.

Process Integration

The popularity of gasification as an economical and efficient disposal method for refinery bottoms presents interesting possibilities for the integration of acid gas cleanup systems with existing sulfur removal and recovery equipment. All refineries have amine-based systems for handling H_2S. Utilizing any excess capacity in existing units offers the ability to lower capital costs for a new gasifier. As will be discussed, cross contamination of amine solutions is a real concern for such a scenario. However, with proper anticipation of potential contaminants, both from the refinery and from the gasifier, plans can be made for the pre-treatment of gas streams and/or the reclamation of contaminated solution.

There are also ways to create additional capacity in existing desulfurization equipment without capital expenditure. Specially formulated treating solvents are available that operate at higher concentrations than generic monoethanolamine (MEA) or diethanolamine (DEA). These same solutions may be appropriate for use in an integrated gasifier.

Acid Gas Partial Pressure

With physical solvents, acid gas partial pressure provides the driving force for absorption. The higher the pressure, the lower the required circulation rate to affect separation. This improves operating economics for physical solvents.

With chemical solvents, partial pressure is the driving force for mass transfer. The higher the pressure, the lesser number of stages are required to affect separation. This reduces the capital requirement for chemical solvents.

The partial pressure of $C0_2$ and H_2S also affects a solvents ability to selectively remove H_2S while slipping $C0_2$ into the treated gas stream. IGCC applications require selective removal of H_2S and maximum $C0_2$ slip. A solvent with superior slip characteristics offers considerable advantage because more gas will be available to produce power in the turbine.

Selective Acid Gas Removal: IGCC

When the raw material fed to the gasifier contains sulfur, the principle sulfur species in the raw synthesis gas are H_2S and COS. For subsequent combustion in a gas turbine the level of sulfur species must typically be less than 50 ppm. From the perspective of overall energy efficiency, the slip of $C0_2$ through the acid gas removal unit should be as high as possible. Methydiethanolamine (MDEA) is often cited as the solvent of choice in this application, giving good sulfur removal and reasonable $C0_2$ slip. Several authors have addressed the mechanism by which MDEA selectively absorbs H_2S [4,5]

Figure 1 shows a simplified process flow diagram for chemical-based acid gas treating. Cooled synthesis gas enters the bottom of an absorber where it contacts an aqueous chemical solvent solution. The treated gas exits the absorber and continues to the next processing step, which is the gas turbine in IGCC applications. Cool lean solution enters the top of the "absorber and counter-currently contacts the synthesis gas using trays or packing, absorbing acid gas contaminants as it passes down the column. Warm rich solution leaves the bottom of the absorber and is routed to a regenerator. Steam stripping is used to remove acid gas from the solution. This results in a concentrated acid gas stream which can be fed to a Claus sulfur recovery unit. The hot lean solution is then cooled prior to returning to the absorber. A lean/rich cross exchanger is used to reduce the sensible heat load on the regenerator reboiler.

Over the past 15 years, solvents have been developed that allow greater CCH slip compared to MDEA. Selectivity is a function of the ratio of H_2S to CCh, the number of trays in the absorber, and the solvent used for absorption. $C0_2$ and sulfur content can vary widely depending on the feed to the gasifier. Slip values can range from 70-88% of the inlet $C0_2$. A sample synthesis gas stream with moderate CCs content is shown in Example 1.

The economic advantage of greater gas volume going to the power turbine is the most significant reason for using specialty solvents. Note that the specialty amines offer enhanced $C0_2$ slip at the cost of reduced ability to meet tight sulfur specifications. Reduced energy consumption is realized in the treating system when COj slip is increased. This comes from lower solvent circulation (less sensible heat), lower heat of reaction, and the fact that CO? that is not absorbed does not have to be regenerated.

Example 1: Enhanced CO_2 Slip*

Feed Gas: 175,000 Nm³/hr (157 MMscfd)
2690 kPa (390 psia); 40 °C (104 °F)
CO_2: 10.6 mole%
H_2S: 0.6 mole%

Solvent	MDEA	UCARSOL® HS-101	UCARSOL® HS-115	UCARSOL® HS-111
Circ. Rate, m³/hr (gpm)	192 (845)	184 (810)	166 (731)	149 (656)
CO_2 Slip, % of inlet	78	80	84	88
H_2S in Outlet, ppmv	4	4	10	35
Reboiler Duty, Gcal/hr (MMBtu/hr)	9.9 (39.3)	9.5 (37.7)	8.6 (34.2)	7.7 (30.6)

* Based on 50 wt% solutions, 10-tray absorber, 102 kg stripping steam/m³ solution (0.85 lbs/gal), 17°C (30°F) L/R exchanger approach.

Total Acid Gas Removal: Chemical Production

In applications where the gasifier product is destined to be used as a chemical feed stock, complete removal of carbon dioxide and sulfur species is required. CO_2 specifications of less than 100 ppm are typical. CO_2 removal can be accomplished at the same time the sulfur is removed (total acid gas removal) or a portion of the synthesis gas can be treated for further CO_2 removal after the H_2S has been selectively removed. This scheme has the advantage of concentrating the H_2S from the first treater for feed to a sulfur recovery unit. In some chemical feed stock applications it proves economical to generate the synthesis gas from sulfur-free raw materials, such as natural gas, so that only CO_2 removal is required.

In the past, MEA and inhibited MEA have been used for this application. Specialty solvents are now available which offer significantly better operating efficiency versus MEA. An example using a synthesis gas stream with moderate CO_2 content is shown in Example 2.

Significant efficiency is gained from the use of formulated MDEA-based solvents by operating at 50 wt% in lieu of lower concentrations for generic amines. Industry experience indicates that primary amines cannot be operated at higher concentrations without increasing the potential for corrosion. Operating at higher strength reduces the solvent circulation which reduces the sensible heat load on the regenerator. Further energy efficiency is gained because the heat of reaction between CO_2 and tertiary amines is lower than the heat of reaction with primary amines.

Example 2: Efficient CO_2 Removal*

Feed Gas: 153,000 Nm³/hr (137 MMscfd)
2620 kPa (380 psia); 40 °C (104 °F)
CO_2: 9.3 mole%

Solvent	18 wt% MEA	50 wt% UCARSOL® CR-422
Circ. Rate, m³/hr (gpm)	667 (2940)	381 (1680)
CO_2 Specification, ppmv	100	100
Reflux Ratio, mole/mole	2.5	1.25
Reboiler Duty, Gcal/hr (MMBtu/hr)	42.1 (163.3)	23.8 (94.6)

* Based on 30-tray absorber, 17°C (30°F) L/R exchanger approach.

Capital and Operating Costs

In a typical gasification project the cost of the acid gas removal technology represents only a small proportion of the overall project cost. However, the choice of cleanup technology and design of the acid gas removal unit has long term consequences for plant reliability and cost of operation. These can have a significant impact on the ultimate viability of the project. This paper will not address capital and operating costs directly but, instead, will focus on operational difficulties that can occur when chemical solvents are chosen for the acid gas cleanup. Practical solutions for these potential problems are presented.

Feed Contaminants

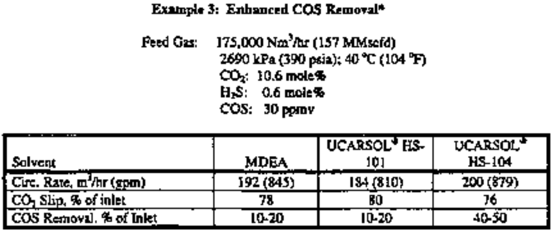
Gasifier synthesis gas contaminants, other than H_2S and $C0_{2>}$ fall into four main categories: metal carbonyls, COS, foam promoters and foulants, and carboxylic acids and their precursors that form heat-stable amine salts.

Iron and nickel carbonyls present an interesting problem. They are only partially soluble in aqueous solutions so consideration has to be given to the potential impact on downstream turbine blades. If the anticipated level of contamination in the treated synthesis gas is not acceptable, it may be advisable to use a physical solvent to achieve total removal of the metal carbonyls. If the level of carbonyls removal by chemical solvents is adequate, they can be removed from the working solution via particulate filters, although provision for handling of potentially hazardous filter cake has to be made. Experience indicates that moist filter cake presents no airborne hazard and protective clothing is adequate to protect workers from dermal contact.

COS Removal

Local environmental regulations typically control the level to which sulfur must b- removed. In cases where very strict effluent levels are required, COS hydrolysis may be recommended upstream of the acid gas removal unit. This step converts all but a few ppm of the COS to hydrogen sulfide. Solvent choice also plays a part in the decision for/against COS hydrolysis as different solvents are able to remove COS to different levels under given conditions. Example 3 demonstrates how one specially formulated MDEA-based solvent is able to enhance COS removal while maintaining most of it's $C0_2$ slip.

The COS removal performance of specially formulated solvents may be sufficient to avoid the installation of a COS hydrolysis reactor. A penalty is paid, however, in reduced $C0_2$ slip. Upstream COS hydrolysis is probably preferred in facilities which must meet stringent total sulfur emission levels.



Example 3: Enhanced COS Removal*

Feed Gas: 175,000 Nm^3/hr (157 MMscfd)
2690 kPa (390 psia); 40 °C (104 °F)
CO_2: 10.6 mole%
H_2S: 0.6 mole%
COS: 30 ppmv

Solvent	MDEA	UCARSOL® HS-101	UCARSOL® HS-104
Circ. Rate, m^3/hr (gpm)	192 (845)	184 (810)	200 (879)
CO_2 Slip, % of inlet	78	80	76
COS Removal, % of Inlet	10-20	10-20	40-50

* Based on 50 wt% solutions, 10-tray absorber

A significant number of operational problems associated with wet solvent systems can be traced to solvent contamination by soot/particulates, iron sulfide, tars, or surface active species such as hydrocarbons. For the most part these are introduced unintentionally with the synthesis gas. Adequate and reliable pre-treatment of the synthesis gas is perhaps the best way of minimizing contamination of the acid gas cleanup solvent. In a typical gasification process the hot synthesis gas exits the gasifier and is passed through a series of waste heat boilers, quenches, and water washes, to recover sensible heat as steam, and remove the soot, tars, and higher boiling hydrocarbons that are unavoidably formed during the gasification process. A wide variety of water washes are employed [6]. These washes are not always as efficient as expected, particularly with very fine aerosols or particulates. Often times they simply malfunction or are under-designed for startup or upset conditions. Union Carbide has developed recommendations for dealing with each contaminant based on experience in hundreds of chemical solvent-based treating units.

Clean uncontaminated treating solutions have a very low tendency to foam. This is confirmed by reports in the literature [7] as well as Union Carbide's field experience. It has also been confirmed that the addition of sparingly soluble contaminants, such as high molecular weight hydrocarbons, tars or lubrication oils, increases the foaming tendency. Operationally, foaming can lead to increased solvent losses and off-specification treating.

_ An activated carbon filter is recommended in the chemical treating system to purify a 10-20% slip-stream of cool lean solution. This is usually adequate to take care of chronic contamination problems. Anti-foam agents are administered as needed to suppress foaming during acute contamination episodes. Administering anti-foam to a system on a routine basis should not be required and will shorten the life of the activated carbon bed, but injection systems should be setup so that it can be added quickly when necessary.

The level of metal carbonyls and particulates in the treating solution can be reduced by filtration and filter suppliers recommend a variety of operating schemes and filter types. To avoid contamination of the regenerator, rich-side filtration is recommended, though worker safety must be addressed when H,S is present in the rich solution. At a minimum, 10-20% slipstream filtration should be coupled with carbon filtration of the cool lean solution. The more filtration of the working solution provided, the better the solvent will perform and the more trouble-free the unit operation.

Heat Stable Amine Salts

Perhaps the most significant contaminants are Heat-Stable Amine Salts (HSAS). These are formed when the basic solvent reacts with a relatively strong acid. HSAS are one of the more intractable results of contamination. Although degradation of the amine can also lead to their formation, HSAS precursors are usually introduced with the synmesis gas.

The introduction of any relatively strong acid into the amine system will result in the formation of HSAS, a reduction in pH, and deactivation of the solvent from an acid gas removal perspective. If instead of reacting with a weak acid in the synthesis gas such as H2S (which has a pK, of 7.05 at 20^{C}C) the solvent reacts with a stronger acid (pK_a <6) it proves impossible to reverse this to any great extent at normal stripper conditions. The resulting salt is said to be heat-stable because of this inability to reverse the reaction. For example, with formic acid (pK_t 3.76), the amine formate salt is readily formed but cannot be reversed:

R,N + HCOOH •* R_3NH + HCOO' + Heat (1)

Low levels of acids or acid precursors are absorbed into the solvent from the synthesis gas being treated. Since they can only be lost from the system via mechanical losses, and not by vaporization, they tend to steadily accumulate. Impurities in the gasifier feed stock can lead to the introduction of HSAS in the solvent but one of the major sources of HSAS may be carbon monoxide (CO), which can lead to the creation of formate anions.

While the partial pressure of CO in synthesis gas can vary widely, it is true to say that it is always significantly higher than that encountered in other gas treating applications. One unfortunate result of this is the generation of formates, most likely as a consequence of the following simple reaction [8]:

OH* + CO -^ HCOO" (2)

Although the rate of reaction (2) is very slow, it is irreversible and formate HSAS will steadily accumulate in solution over time. For example, in one system treating synthesis gas with a CO partial pressure of 10 bar, formate anions build at a rate of 150 ppmw/day. In addition to the CO partial pressure, the rate of formate formation via this route increases as a function of increasing pH and temperature. Unfortunately, reducing any of these three parameters is either impractical or counter-productive to the main purpose of the cleanup unit, which is removal of H_2S and/or $C0_2$ to very low levels.

Other potential routes to formate from CO are via amide or formate ester intermediates, particularly in total CO, removal systems, but these routes need further verification before they are proven. Fortunately amides themselves do not pose any significant corrosion problems compared to formate anions. Suffice it to say that, whatever the mechanism, formate accumulation is an unavoidable consequence of treating synthesis gas and provision has to be made ahead of time to control, mitigate, and ultimately remove formates from the system.

Two nitrogen based contaminants, ammonia and hydrogen cyanide (HCN), are often encountered and are absorbed from the synthesis gas by chemical-based treating solutions. The various water-wash and quench systems upstream of the acid gas cleanup unit should remove the majority of these species but a small amount will still get through to the treating system. Ammonia does not lead to HSAS formation and can be easily removed by purging regenerator reflux water. However, cyanide incursion is a more serious problem since it can be quite corrosive, forming soluble ferrocynaide complexes, as well as acidic anions which form HSAS. HCN itself is a weak acid but it reacts in a basic solution and converts to stronger acids that do form corrosive heat-stable amine salts.

Problems Associated With Heat Stable Amine Salts

The greatest problem posed by HSAS is the increased potential for corrosion [9, 10] Although there is no definitive explanation for HSAS corrosion one promising hypothesis is that as the anion level increases so does the level of undissociated acid in equilibrium with the anion [11]. The undissociated acid is the active species promoting corrosion by catalyzing the cathodic reaction. If the pH and acid loadings (HSAS as well as acid gas) are known, the level of undissociated acid can be calculated, taking into account the amine and acid pK, values. This exercise reveals that the most corrosive HSAS are those associated with the medium strength acids (e.g. formic, acetic and glycolic) rather than the stronger acids, since the former lead to the greatest concentration of undissociated acid in solution. Higher temperatures increase the concentration of undissociated acids, making hot lean areas of the treating unit particularly susceptible to corrosion.

By themselves, the typical HSAS encountered in gasification applications (formates and thiocyanates), being soluble and ionic in nature, do not promote foaming. However, by increasing corrosion rates they can increase the particulate load and thus indirectly cause foaming.

HSAS Control & Removal Strategies

The best solution to HSAS problems is to prevent the precursors from entering the amine system in the first place. The pre-wash systems discussed earlier should achieve a good degree of reduction. However, with unavoidably high CO partial precjures, pre-washing will not eliminate all HSAS problems. Options have to be available to control and treat HSAS problems when they do occur.

One apparently simple solution to increased HSAS levels is to purge contaminated solvent and makeup with fresh material. Unfortunately this significantly increases the operating costs of the cleanup unit. The biological oxidation demand on the waste treatment system is also increased when any contaminated solvent solutions are sent to the sewer. With new discharge limits imposed on waste treatment systems this is not always a feasible proposition. Deliberate purge-and-makeup is thus neither an economically nor environmentally attractive option.

In MEA systems HSAS problems can be taken care of by the use of a reclaimer; a semi-batch distillation process operated at atmospheric pressure. A slip stream of MEA solution is fed to the reclaimer and water and MEA are stripped overhead, leaving behind MEA degradation products, HSAS and, if used, corrosion inhibitors. This approach cannot be applied to systems running on MDEA or MDEA-based formulated products since the atmospheric boiling points of MEA and MDEA are 171 °C (340°F) and 247°C (477°F) respectively. Significant thermal degradation would result if MDEA was reclaimed at atmospheric pressure. The costs and operating complexity associated with setting up an on-line vacuum reclaimer are considered prohibitive. Therefore, to fill the need for on-line reclaiming of MDEA-based specialty products, Union Carbide developed an electrodialysis process. This technology, commercially known as the UCARSEP® Process, has been successfully used in the field [12]. By coupling this with a strategy of HSAS control via neutralization the advantages of this technology are further strengthened.

HSAS Control Via Neutralization

One proven means of mitigating the effects of HSAS is to neutralize using a stronger base than the amine in question. This will raise the system pH, deprotonate the amine and render it available for gas removal purposes again. The overall effect is shown below:

H $R,NtT + OH^{-}$ •» $R,N + H,0$ (3)

There is a lot of evidence in the literature for the benefits of neutralization as a means of controlling HSAS problems [9, 10]. More importantly, this is also supported by industry experience[13]. Caustic has been used as the strong base but this is not the most suitable since sodium salts are not always soluble in chemical solvent systems. Caustic can also be an unintentional source of chlorides and if not administered carefully, can lead to concerns with over-neutralization, precipitation/fouling, and stress corrosion cracking. A proprietary neutralizing agent has been developed by Union Carbide for use in the many situations where caustic proves unsuitable.

Union Carbide's experience has been that without neutralization, HSAS anion levels of up to 10,000 ppmw can be tolerated without significant corrosion. Since neutralized salts are less corrosive than the corresponding amine salts [9,10], a higher level of anions are permissible if neutralization is practiced. Trouble-free operation with anion levels as high as 50,000 ppmw is possible with judicious and regular neutralization to maintain the HSAS levels at -1 wt%. Any system's corrosion-free contaminant level is influenced by the type of anion as well as its concentration. Although it can be misleading to generalize, it has been found that a limit of <30-40,000 ppmw anions (coupled with <l-2 wt% HSAS) to be both a safe and practical target Neutralization is thus a very pragmatic and effective solution to the HSAS problem. For example, after implementing a program of neutralization, one user reported greatly improved operation: comparing the six month period before and after treatment, the number of heat exchanger washes was reduced from four to none, the number of absorber washes went from ten to none, and the number of filter changes was reduced from sixteen to four [10].

Being able to operate safely at higher anion levels has the added benefit of extending the time before solvent reclamation is required. Depending upon the relative rate of incursion and loss, the need for reclamation may be averted completely. As the level of contamination increases, mechanical solution losses, which are fairly constant if viewed over a long enough time frame, account for larger and larger contaminant losses. This increases the time between reclamation and decreases the amount of salt that has to be removed when reclamation is required. Taking advantage of unavoidable system losses in this way is far removed from setting up a deliberate purge-and-makeup procedure to control HSAS.

Electrodialysis Technology for HSAS Removal

Electrodialysis (ED) has been widely used in the water treating industry for many years. Recognizing that it had beneficial characteristics for salt removal and fit very well with neutralization, Union Carbide adapted it to the unique conditions encountered in acid gas cleanup [14].

ED is a separation process in which ion permeable membranes are placed in an electric field to facilitate the removal of substances that ionize in solution. These semi-permeable membranes contain electrically charged functional sites chosen such that they are selective and allow the passage of either anions or cations, but not both. By correct sequencing, anions and cations can be extracted from one solution into another as shown in Fig 2. The membranes are sequenced such that when the contaminated amine solution enters the channel between an anion and cation permeable membrane, the anions move towards the anode through the anion permeable membrane, and the cations move towards the cathode through the cation permeable membrane. On the other side of both membranes an aqueous brine solution flows and the transported ions are collected and swept out of the system for disposal.
The technology can be tailored to the specific requirements of any treating unit to provide a dedicated on-site HSAS removal capability. Typically several hundred cell pairs are required but the exact number and membrane area needed are governed by the required salt removal duty. However, the overall process is very compact and the space requirement is small.

For systems where a permanent unit cannot be justified because die contamination problem is periodic or controllable through judicious neutralization, a mobile ED unit capable of removing up to -0.2 mole/sec of salts has been built. The unit can be brought on-site and cleanup on-line in a minimal amount of time. Only a small slip-stream of contaminated lean amine solution is required (typically *<l%* of circulation) and experience has shown that the operation of the treating unit is not affected by the reclamation. The process is fully automated and operates 24 hours per day. Process and utility hookups are simple and power consumption costs are minimal. A source of good quality water for brine make-up is required. Water has to be added to the brine loop to maintain a constant salt concentration in the brine, but water is neither added nor taken out of the solvent itself.

One of the benefits of the ED process is that the aqueous brine stream produced is considered to be biodegradable and non-hazardous. The brine is homogeneous, has a pH typically in the range of 9-10, and does not require any post-treatment before discharge to a conventional waste water treatment system. Unlike conventional ion exchange absorption processes, the volume of brine is simply proportional to the amount of salt removed since flushing or back-washing with rinse water or regeneration chemicals is not required [15]. In this way the hydraulic load and biological oxygen demand on the waste water treatment system are minimized.

Summary

Chemical solvent-based processes are well suited for acid gas cleanup of gasifier product streams. The combination of a specially formulated solvent and well designed equipment can be used to meet a variety of product gas and/or emission requirements. Superior selectivity for H_2S over $C0_2$, enhanced COS removal and efficient total $C0_2$ removal can be achieved more economically with specialty solvents than with corresponding generic amine solutions.

In providing the intimate contact necessary to meet H_2S and CO, specifications, there is more than adequate contact for a chemical solvent to absorb other contaminants from the gas. Solvent contamination can be directly linked with increased levels of foaming, fouling, and corrosion. These symptoms in turn result in increased solvent losses, off-specification operation, and possible equipment failure/replacement. The extent to which the acid gas cleanup systems can handle these diverse contaminants, or to which provision is made to remove them upstream, will have a great impact upon the operability of the unit and overall plant reliability. Since reliability is a key concern in gasification applications it is imperative that these issues are taken into account at me design stage.

References

1. "Gas Purification", 4th ed., Chapter 2, A.L. Kohl and F.C. Reisenfeld, Gulf Publishing Company, Houston, 1985.
2. "Comparison of Physical Solvents Used for Gas Processing'" R.W. Bucklin and R.L. Schendel, pp .42-80, Acid and Sour Gas Treating Processes, ed. S.A. Newman, Gulf Publishing Co., 1985
3. "Purification and Recovery Options for Gasification", DJ. Kubek, E. Polla, and F.P. Wilcher, Presented at "Gasification Technologies Conference", San Francisco, Oct 2-4, 1996
4. 'Tertiary Ethanoloamines More Economical For Removal of H2S and Carbon Dioxide", F.C. Reisenfeld et al, Oil & Gas Journal, pp 61-65, 1986
5. "Increase Amine Unit Capacity With UCARSOL Solvents", D.L. Dwyer, Presented at AIChE Spring Meeting, Houston, TX, 1993
6. "The Shell Gasification Process for Conversion of Heavy Residues to Hydrogen and Power", L.O.M. Koenders, S.A. Posthuma, P.L. Zuideveld, Presented at "Gasification Technologies Conference", San Francisco, Oct 2-4, 1996
7. "Analysis of Foaming Mechanisms in Amine Plants", C.R. Pauley, R. Hashemi, S. Caothien, Presented at the AIChE Summer Meeting, Denver, CO, August 22-24, 1988.
8. "Absorption of Carbon Monoxide into Aqueous Solutions of KJCOJ, Methyldiethanolamine, and Diethylethanolamine", C.J. Kim, A.M. Palmer, and G.E. Milliman, Ind. Eng. Chem. Res., Vol 27, pp 324-328, 1988
9. "The Role of Anion Contaminants on Corrosion in Refinery Amine Units", L.E. Hakka, S.F. Bosen and H.J. Liu, Presented at the AIChE Spring Meeting, Houston, TX, 1995.
10. "Neutralization Technology to Reduce Corrosion from Heat Stable Amine Salts", H.J. Liu, J.W. Dean, and S.F. Bosen, Paper 95572, Presented at "Corrosion in Gas Treating", Corrosion 95, NACE International Annual Conference and Corrosion Show, Orlando, FL, March 29, 1995.
11. "Controlling Corrosion in Amine Treating Plants", R.B. Nielsen, K.R. Lewis, J.G. McCullough, and D.A. Hansen, Laurence Reid Gas Conditioning Conference Proceedings, University of Oklahoma, Norman, OK, 1995.
12. "Clean Amine Solvents Economically and On-line", J. Price and D. Burns, Hydrocarbon Processing, Vol 74, No. 8, pp 140-141, August 1995.
13. Proceedings of the 1994 NPRA Question & Answer Session on Refining & Petroleum Technology, Session V.A, Question 3, Washington D.C. 1994
14. "Removal of Salts From Aqueous Alkanolamines Using an Electrodialysis Cell with Ion Exchange Membranes", R.A. Gregory and M.F. Cohen, European Patent 286143, 1988
15. "The On-Line Removal of Non-Regenerable Salts from Amine Solutions Using the UCARSEP Process", D. Burns and R.A. Gregory, Seventy-Fourth Annual GPA Convention Proceedings, San Antonio, TX, March 13-15, 1995

Figure 1 - General Flow Sheet for Chemical Solvent Based Process

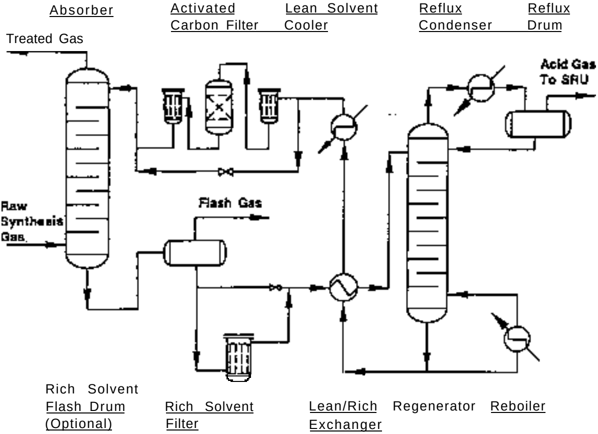


Figure 2 - Principle of Amine Reclamation Via Electrodialysis

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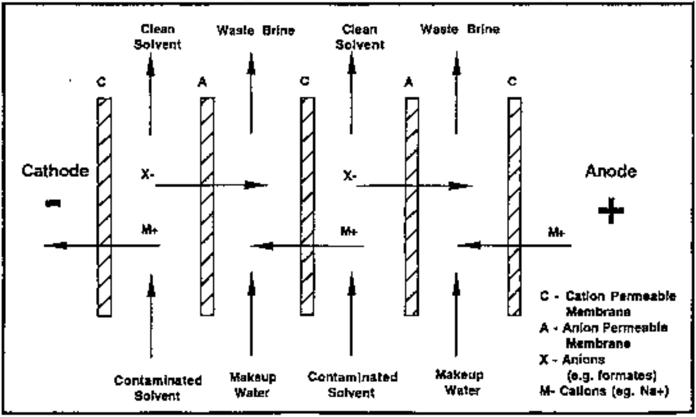

ROSE TREATMENT TO RAISE THE PROFITABILITY OF REFINERY RESIDUE GASIFICATION COMBINED CYCLE (GCC)

By

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INTRODUCTION

The refining industry in Europe is facing significant changes due to the progressive tightening of product quality specifications, changing patterns of product demand, persistent low margins and pressure on capacity utilisation. These factors are leading to refinery closures and asset restructuring deals. There is a well-established long-term trend, particularly in Europe, for a reduction in the production of heavy fuel oil from refineries. The three proposed Italian Refinery Gasification Combined Cycle Projects are a reflection of this trend, a trend which is a direct consequence of the fall in demand for heavy fuel oil both in Europe and elsewhere. In addition the sulphur content of fuel oils is also under pressure in most advanced industrial countries. In Europe, these factors are underlined by the pressures exerted by the environmental lobby on politicians and governments at both national and supra-national levels.

This paper describes one approach showing how a European refinery containing a visbreaker unit, can reduce the production of heavy fuel oil in a progressive, stepwise manner. The underlying philosophy is to maximise the utilisation of existing refinery units whilst installing appropriate additional plant units in a stepwise fashion to achieve operating objectives. Preliminary economics are also presented for the solutions proposed.

The paper uses The M. W. Kellogg Technology Company's 'state-of-the-arf ROSE™ (Residuum Oil Supercritical Extraction) solvent deasphalting unit technology to reduce the amount of high sulphur fuel oil produced by a Refinery. In general, a ROSE unit can take a feedstock of either vacuum residue or visbroken vacuum residue and recover a deasphalted oil fraction (DAO), typically for feeding to a Fluid Catalytic Cracking Unit (FCCU) or hydrocracker. In this paper we use the example of adding a ROSE unit to a refinery containing an existing visbreaker unit and with visbroken vacuum residue being utilised as a ROSE unit feedstock. The DAO fraction is then fed to an FCCU. This paper is one of a series (Ref. 1-4) dealing with applications of ROSE technology.

This paper also draws on the examples offered by the proposed Integrated Gasification Combined Cycle (IGCC) Projects in Italy which provide an effective and environmentally-friendly means of disposing of asphaltene-rich residues from a refinery as the feedstock to an adjacent IGCC power plant. This is a realistic solution where there is demonstrable demand for additional electrical power and where the electricity can be sold at a reasonable price. The nature of the proposed Italian refinery residue to electricity IGCC projects is discussed in the next section.

REFINERY GASIFICATION COMBINED CYCLE PROJECTS

Refinery IGCC projects disposing of refinery residues and exporting base load electricity are expensive by refinery standards (S5,000-$10,000 per BPSD). Until now, most refiners have preferred to use gasification on a small scale in units producing high value utilities such as hydrogen. Such units are under refinery management and control whereas bottom-of-the-barrel gasification plants require support and assistance external to the refinery.

In operation, a refinery plant supplying residues to a power station is demand driven and has to sell electricity at all times at the price set by oil product markets, and not by the market price for the exported commodity, generally electricity. In practice, this makes the power plant a base-load supplier. As the power station is a base-load supplier, the refinery has an obligation to provide feed to the power station, possibly even during refinery shutdowns. This aspect has to be carefully considered as a typical feed rate for a 500 MW IGCC Plant is 120 tonne per hour of residue. To guarantee an electricity supply for a short 7 day refinery shutdown would require a feedstock storage capacity of 20,000 tonne. An alternative during refinery shutdowns would be to substitute the feedstock with an imported supply of heavy fuel oil. This is only feasible if there is sufficient flexibility built into the IGCC design.

The most recent commercial application of gasification is in Italy where one major refinery residue to electricity IGCC project is underway and two more are close to financial closure. They are owned by joint venture companies outside the refinery but in which the refiner has a major equity stake together with a specialised investor. The joint venture company negotiates a supply contract with the refinery and an offtake contract through the agency of the specialised investor. These projects utilize quench-based gasification technology schemes (and therefore capable of accepting a large range of feedstocks) which have all qualified for financing.

API-API Energia

API Energia is located at Falconara on the Adriatic Coast and is 50% owned by ABB who is also the Main Contractor, supplier of the integration scheme and sub-Contractors for the combined cycle unit. Net output is 233.5 MW.

ERG Petroli - **ISAB** Energy

The only one of the three projects to integrate ROSE deasphalting technology, this project is now underway as the financial package was agreed recently (termed 'financial closure'). This project is located at the Sicilian Refinery of ERG Petroli at Priolo. The Project is 51 per cent owned by ERG Petroli and 49% by Mission Energy, a subsidiary of Southern California Edison. High pressure Texaco Quench Gasifiers will feed two Siemens V94.2 gas turbines modified and supplied by Ansaldo, Genoa. The 500 MW plant is integrated with a Jacobs Engineering patented CPG (Clean Power Generation) flow scheme (5) which uses a fuel gas expander coupled with a water desaturation/resaturation circuit.

- Saras - Sarlux

Sarlux is located at the Saras refinery at Sarroch, Sardinia. The Project is 60% owned by Saras and 40% by Enron. Output is 500 MW from a combination of Texaco Quench Gasifiers and GE gas turbines.

RESIDUE UPGRADING SOLUTIONS

The philosophy taken in this paper is to maximise the utilisation of existing refinery equipment whilst considering the addition of plant units in a stepwise manner to reduce heavy fuel oil production. An existing 'generic* refinery and the solutions proposed are all based on processing 200,000 BPSD of Arabian Light crude oil (33.2 °API) to produce an estimated 28,500 BPSD of vacuum residue (565 °C plus cut, 6.5 °API) as bottoms from the vacuum tower.

Base Refinery Configuration

The base refinery processing the above crude includes atmospheric and vacuum distillation units, visbreaker, VGO hydrotreater and an FCC. All the vacuum residue from the vacuum unit (28,500 BPSD) is fed to a single-pass visbreaker as shown in Figure 1. The visbreaker unit involves mild thermal cracking of vacuum residue in a visbreaker furnace and the furnace effluent is separated in atmospheric and vacuum columns to produce an estimated 71 weight per cent yield of visbreaker vacuum tar as bottoms from the vacuum unit The visbreaker vacuum tar is blended with cutter stock comprising FCC cycle and decant oils and kerosene to produce a saleable high sulphur fuel oil with acceptable specific gravity, sulphur content, viscosity and cold flow properties. Viscosity blending has been carried out using the Refutas method. A material balance for the visbreaker is presented in Table 1.

To simplify economic considerations, the FCCU is considered to operate on a mixture of hydrotreated VGO from the vacuum and visbreaker vacuum units and separately purchased (unhydrotreated) atmospheric residue. The impact of Options 1 and 2 below is to substitute part or all of the atmospheric residue with DAO, with the simplifying assumption being made of minimal impact on FCCU throughput, product yields and catalyst make-up rate and hence no allowance has been included for the impact on FCCU economics.

The following two processing options are considered :

OPTION 1 Add a 19,400 BPSD Solvent Deasphalting Unit Downstream of the Visbreaker.

This solution is characterised by the addition of the ROSE solvent deasphalting unit, which is exclusively licensed by The M W Kellogg Technology Company, to process 19,400 BPSD of vacuum tar from the visbreaker. In the ROSE process, vacuum residues or in this case visbroken vacuum residues are physically separated into deasphalted oil (DAO) and asphaltenes.

The ROSE process is well-established commercially and has been used extensively for the following reasons:

i) To reduce heavy fuel oil production
ii) To add to the feed to an FCCU or hydrocracker
iii) To displace purchases of atmospheric residue
iv) To relieve overloaded vacuum towers
To manufacture bitumen
vi) To replace delayed cokers
vii) To produce lube bright stocks

In both options in this paper, ROSE is being used for reasons i) and iii) above. The ROSE process is relatively inexpensive in terms of both capital cost and operating expense when compared to residue conversion processes such as delayed coking or residue hydrocracking.

A material balance for Option 1 has been prepared in Table 2 and a schematic in Figure 2.
Because of pumping requirements for the asphaltene fraction and also the gasification burners in Option 2 (see below) may require a minimum asphaltene ring and ball softening point, it is proposed to limit the DAO yield in both options to 38 weight per cent with a 62 weight per cent yield of ROSE bottoms (asphaltenes).

ROSE bottoms are blended with cutter stocks comprising FCC cycle oils, decant oil and kerosene. For Option 1 the total amount of high sulphur fuel oil produced from this blend is estimated at 2060 kilotonne per year, a 19 per cent reduction on the Base Case. As in the Base Case, blending is carried out to produce a fuel oil with acceptable specific gravity, sulphur content, viscosity and cold flow properties.

OPTION 2 Add Integrated Gasification Combined Cycle Power Generation Plant fed by the Bottoms from a Solvent Deasphalting Unit

This solution involves the addition of an Integrated Gasification Combined Cycle (IGCC) Power Plant to gasify ROSE bottoms and produce electrical power. A material balance has been prepared on this basis in Table 3 and a schematic in Figure 3. As explained above, it is proposed to limit the DAO yield in both options to 38 weight per cent with a 62 weight per cent yield of ROSE bottoms (asphaltenes).

The gasification of ROSE asphaltenes and subsequent combustion of syngas results in a further substantial reduction in fuel oil production as cutter stock is no longer required. The 915 kilotonne per year of FCC oils containing an estimated 2 per cent sulphur can be sold directly as medium sulphur fuel oil (or the LCO could alternatively be upgraded to diesel by suitable hydrotreating).

The IGCC Power Generation Plant proposed is based on a 'developed plant design' utilising Jacobs Engineering's patented Clean Power Generation (CPG) flowscheme. A Mediterranean location is assumed with a design ambient air temperature of 35 °C. This results in a lower overall thermal efficiency for the GCC Plant than would be expected in a more temperate climate. Net power output is estimated at 360 MW based on the 41 per cent overall thermal efficiency assumed here.

Technical Discussion

The Kellogg ROSE™ Process Technology utilises super-critical fluid technology to recover a deasphalted oil (DAO) fraction and as a result offers a major operating cost advantage over a conventional solvent deasphalting unit An existing conventional solvent deasphalting unit is easily revamped to the more efficient ROSE technology with attractive benefits in terms of reduced operating costs.

An appropriate outlet is clearly required for tbe DAO as well as the asphaltenes. In this paper, the DAO fraction produced in Options 1 and 2 from the ROSE unit is fed to an FCCU. Alternatively the DAO may also be an excellent feedstock for a hydrocracker.

A key result of adopting Options 1 or 2 is to significantly reduce the amount of heavy fuel oil produced. Fuel oil production declines by 19 per cent in Option 1 (adding ROSE) and a much more substantial 64 per cent in Option 2 (adding ROSE plus IGCC). The latter option illustrates the major benefit in terms of reduced fuel oil production of combining the two plants.

The attractiveness of Option 1 can be further significantly improved with significantly higher economic returns if part of the bottoms from the ROSE unit can be sent to a bitumen unit in the Refinery. This results in substantial savings in the quantities of cutter stock (kerosene or gasoil) required to blend with the remaining bottoms and hence achieves a further reduction in the amount of heavy fuel oil produced in Option 1.

The flue gas from the IGCC plant in Option 2 offers the prospect of fairly low SOx and NOx emissions, and these emissions can be further reduced by a catalytic treatment step or steam injection in the gas turbine.

Basis of Economic Analysis

Preliminary economics are presented for each option on a Mediterranean (European) location with capital costs, operating costs and price set on a fourth quarter 1995 basis. The price set used is given in Table 4. As previously mentioned, a basic assumption of this paper is that deasphalted oil (DAO) from the ROSE unit replaces purchased atmospheric residue. Because of this and similar expected yields over an FCCU compared with atmospheric residue, DAO is valued at parity with atmospheric residue. (Based on client information, atmospheric residue is priced at LSFO value plus $13 per tonne). As cutter stock, FCC cycle oil is valued here at diesel minus S3 per barrel and decant oil at HSFO value. Kerosene is used in preference to gas oil as an additional but expensive low viscosity cutter stock.

Economics for the IGCC Power Plant in Option 2 are based on a 'developed' plant design and a corresponding capital cost of $1500 per kilowatt of net power output, which includes the capital cost of a sulphur recovery unit but excludes the cost of an air separation unit. This is believed to be lower than the likely capital cost for the Italian projects, but reflects potential for capital cost savings on future projects. The cost of oxygen (at 95 per cent purity) fed to the GCC Plant is taken at an 'over the fence' figure of $41 per tonne.

For both options, a time on stream of 8000 hours per annum is assumed. This may actually slightly underestimate the availability of the IGCC Power Plant in Option 2. The plant availability is likely to be determined by the availability of the turbines. By allowing for additional production and storage capacity, returns and cash flow for Option 2 may thus be enhanced by a further few per cent.

A discounted cash flow approach has been adopted based on the incremental pre-tax cash flow relative to the base refinery configuration to obtain the internal rate of return (the discount rate at zero NPV for the Project) for both Options 1 and 2. A 20 year life is assumed for both ROSE and IGCC Plants.

For Option 2, the sale price for electricity from the IGCC Plant is taken as $0,075 per kWh for years 1 to 8 and $0,045 per kWh thereafter, reflecting an element of subsidy in the early years of the Project (believed to be the situation for the Italian Projects). An overall thermal efficiency of 41 per cent is used for the IGCC Plant, applicable to a relatively hot climate.

Results of the Economic Analysis

Estimated capital costs are $31 million for the ROSE Unit in Option 1, contrasting with the $571 million for the Rose Unit plus IGCC Plant in Option 2. ROSE unit cost qualifications and exclusions are given in Table 6. These major differences in CAPEX reflect the modest level of capital expenditure for the Rose Unit compared with the one and a half orders of magnitude higher capital cost of the IGCC Plant in Option 2, estimated at $540 million for a 360 MW plant. Obviously without the ROSE unit, the investment cost would be much higher.

Results of the preliminary economic analysis for Options 1 and 2 are presented in Table 5. The internal rate of return is estimated at 35 per cent for Option 1, equivalent to a simple payback period of 2.9 years. The corresponding internal rate of return for Option 2 is estimated at 28 per cent, and this is equivalent to a simple payback period of 3.1 years, based on the cash flow in years 1 to 8.

Economics for Options 1 and 2 are very sensitive to the price of HSFO. For example, if the price of HSFO were to decline from the S88 per tonne (LSFO at $101 per tonne) to S60 per tonne then the internal rate of return for Option 1 increases from 35 per cent to 80 per cent, and for Option 2 the internal rate of return increases from 28 per cent to 41 per cent. This reflects a reduction in low value HSFO production achieved by both Options 1 and 2 and indicates a possibility of living without price subsidies on electricity for Option 2.

Economics for Option 2 are highly sensitive to the price of electricity and the capital cost of the IGCC Plant. Without the higher (subsidised) price for electricity in the early years of the Project, the internal rate of return for Option 2 falls from 28 per cent to 14 per cent. To reiterate, economics as summarised in Table 5 are based on a 'developed' capital cost of the IGCC Plant of $1500 per kilowatt of power output, excluding an air separation unit (ASU). Industry estimates of the minimum achievable capital cost of an IGCC Plant are currently pitched at about $1200 per kW, with little dependence on scale and this figure may or may not include the air separation unit. For an IGCC Plant capital cost of $1200 per kW (excluding the ASU) and the price set in Table 4, the internal return for Option 2 (Rose plus IGCC) is estimated at 19 per cent without any electricity price subsidy, suggesting that if the IGCC Plant capital cost can be reduced to this level, no subsidy, after all. is needed to achieve a viable project based on the ROSE plus IGCC Plant configuration.

The location of the IGCC Plant can also influence the rate of return for Option 2. For a colder climate and an overall thermal efficiency of 45 per cent rather than the 41 per cent assumed above, the IRR for Option 2 is increased by about 4 per cent, from 28 per cent to 32 per cent, assuming the same price set.

Summary

This paper demonstrates the economic attractiveness of utilising the ROSE technology to reduce the amount of high sulphur fuel oil (HSFO) produced by a refinery with a visbreaker. Good returns and short payback periods are achievable. The lower the price of HSFO relative to crude and other products, the more attractive the economics get.

This paper also demonstrates the validity of the gasification combined cycle route as a capital - intensive yet environmentally-friendly means of eliminating the bottom of the barrel from a refinery, probably through a joint-venture company. The attractiveness of the ROSE™ plus IGCC Plant option is likely to increase once the 'dash for gas' is behind us and large quantities of natural gas are no longer available at bargain prices. In either case, incorporation of ROSE technology is the key to attaining an economically viable project

A ™ This paper also suggests that if the capital cost of the IGCC Plant could be reduced to the bottom end of industry estimates, or if the price of HSFO relative to crude and other products were to fall significantly from today's levels, then the ROSE plus IGCC option would be economically viable without any electricity price subsidy. The location of the plant can also have an influence, both via product prices and the overall thermal efficiency of the Plant.

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Cost Effective Approaches to Heavy Oil Conversion, Degnan J.J.. Sloan H.D. and Simons H.J., Financial Times Ninth Petroleum and Gas Conference, European Oil Refining and the Market to the Year 2000, Amsterdam, 15-16 Nov 1994

Bottom of the Barrel Upgrading Utilising Rose Technology, Sloan H.D.,
Presentation to The MW Kellogg European Refining Conference, London, 12-13 Oct 1995

First Steps Towards Major Residue Conversion Projects, Simons H.J., Ditz J.M., Sloan H.D.and James J., The MW Kellogg European Refining Conference, London, 12-13 Oct 1995

Advances in Solvent Deasphalting Technology, Northup A.H. and Sloan H.D.,
National Petroleum Refiners Association Annual Meeting. San Antonio, Texas. 17-19 Mar 1996

Development of an IGCC Flowscheme for ISAB S.p.A., Sicily, Griffiths J. and Magrath A.F.,
ASME Cogen Turbo-Power 1993 Conference. Bournemouth, Sept 1993

TABLE 1

BASE CASE BALANCE
VISBREAKER

FEED (V.R.)	1548 kt/yr
CUTTER STOCK	1426 kt/yr

PRODUCTS ON VACUUM RESIDUE

PRODUCT	WT %	**kt/yr**
Gas	2.7	43
Cracked Naphtha	6.2	96
Atm Gas Oils	8.4	130
Vacuum Gas Oils	11.2	173
Visbroken Vac Tar	71.5	1106

Total Fuel Oil = 2532 kt/yr

CUTTER STOCK COMPRISES FCC CYCLE & DECANT OILS AND KEROSENE

TABLE 2

OPTION 1 BALANCE
VISBREAKER + ROSE

FEED (V.R.)	1548 kt/yr
CUTTER STOCK	1375 kt/yr

PRODUCTS ON VACUUM RESIDUE

PRODUCT	**WT** %	**kt/yr**
Gas	2.7	43
Cracked Naphtha	6.2	96
Atm Gas Oils	8.4	130
Vacuum Gas Oils	11.2	173
DAO	27.2	421
ROSE Bottoms	44.3	685

Total Fuel Oil = 2060 kt/yr

CUTTER STOCK COMPRISES FCC CYCLE &
DECANT OILS AND KEROSENE

Figure 1 : Base Case
Overall Process Scheme and Material Balance

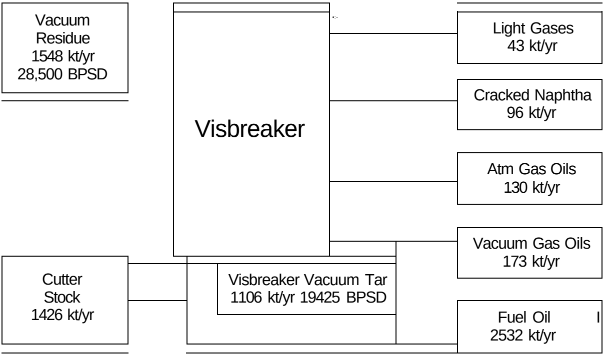

Figure 2 : Option 1
Overall Process Scheme and Material Balance

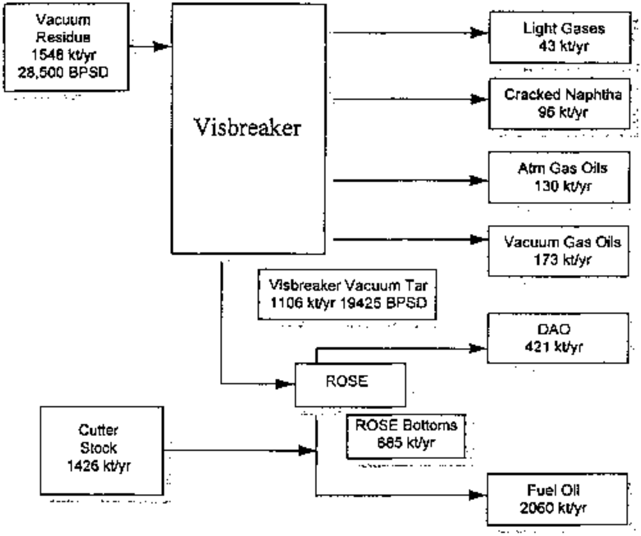

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TABLE 3

OPTION 2 BALANCE
VISBREAKER + ROSE + GCC PLANT

FEED (V.R.) CUTTER STOCK	1548 kt/yr NIL

PRODUCTS ON VACUUM RESIDUE

PRODUCT	WT%	kt/yr
Gas	2.7	43
Cracked Naphtha	6.2	96
Atm Gas Oils	8.4	130
Vacuum Gas Oils	11.2	173
DAO	27.2	421
Electrical Power (MW) (From ROSE Btms)	44.3	360 MW

FCC Cycle & Decant Oils to Fuel Oil = 915 kt/yr

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Figure 3 : Option 2
Rose Unit and GCC Plant Added to Visbreaker Overall Material Balance and Process Scheme

Vacuum
Residue
1548kt/yr
28,500 BPSD

Light Gases
43kt/yr

Cracked Naphtha
96 kt/yr

Visbreaker

Atm Gas Oils
130 kt/yr

Vacuum Gas Oil:
173 kt/yr

Visbreaker Vacuum Tar
1106 kt/yr 19425 BPSD

DAO
421 kt/yr

360W Electrical
Power Output

ROSE Bottoms
685 kt/yr

Cutter Stock
to Fuel Oil
915 kt/yr

Gasification
Combined
Cycle Plant

w

TABLE 4

PRODUCT PRICE SET USED

Spot Prices, Mediterranean, 4Q95 Average

PRODUCT	PRICE
Kerosene (cutter stock)	167.0
DAO	114.3
Base Cutter Stock (FCC Cycle/Decant Oils)	104.5
HSFO	88.0

ELECTRICAL POWER	**$/MWh**
Year 1 to 8 (Subsidised)	75.0
Year 9 + (no subsidy)	45.0

Oxygen priced at $41.3 per tonne

TABLE 5

PRELIMINARY ECONOMICS

CASE	**DESCRIPTION**	**CAPEX ($MM)**	**IRR (%)**	**SIMPLE PAYBACK PERIOD (YRS)**
Option 1	Add Rose	**31**	**35**	**2.9**
Option 2	Add Rose + Gasification Combined Cycle Plant	**571**	**28**	**3.1**

Notes:

1) Economics are on a pre-tax, cash flow basis.
2) Uses price set in Table 4.
3) Capex excludes working capital requirements.

TABLE 6

ROSE UNIT COST QUALIFICATIONS

INSIDE BATTERY LIMITS COST

ORDER OF MAGNITUDE, 4Q95, U.S.G.C. FACTORED TO U.K.

LOCATION FACTOR OF 1.0 RELATIVE TO U.K. TAKEN FOR UNSPECIFIED MEDITERRANEAN LOCATION

CLEAR, LEVEL ACCESSIBLE SITE

SPREAD FOOTING FOUNDATIONS

UTILITIES AT THE BATTERY LIMIT

40 HOUR WORK WEEK

DISTILLATES AND POWER FROM REFINERY HEAVY RESIDUES

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Abstract

A process is under investigation in Eniricerche which integrates distillate recovery and power production from refinery heavy residues. Thermochemical treatment of the feed is accomplished in two subsequent steps, pyrolysis and gasification, while electric power is generated in a high efficiency combined cycle. The pyrolysis of the residue allows to obtain a mixture of gases and vapours, plus a substream of coke. The coke is fed to a gasifier to produce a fuel gas, which is then mixed with the pyrolysis gas, cleaned and finally burned in a combined cycle.

An externally heated rotary kiln is being considered for the pyrolysis step to attain continuous operation.

For coke gasification, the use of a moving bed catalytic reactor is under evaluation. As catalytic gasification needs less oxidant to be operated, air can be used instead of oxygen, avoiding the costly and energy demanding air separation unit. Besides that, the low level and the chemical composition of the coke ash can translate into high recovery and low deactivation rate of the catalyst, minimising the necessary make-up stream.

The process performances are being evaluated through computer simulation, enabling to choose the best integration options for the various components considered. Experimental activity on both pyrolysis and gasification and gasifier mathematical modelling are being extensively used to gain all the necessary inputs to process simulation.

Introduction

Heavy crudes with high metal and sulphur content require very severe processing. A simple way to utilize their residues is blending them with gasoil to produce a fuel to be burned in boilers or furnaces. This practice is becoming increasingly difficult, due to the more and more tightening environmental regulations and the reluctancy towards the implementation of expensive flue gas cleaning units: shifting to a cleaner even if more costly fuel is often the easiest and less troublesome choice.

The consequence of this is the search of alternative utilization routes, especially for the heaviest and more "difficult" residues.

In Italy, economic and regulatory pressures are at the base of a surge of interest in the process of gasification as a means to convert a variety of hydrocarbon feedstocks into a fuel gas and then into electic power in high efficiency low emission IGCCs.

In fact, the Italian State-owned power grid is currently forced to pay the electricity produced in this way considerably more than the free market would naturally allow. The IGCC option is thus made economically very attractive, despite its high .investment costs. The expected beneficial effects of this policy are the disposal of a potentially very unclean product in an environmentally sound fashion and

the limitation of the strong deficit in Italian electricity production system in a socially acceptable way and without further use of naturally premium fuels like natural gas.
However, when free market would eventually determine electricity price, less capital cost intensive residue conversion routes would probably be preferred. In addition, higher future market demand for light products is foreseen. A possible way by which an oil company can cope with these trends is to enhance its refinery conversion capacity into light products, feeding only the very ultimate residue to the gasification plant.

A preliminary investigation has been carried out in Eniricerche to evaluate an integrated process able to produce both distillates and power from heavy petroleum residues or asphalt. This process performs three main operations: pyrolysis, gasification and power production.
Pyrolysis is an inherently low capital intensive step and enables to decrease the size and hence the cost of the subsequent sections. In order to attain a continuous and flexible operability of the pyrolysis reactor, where flexible refers mainly to the feedstock quality, the adoption of a rotary kiln instead of a more conventional delayed coker can be considered, especially when a relatively low flowrate of a very "difficult" feed, like for example hydrogenation residues, is to be processed /1, *21.*
Gasification is used to convert the pyrolysis coke into a fuel gas. Air is used as the oxidant to avoid the economic and energetic costs of an air separation unit. Consequently, in order to keep the fuel gas heating value sufficiently high for its use in a standard gas turbine with only limited modifications, air consumption is minimized by performing the gasification process in presence of a catalyst. Moving bed reactor has been considered a suitable technology for catalytic gasification: it allows, in fact, to obtain a relatively low temperature of the fuel gas, with a higher cold gas efficiency and a lower loss of the possibly volatile catalyst, as a matter of fact, a catalytic moving bed reactor for petroleum coke gasification has not been developed to date, even if encouraging results on a laboratory scale have been obtained /3, 4/.

In Figure 1 the schematic view of the resulting process is shown, the name of which, Integrated Pyrolysis and Gasification Combined Cycle (IPGCC), comes after the presence of both a pyrolysis and of a gasification stage.
As our interest has been aimed at the utilization of very heavy and "difficult" residues, a rotary kiln pyrolyzer is adopted, which is fed by a mixture of the residue and a recycle stream of coke and mineral matter from the gasifier. The coke produced in the pyrolysis step is then fed to the gasifier. The recycle has the fundamental function of recovering the catalyst, still active at the exit from the gasifier. Furthermore, all the solid material at the gasifier bottom exit can be recycled back to the pyrolyzer and used there to improve pyrolysis reactions and dilute the feed, preventing agglomeration.
With less problematic feeds, a conventional delayed coking unit should be more conveniently used. In this case, the recycle stream from the gasifier must be sent back downstream the coking unit in the gasifier feed preparation unit.
After separation, the liquids and gases developed during the pyrolysis stage have different destinations: the first ones are sent back to the refinery, while the others, after compression, are mixed with the fuel gas produced in the gasifier and sent to a cold cleaning process and then to a high efficiency combined cycle.

Note that petroleum residues are usually characterized by a low ash content, which enables an easy recycle and a low make-up. Besides, due to ash composition, a low deactivation rate of the catalyst is possible: ash components are in facts mainly transition metals, which have generally a low interaction with the catalysts being considered, not poisoning or deactivating it and leaving its activity still unchanged when exiting from the gasifier to be recycled.

E.niricerche R&D activity on the IPGCC process

Some aspects of the IPGCC process call for some deep analysis to prove that its realization may be possible and really profitable.

First of all, an extensive work is necessary regarding the pyrolysis stage, to assess its operability in the required conditions and to give information such as mass and energy balances for the feeds of interest and the quality of the producible liquids, to be sure they can really be a valuable stream for the refinery.

The catalytic gasification stage must be thoroughly examined too, in particular as regards the advantages resulting from the catalyst use and the assessment of any limitation in the char massflow recycled back to the pyrolyzer.

There are five main activity lines in the IPGCC R&D project currently in progress in Eniricerche: pyrolysis tests, catalytic gasification kinetic studies, gasifier modelling, process simulation and economic evaluations. The first two imply experimental work on pyrolysis and gasification, while the last one collects the results of all the other lines to produce the final economic figures. In the following each line is analyzed in more detail.

Pyrolysis

To allow the collection of representative experimental pyrolysis data, a bench scale rotary kiln reactor has been built. The apparatus is conceived in such a way as to allow the study of a wide range of operating conditions, reaction temperature, gas/solid residence time, solid/liquid feed flowrates. Figure 2 and Figure 3 show the yields obtained for the feeds of Table 1. A remarkable amount of liquids turns out to be obtainable from pyrolysis and their quality seems to be interesting as well. In fact, a high percentage of them are middle distillates or even lighter fractions, while the hardest fraction constitues only a few percentage points.

Gasification: experimental

The relevant steam gasification reactions, based on carbon are /5, 6/:

$C + H_20 <=> CO + H,$ (1)

$CO + H_20 <=> CO, + H.$ (2)

$CO + 3H_2 <=> CH_4" + H,0$ (3)

$C + 2H_2 <=> CH_4 "$ (4)

The production of synthesis gas would be governed by equation 1), while the combination of reaction 1) and 2) leads to an increase of hydrogen concentration. A combination of reactions 1) to 4) leads to methane containing gas production.

As reaction 1) is strongly endothemic in all gasifiers heat has to be supplied at a high temperature level. When oxygen is used, heat is provided by partial combustion. Air separation can be avoided, but that implies a high nitrogen dilution of the product gas.

Anyway, an autothermal gasification could theorically be achieved by the combination of reactions 1), 2) and 3) performed in the same reactor; in fact at 923 K the overall reaction would be substatially thermoneutral :

$2C + 2 H_2O = CO, + CH_4 *$ (5)

To achieve this condition a very active catalyst is needed with the following funtionalities:
- high activity towards steam gasification: reaction 1);
- high activity and selectivity towards the formation of methane (and carbon dioxide).

Experimental work has been done to study the catalytic behaviour of a potassium-based catalyst in a pressurized thermobalance. The kinetics of petroleum coke catalytic gasification were investigated for the steam-coke reaction under highly differential conditions at pressures between 0.1 and 3.0 MPa.

Some important results are shown in Table 2, which demonstrate how the catalyst can improve gasification kinetics.

Furthermore, from reactivity data collected at 3.0 MPa catalyst addition is equivalent to a 150 - 170 °C increase of the reaction temperature: this enables a limitation in the temperature profile inside the gasifier, with consequent less oxydant demand, i.e. less nitrogen dilution when using air instead of pure oxygen.

Feed sulphur content does not seem to play a significant role in catalyst activity, a key factor when feed flexibility is considered.

The steam gasification kinetic data were confirmed in laboratory scale fixed bed gasification tests. Moreover, high activity and selectivity towards methane (and COj) formation were observed in tests with synthetic (CO, H2, H20) gas mixtures.

A catalyst recovery up to 96% was obtained after steam gasification tests at 750 °C.

Gasification: modelling

The modelling activity has been performed in order to assess the compatibility between the integration constraints and the most profitable operating conditions or the design requirements, namely the possibility to recycle a not negligible flow of solids, keeping low oxydant and steam consumptions and reasonable reactor dimensions. A computer code has been developed in Eniricerche for the purpose, giving temperature and concentration profiles along the reactor height *171.* When comparing catalytic and non catalytic gasification of coke, the differences concerning temperature profiles and gas yields, especially as regards methane content, can be easily observed.

From the modelling studies it has come out that there are some limitations in the recycle flowrate through the gasification reactor. In fact, with high recycle rates, the gasifier cannot be properly operated, due to a too high mismatch in solid and gas flowrates at the base of the gasifier. Thus, the recycle must be limited and cannot be used to provide the pyrolyzer with all the thermal energy it requires: the energy balance of this reactor is in fact fulfilled by burning some of the available fuel gas.

On the other hand, the fuel gas heating value is quite satisfactory. In conclusion, the mixture of this gas with that coming from the pyrolyzer seems to have enough energy density (ranging from 11 to 15 Mj/Nm^3 for the two feeds of Table 1) to feed a standard gas turbine without the necessity of extensive modifications.

Process simulation

While for the most unconventional part of the process experimental or modelling work is essential to obtain meaningful results, the remainder of the process, say the filtration and cleaning steps and the power island, belongs to almost standard design, except maybe for some aspects of the gas turbine operation.

A commercial computer code, ASPEN, is being used to evaluate the performances of the process by calculating mass and energy balances for all the individual components, the pyrolyzer, the gasifier, the combined cycle and so on. In principle, different arrangements could be analyzed in order to look for the best integration of all the single blocks.

For a feed flowrate of 64.1 t/h (500000 t/y @ 89% stream factor) of hard (pentane) asphaltenes ex-Ural a net power production of 195 Mw_e has been estimated. This figure drops to 140 Mw_e for an equivalent feed flowrate of Gela residue, due to the higher yield of distillates. These results have been obtained with a high performance gas turbine (firing temperature 1288 °C, compression ratio 14.7) and a double pressure plus reheat steam cycle.

Economic evaluations

A very preliminary economic analysis is being made, based on the available laboratory results, literature information and process simulations.

Compared to processes (like IGCCs) producing only power, the EPGCC, with a lower power output for the same feed flowrate, is favoured by a low electricity price scenario.

Furthermore, as it allows the production of a certain amount of distillates, the economic profitability of the IPGCC process depends, among other things, also on the yield and valorization of those distillates. This aspect is closely connected with the quality of the feed: for example, a too high sulphur content adversely affects the valorization of the distillates and the IPGCC option can turn out to be uneconomic, especially in presence of high distillate yields, as shown in the following.

For a plant size suitable for 500000 t/y feed flowrate (89% stream factor) of hard (pentane) asphaltenes ex-Urai a capital cost of about 400 million USS has been evaluated *(*± 25% accuracy). Assuming a cost of 32 USS/t for the feed and a price of 140 US$/t for the distillates and of 0.05 USS/kwh for the electricity, a DCF-IRR higher than 9% has been calculated (constant dollars, 15 years' plant operating life).

A similar evaluation for the other feed considered in this paper (Gela residue) shows a dramatic drop of the DCF-ERR figure. This is due principally to the higher cost of the feed (estimated at 50 USS/t), the enhanced yield of liquids with higher sulphur content and consequent low value (of the order of 110 USS/t) and the lower power production.

Conclusions

Some concluding remarks can be drawn at this point. Eniricerche R&D activity has allowed an evaluation of an original process for the utilization of very hard petroleum residues to produce power and distillates. There has been the confirmation of some very interesting technical features of the EPGCC that can be important assets: high flexibility with respect to the feed; ability to enhance refinery conversion capacity through the production of distillates; use of air in the gasification reactor, avoiding an air separation unit; high efficiency of the power island thanks to a combined cycle arrangement.

Also economically the EPGCC can be interesting in principle, especially when a high valorization of the distillates is possible and when no funding is present as favourable purchase prices for the electricity produced. It must be remembered that this last option is an eventuality which one can surely expect in the future, due to the political difficulties in granting those prices, especially in an European view.

References

I/ H. P. Wenning, Ruckstandsschwelung in der Drehtrommel, *Btindminister for Forschtmg und Technologie,* AbschluOericht 03E-6104 A, 1987.

/2/ A. Schleiffer, WeiterentwickJung der Ruckstandsschwelung in der Drehtrommel, *Btindminister for Forschung und Technologie,* AbschluGericht 0326617 A, 1990.

/3/ Y. Ohtsuka, A. Yamauchi, Q. Zuangh. Catalysis of petroleum coke gasification by calcium hydroxide and carbonate. ACS Division of Fuel Chemistry Prepr. 1, 221-225, 1996.

/4/ D. M. Ginter, G. A. Somorjai, H. Heinemann. Factors affecting the reactivity of chars and cokes during low-temperature (640 °C) steam gasification. Energy and Fuels, 7, 393-398, 1993.

151 J. A. Moulijn. F. Kapteijn. The Mechanism of Alkali Metal Catayzed Gasification of Carbon. Erdol und Kohle, 40, 1, 15-21, 1987.

161 K. J. Huttinger. The Potential of Catalysed Coal Gasification. Erdol und Kohle, 41, 9, 376-378, 1988.

I7/ M. Bistolfi. A Moving Bed Reactor Model for Catalytic Coal Gasification. Paper X-8 presented to the Joint Meeting of the French, Italian and Swedish sections of the Combustion Institute. September, 1992.

Table 1 - Feed Analysis

	Gela Residue	Hard (pentane) Asphaltene ex - Ural
Elemental Analysis (wt %, dry)		
Carbon	81.59	83.94
Hydrogen	9.91	7.82
Nitrogen	0.56	1.26
Sulfur	7.77	3.91
Metals (wt ppm.)		
Nickel	186	324
Vanadium	127	973
Conradson Carbon (wt %)	18.13	40.78
Density at 15 °C (g/ml)	0.97	1.17
Viscosity (cSt)		
at 100 °C	2,177	
at 200 °C	–	**1,171,000**
Higher Heating Value (MJ/kg)	40.7	38.3

Table 2 - Petroleum coke calalytic steam gasification
(T= 973 K P = 3.0 MPa)

Coke sample (*)	RS	GS	VE
Elemental Analysis (wt.%, dry)			
Carbon	83.42	93.58	89.16
Hydrogen	2.34	3.82	3.61
Nitrogen	1.41	1.34	1.59
Sulfur	11.99	0.74	4.92
Ash	0.84	0.52	0.72
Activation Energy (kJ/mole)			
As received	218	195	206
With catalyst	160	132	164
Reactivity (**) (g/h)/g			
As received	0.36	0.19	0.2
With catalyst	12.35	11.56	10.95
Enhancement Factor (***)	34.3	60.8	54.6

(*) from coking at T = 773 K

(**) reactivity has been calculated at 50% Carbon conversion and has been espressed as:

$$r_s = -1/m_c \cdot dm_c/dt = 1/(1-X_c)^n \cdot dX_c/dt = k$$

where:

r_s = carbon gasification rate
m_c = mass of carbon
t = time
X_c = carbon conversion
n = 2/3 (supposing the progress of the reaction at the spherical surface))
k = reaction rate constant

For the temperature dependence of the apparent rate constant there is:

$$k = k_0 \cdot e^{(-E/RT)}$$

(***) ratio between the reaction rate with catalyst and the reaction rate without catalyst.

Figure 1 - Integrated Pyrolysis catalitic Gasification Combined Cycle
(schematic)

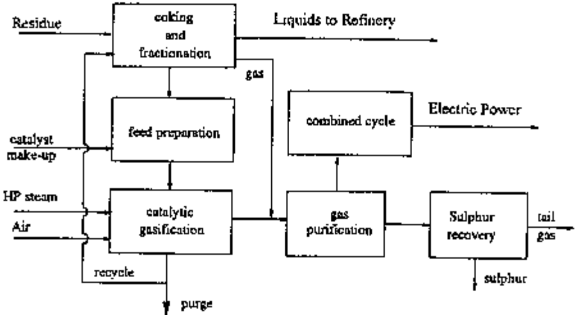
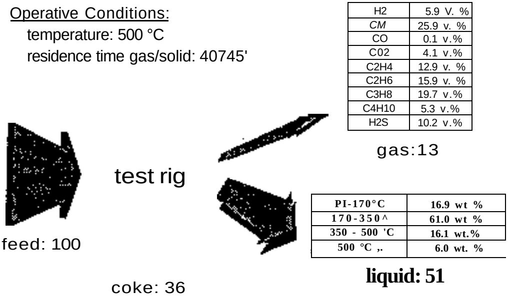


Figure 2 - Coking Products Distribution
From Gela Residue



Operative Conditions:
temperature: 500 °C
residence time gas/solid: 40745'

H2	5.9 V. %
CM	25.9 v. %
CO	0.1 v.%
C02	4.1 v.%
C2H4	12.9 v. %
C2H6	15.9 v. %
C3H8	19.7 v.%
C4H10	5.3 v.%
H2S	10.2 v.%

gas:13

test rig

PI-170°C	16.9 wt %
170-350^	61.0 wt %
350 - 500 'C	16.1 wt.%
500 °C ,.	6.0 wt. %

liquid: 51

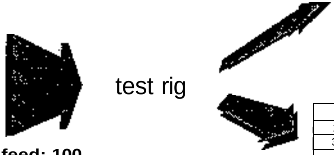
feed: 100

coke: 36

Figure 3 - Coking Products Distribution

From Hard (pentane) Asphaltene - ex Ural

Operative Conditions:
temperature: 500 °C
residence time gas/solid: 40745'



H2	28.3 v.%
CH4	35.5 v. %
CO	
C02	0.3 v. %
C2H4	11.5 v.%
C2H6	9.6 v. %
C3H8	10.2 v. %
C4H10	1.8 v.%
H2S	2.8 v. %

gas:19

P!-170°C	12.6wt%
170-350 "C	49.9 wt. %
350 - 500 °C	31.9 wt%
500 °C +	5.6 wt %

feed: 100

liquid: 33

coke:48

i%

CFB Gasification - Energy from Biomass and Waste

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1. Introduction

This paper presents an overview on the Circulating Fluidized Bed Combustion (CFBC) and the Circulating Fluidized Bed Gasification (CFBG) processes. CFBC units are state of the art and have proven their capability of converting biomass into power and/or steam. The application of CFBG in industry and power production is outlined. Process parameters of plants already in operation or under construction are shown. Decision criteria for the selection of either CFB combustion or gasification based on available feedstocks and products required are discussed.

Lurgi AG is a group of engineering contractors operating world-wide and focusing on plants for the oil, gas, chemical, metallurgical, energy and environmental sectors as well as the polymer and synthetic fibres industries. 400 processes including 200 proprietary processes for the engineering and construction of turnkey plants and plant units are being offered.

Our organisation and working methods permit each project to be realized on a flexible and individual basis. Lurgi is free to select the most suitable suppliers in terms of quality, reliability and financing including those in the client's own country or in third countries.

2. CFB Combustion and Gasification Technologies

CFB reactors are in commercial operation for reduction processes and for combustion and gasification of solid fuels. In this, paper reduction processes are not considered. The fact, that world-wide over 80 CFB combustion plants using Lurgi licences are commercially operating proves that this technology is well accepted. Lurgi's CFB gasification technology is at present applied in two industrial plants. This process is, however, the key process for our advanced thermal waste treatment processes and also well suited for biomass gasification and for the replacement of water gas fixed bed gasifiers.

2.1 CFB Combustion

The CFB combustion process utilises a fluidized bed combustor in which crushed coal or other fuels and limestone (for desulfurization, if required) are suspended in a stream of air flowing upwards due to the fine particle size of the fuel feed and the high gas velocities (approx.
bulk of the solids is carried out of the combustor with the flue gases, collected in a
one and returned to the combustor. This gives the process its name: Circulating

2.1.1 Combustion

The fuel is typically burnt at a temperature of about 850°C. It is fed directly to the combustor without requiring costly fuel preparation and distribution systems. The combustion air is introduced in two stages: Primary air through the nozzle grate at the bottom of the combustor and secondary air part way up the combustor above the fuel feed point. The limestone required for desulfurization is added near the bottom of the combustor.

2.1.2 Steam generation

Depending on the mode of operation of the CFB power plant (base load or intermediate load) as well as the plant size and the type of fuel used, the plant may be designed either with or without a fluid bed heat exchanger (FBHE). In FBHE-equipped plants, the heat transfer surfaces for steam generation are located in the combustor and in the FBHE as well. On leaving the recycling cyclone, a portion of the hot solids is diverted to the fluid bed heat exchanger where it is cooled before being returned to the combustor. The heating surface for economizer and superheat duty is typically installed in the convective pass.

2.1.3 Flue gas clean-up

The flue gases are cleaned of particulates in a downstream electrostatic precipitator or baghouse filter. Add-on flue gas desulfurization and/or NO_x removal systems are not required in CFB power plants. Gaseous pollutant control $S0_2$ capture: The fine-grained limestone required for desulfurization is fed to the process near the bottom of the combustor.

Dnsulfurization takes place directly in the combustion zone. The reaction steps are as follows:

- Oxidation of the fuel-bound sulphur
  $S + 0_2 \rightarrow S0_2$
- Calcining of the limestone to form calcium oxide:
  $CaC0_3 \rightarrow CaO + C0_2$
- , and the decisive reaction - gypsum formation
  $CaO + S0_2 + 1/2\ 0_2 \rightarrow CaS0_4$

NO_x suppression: The combustion temperature of 850°C is generally too low to allow for any significant oxidation of atmospheric nitrogen. Formation of nitrogen oxides from fuel-bound nitrogen compounds is suppressed by staged combustion air addition. This allows for low NO_x emission levels (< 200 mg/m^3_n). Chlorine and fluorine compounds are largely retained in the ash.

2.1.4 CFB ash

The only CFB by-product is dry ash which contains the original ash from the fuel, the gypsum formed as well as a small amount of free lime (CaO) and residual amounts of carbon. This ash is well suited for blending into cements and other construction materials. Due to its basicity and its hydraulic characteristics, CFB ash can be landfilled without any problems.

2.1.5 Design Features

The CFB is well-suited for power plants with capacities of 60 to 1000 MW_{th} per unit. Its excellent part-load behaviour and load following capabilities as well as its ability to start up quickly after temporary shutdowns (overnight, weekends) make it an ideal system for co-generation plants and industrial applications. The excellent heat storage capacity of the hot ash prevents the plant from cooling down excessively and hence, ensures rapid restarting. Fuel flexibility: In addition to high-grade coals, CFB firing systems are well suited to burn low-quality fuel with high ash and sulphur contents. In particular, fuels which are either difficult to burn or cannot be burnt at all in conventional steam generators may lend themselves well to CFB combustion (shale oil, carbon containing refuse e. g.). High-carbon burnout: 98-99 % carbon burnout can be achieved due to intimate gas/solids mixing and the long retention time of the fuel in the circulating fluidized bed. "In-situ" pollution control: Desulfurization is accomplished in the combustion zone itself by the addition of small quantities of limestone. At a Ca/S molar ratio of approx. 2, over 90 % of the sulphur contained in the fuel is converted to gypsum. Low NO_x emissions: Low combustion temperatures in combination with staged combustion, a typical feature of the CFB process, permit NO_x emissions to be reduced to less than 200 mg/m^3_n. Chlorine and fluorine are largely retained in the ash. Flue gas dedusting to the statutory emission limits (less than 50 mg dust/m^3,,) is accomplished in electrostatic precipitators or baghouse filters. The CFB ash lends itself for use as an aggregate for cement or other construction materials or can be landfilled without any problems. Excellent operability: part-loads down to 25 % are well within reach at load-change rates up to 7 % per minute. Simplified fuel preparation and feeding: Normally, crushers are sufficient for preparing the CFB fuel. Only when firing fuels which exhibit both high ash and moisture contents are crushing and drying required. Compact plant design: Thanks to their compact design, CFB power plants can be located in densely populated areas. CFB plants have lower space requirements than conventional steam generators with downstream flue gas cleaning equipment (Fig. 1).

2.2 CFB Gasification

The atmospheric CFB gasification [2] is suitable for feedstocks like coal, biomass or wastes. The Lurgi CFB gasifier operates at near atmospheric pressure and is therefore well suited for smaller capacities (i. e. up to around 20 t/hr of coal). The CFB gasification unit (Fig. 2) consists of a vertical, cylindrical, refractory lined vessel with recycle cyclone, bottom ash cooling, and if required, dry fly ash removal and wet gas scrubbing systems. The CFB gasifier operates in a mode between the classical bubbling bed and the pneumatic transport reactor. Under those conditions the slip velocity between solids and gas (or the velocity differential) is highest, leading to maximum heat and mass transfer between gas and solids, requiring the smallest reactor diameter of all fluidized bed principles. Coal, biomass, wastes or other solid fuels are introduced into the reactor near its bottom. Gasification agents - depending on product gas specification - air, oxygen and steam, or oxygen and carbon dioxide are introduced through a nozzle grate in the lower part of the reactor. Ash is partly withdrawn through the reactor's grate (bottom ash) and partly recovered from the product gas (fly ash). Gasification reactions are starting close to the bottom of the reactor at the fuel feeding point. Reaction temperature typically ranges from 800-1050°C, depending on the type of feedstock. The dust laden product gas leaves the reactor at its top and passes through a cyclone. The major portion of the dust is removed from the gas and recycled to the gasifier bottom through a stand pipe with seal pot, leading to high carbon conversion. The product gas is then cooled, dedusted and purified depending on the requirements of its further use. Commercial gasification plants are in operation or under construction with capacities of up to 100 MW_{th}. The HTW (High Temperature Winkler-Gasifier) (Fig. 3), a development of Rheinbraun/Germany, operates as a bubbling fluidized bed at pressures between 1 and around 2.5 MPa [3]. Thus the HTW gasifier lends itself to larger capacities of up to around 700 MW_{th} in one reactor. It is being engineered and marketed by Lurgi in co-operation with Uhde Co., Germany. The HTW gasification system consists of a vertical refractory lined, cylindrical vessel with recycle cyclone, a coal feeding system - screw or gravity depending on feedstock - into the pressurised gasifier, and an ash cooling and removal system. The gasification agent (a mixture of air or oxygen and steam, depending on the use of the product gas) is introduced at different levels into the fluidized bed and into the freebord for further gasification and decomposition of hydrocarbons in the gas. The major portion of the entrained dust is removed from the hot crude gas -typically around 850-1000°C - by a cyclone and recycled to the fluidized bed via a down pipe. The hot crude gas is then cooled and treated to the conditions required for further use. The operation of the gasifier at elevated pressure leads to a high gasifier capacity of about 100 MW_{th}/m^2 of gasifier cross-sectional area when using oxygen, and 50 MW_{th}/m2 when operated in air-blown mode (at 2.5 MPa pressure). Commercial scale plants in operation include Rheinbraun's demonstration plant in Germany, where methanol synthesis gas is produced from lignite (capacity approx. 730 tpc dried lignite). Decisive for the application of either atmospheric or pressurised gasification is the required thermal capacity of the plant and the availability of biomass as feedstock. The typical plant capacity for ACFB-gasification is in the range of 10 MW_{th} to 100 MW_{th}. For larger plant capacities the pressurised HTW gasifier appears to be more economical.

A wide variety of biomass feedstocks like

woodchips, treebark, forest wastes

miscanthus, straw, other farmed biomass

biomass wastes (bagasse etc.)

municipal waste, sewage sludge, paper sludge, plastic refuse, etc.

is suitable for CFB gasification.

As an alternative to combustion plants, which produce steam for power production or CHP only, the gasification plants convert the fuel into a product gas for a variety of uses:

fuel gas used for supplemental firing of existing power plants, kilns etc.

synthesis gas for chemical processes (methanol etc.)

fuel gas for combined cycle power generation

3. Application of CFB Process

3.1 CFB-Combustion

Since 1985 when the first CFB power station started commercial operation, more than 80 CFB power stations have been ordered. The capacity ranges from 20 to 250 MWe [4].

Various feedstocks are utilised depending on the plant's location:

hard coal, lignite, anthracite culm, paper sludge, petcoke, oil shale, RDF and wood

Most of the plants are located in the USA, Germany, Italy, France, Japan, South Korea, China, India and Slovakia. Of these plants there are few which utilise wood only.

•

	Capacity	**Steam**	**Start up**	
Fresno, CA, USA	89 MW_{th}	100 t/h 87 bar 515°C	1988	wood
Rocklin, CA, USA	89 MW_{th}	100 t/h 87 bar 515°C	1989	wood
Mecca I+II, CA, USA	2x79 MW_{th}	103 t/h 89 bar 496°C	1992	waste wood

The successful operation of these biomass burning plants proved the ability of CFB combustion technology to utilise unconventional feedstocks [5].

3.2 CFB Gasification

CFB gasification can be efficiently used as a front end process in the following applications: [6], [7]. [8]

- *Biomass to Electric Power*

CFB gasification attached to Power Plant Boiler:
Gas from biomass is used as substitute fuel in existing coal or heavy oil fired power plants.

CFB gas for Combined Power Cycle:
CFB gas from biomass is cleaned and conditioned for combined cycle power generation (Fig. 4).

- *Waste Fuels for Cement Kiln Firing*

CFB gas from waste is used as substitute fuel for precalciners and/or main burners (partial substitution) of cement kilns.

- *Biomass for Lime Kiln Firing*

CFB gas from tree bark / wood waste / paper sludge is used for lime kiln firing for instance in pulp mills.

- *Waste to Electric Power*

CFB gas from municipal waste or RDF (Refuse Derived Fuel) or sewage sludge is purified and used as substitute fuel in gas fired power plant boilers e.g.

- *Biomass to Synthesis Gas*

Gas from biomass is purified and converted to meet the gas specifications required for the downstream synthesis units.

- *Biomass to Methanol and Electric Power*

CFB gas from biomass is used for methanol production.

The purge gas from the methanol plant is burnt in a gas turbine for power generation.

3.3 Status of Technology

The Lurgi CFB gasification pilot plant has been operated since 1983 for more than 5000 hours. The gasifier has a thermal capacity of 1.7 MW_{th}. During the test periods various feedstocks such as hard coal, lignite, biomass, petcoke and all kind of waste material have been tested successfully. The CFB gasification technology is commercially available since 1985.

Location	Capacity	Product	Start up	Fuel
Pols, Austria	27MW_{th}	Fuel gas	1987	tree bark
Rudersdorf, Germany	100 MW_{th}	Fuel gas	1996	wood, waste wood RDF, lignite waste, rubber waste
Project 1	12 MWel	El. Power	1998	short rotation forestry product
Project 2	20 MW_{TH}	Fuel gas	1999	municipal waste
Project 3	30 MWel	El. Power	1999	wood, waste

4. Experience with CFB Gasification

The operation of the CFB gasifier at Zellstoffwerke Pols AG has proven that gas produced from tree bark in a CFB is suited to bum quick lime in a rotary kiln [9].

In this plants the bark drying process as well as the CFB gasification process was optimized with respect to gas quality. According to the fact that at Pols the lime was cycled in a closed loop the addition of the produced gas dust proved, however, to be harmful in the long run.

The Rudersdorf (100 MW_{th})CFB plant came on line in September 1996 and achieved an availability of approx. 80 % by December 96. The plant provides reliably gas for the precalcining. The ash from the CFB gasifier is used as an additive to the raw meal of the cement process.

5. Conclusion

Lurgi CFB combustion and CFB gasification technologies are commercially available for a wide variety of feedstocks. Both technologies have proven their reliability. Results from continued operation concluded that use of CFB technology to burn or gasify biomass achieves high conversion efficiencies and the required low emission levels. The process of choice has to be selected case by case and depends on the client's specific requirements.
With regard to the combustion, gasification, gas clean-up and synthesis processes Lurgi can offer, it is prepared to maximize local project content (detail engineering, manufacturing of much of the equipment e. g.) - this means that forex outlays are significantly reduced. In addition Lurgi's willingness to cooperate to the maximum extent with the relevant local organisations (Design Institutes, manufactures etc.) will ensure that biomass or other solid feedstocks can be converted to high-value products in an economical and environmental - friendly manner.

References

[1] Lurgi Publication, CFB Power Plants, 198 e/8.92/2.50

[2] Loffler, J. C, et al., Gas from Biomass and Waste for Industrial Use, Conference on Thermal Power Generation and The Environment, Hamburg 1-3 September 1993

[3] Herbert, P. K. et al., The Rheinbraun HTW Gasification Process as an Integral Part of Combined Cycle Power Stations. An Update of Development, 12th Conference on Coal Gasification Power Plants, San Francisco, CA, October 27-29,1993

[4] Cleve, K. et al., Large CFB Power Plant Design and Operating Experience - Texas - New Mexico Power, 13th International ASME Fluid Bed Conference Orlando, Florida, May 1995

[5] Gershengoren, E. et al., Design and Operation of a Wood Fired CFB Steam Generator, Paper presented by Ultrasystems Engineers and Constructors, Inc. Irvine, California

[6] Hirschfelder, J. et al., Gaserzeugung aus Biomasse und Restabfallen in der zirkulierenden Wirbelschicht, Chemie Ingenieur Technik 9/95

[7] Loffler, J. C. et al., Biomassenverstromung durch Vergasung und integrierte Gasturbinenprozesse, VGB Kraftwerkstechnik 74 (1994), Heft 7

[8] Greil, C. et al., Circulating Fluidized Bed Technologies for the Conversion of Biomass into Energy, Paper presented at UNIDO Symposium, Vienna, Austria

[9] Loffler, J. C. et al., Low BTU Gas from Tree Bark for Firing a Rotary Lime Kiln in a Sulphate Pulp Mill

Abbreviations

CFBC	Circulating Fluidized Bed Combustor
CFBG	Circulating Fluidized Bed Gasifier
FBHE	Fluid Bed Heat Exchanger
IGCC	Integrated Gasification Combined Cycle
LHV	Lower Heating Value
ACFB	Atmospheric Circulating Fluidized Bed
MW„	Megawatt electric
MW_{th}	Megawatt thermal
CHP	Combined Heat and Power

Basic Flow Sheet of a CFB Boiler

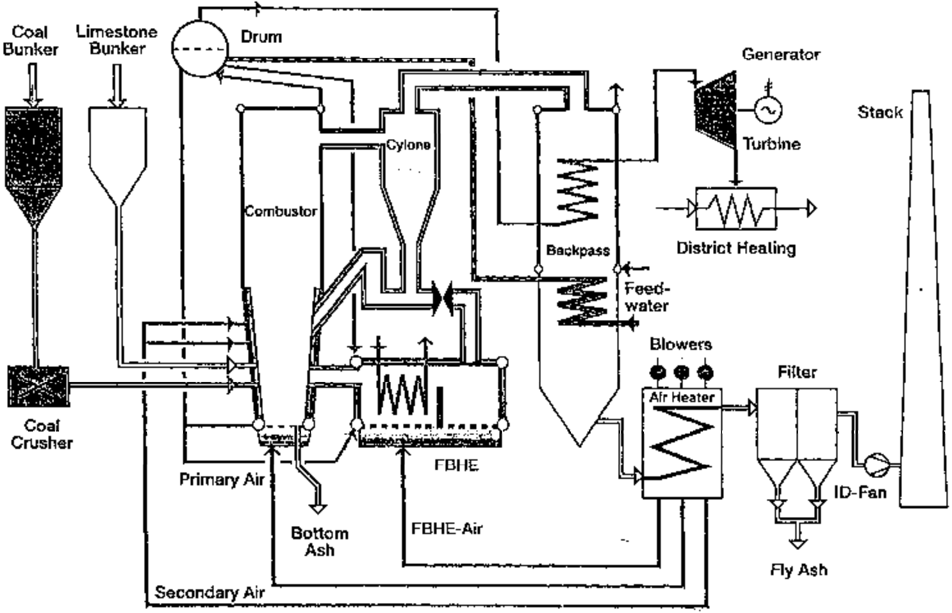

Fig- 1

CFB Gasification

LURGI

Gasifier Unit

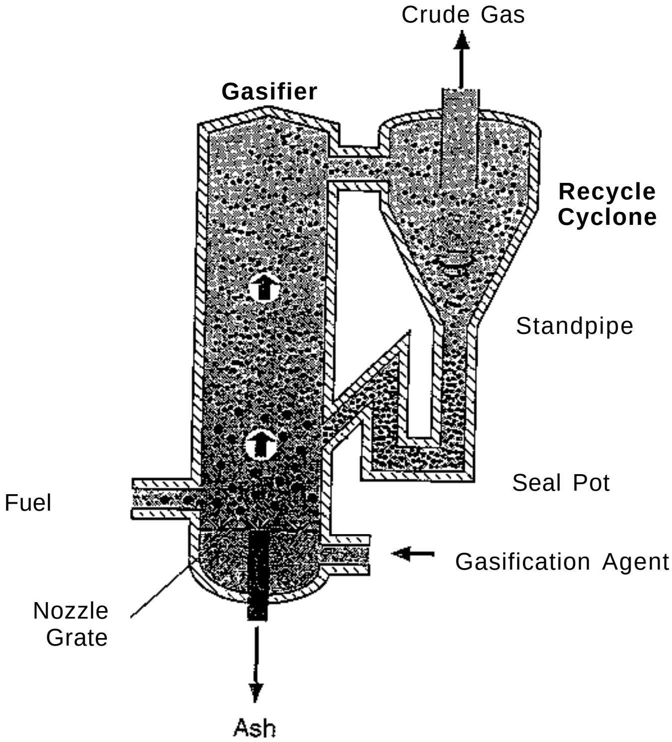

Fig. 2

High-Temperature Winkler (HTW) Gasification System

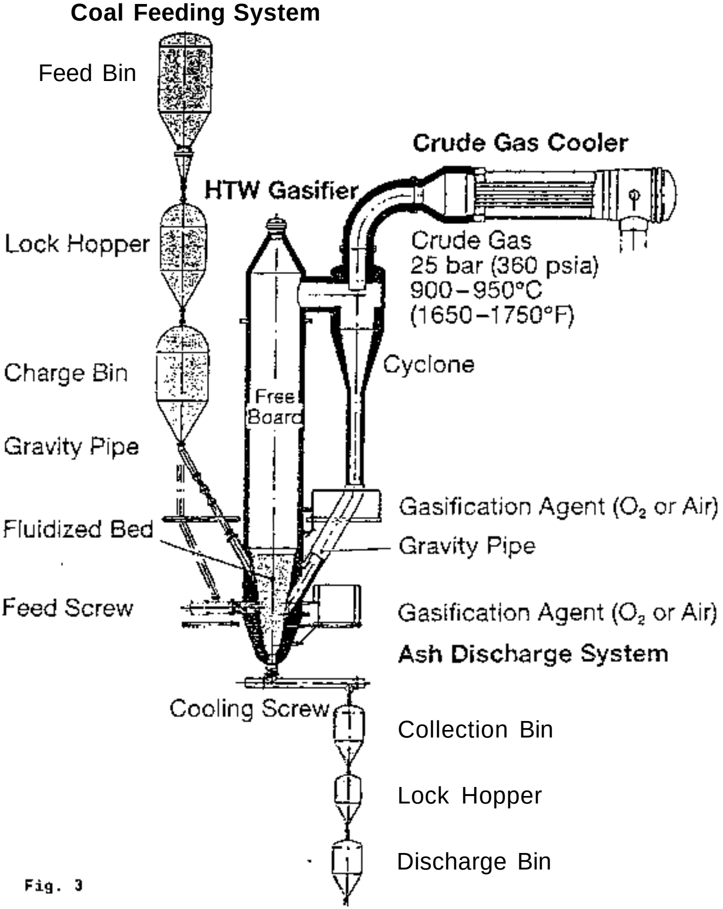

Fig. 3

Overall Process Scheme

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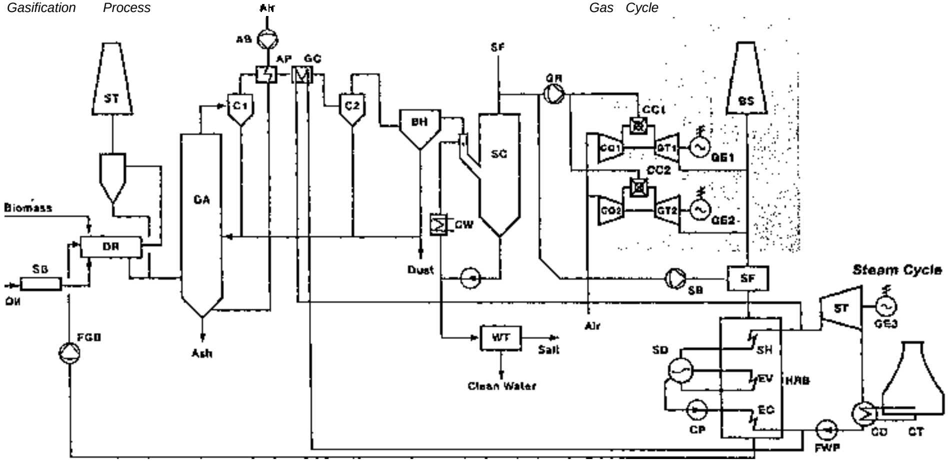
Power from Biomass
Atmospheric Gasification Combined-Cycle



Gasification Process - Lurgi

GA	Gasifier	SF	Start-up Flare
C1.C2	Cyclone 1.2	GR	Syngas Recompressor
AB	Air Blower	FGB	Flue Gas Blower
AP	Air Preheater	SB	Start-up Burner
WT	Water Treatment	DR	
BH	Bag House Filter	ST	Stack
SC	Scrubber	CW	Cooling Water
GC	Gas Cooler		

Gas Cycle

C01.C02	Compressor 1,2
CC1.CC2	Combustion Chamber 1, 2
GT1.GT2	Gas Turbine 1, 2
GE1.GE2	Generator 1,2
BS	Bypass Stack

Steam Cycle

ST	Steam Turbine	FWP	Feed Water Pump
GE3	Generator 3	SF	Supplementary Firing
CD	Condenser	EC	Economizer
CT	Cooling Tower	SH	Superheater
HRB	Heat Recovery Boiler	EV	Evaporator
SD	Steam Drum	CP	Circulating Pump
SB	Syngas Blower		

Fig. 4

THE ROLE OF GASIFICATION IN ENERGY PRODUCTION FROM MUNICIPAL SOLID WASTE

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PURPOSE

Gasification is emerging as an alternative to combustion in the treatment and energy recovery from Municipal Solid Waste. Several innovative proccesses and demostration plants such as Thennoselect in Fondoconte, TPS in Greve, Lurgi and others, are trying to achieve higher electrical efficiencies and lower emissions using this technology.

Most of the proccesses at work today generate a hot gas with a temperature between 900 and 1300 °C wich is cooled and treated later to avoid damage on the engines, turbines or steam cycles were power is produced. Cooling is necessary because the hot gas is highly corrosive due to the temperature and acid gas and other corrosive compounds concentration, wich can only be efficiently removed from the gas when temperatures are low.

In this paper we will deal with the composition of the gas obtained from high temperature gasification of Municipal Solid Waste or Refuse Derived Fuel, its Calorific Value, the amount of energy lost in the cooling of the gas, and so on. The process is analyzed from the energetic point of view and compared with demonstrated technologies, such as mass bum and Refuse Derived Fuel incineration.

METHODOLOGY

After a 3 year program of study, the elemental analysis, Calorific Value and other properties of MSW in Galicia is quite well known. In Table 1 we have the average annual elemental analysis, in Kg per Tn of MSW.

Table 1. Composition of MSW in Kg per Tn that goes into the Management System

C	H	O	N	Cl	S	Ash	Inerts	Moisture
249,9	39,2	163,1	7,6	7,8	2,7	88,5	138,6	302,7

Steps are taken to recover energy from the waste and at the same time safely solve the problem of waste disposal. That so, proven technologies, mass bum grate firing systems and fluidized bed RDF incineration, were first considered but also attention was driven to MSW gasification experience. Both expected emissions and the amount of recovered energy are the main issues when comparing the different technologies and simulation may help in this purpose.

The first scheme considered was grate firing of the raw waste. This kind of plants normally operate with excess air levels between 80 and 110 % to optimize emissions and keep temperature levels in the combustion chamber within an acceptable range. Normally the combustion chamber is not cooled down except in water wall systems. For simulation purposes we then considered the furnace as an adiabatic system. From the composition of the waste we calculated the amount of air necessary to achieve the required 80-110 % excess air level in the furnace. With an algorythm for gibbs free energy minimization in an adiabatic reactor, exit temperature and gas composition is calculated. Equilibrium is a safe assumption when gas temperatures are over 700 °C, as it is the case. Gas composition includes contaminants wich show dependance on the waste composition but also on the temperature and excess air.

To perform calculations both for combustion and later for gasification the program CHEMCAD III was used. More than 40 chemical especies were considered but many were not present in the equilibrium. Carbon, CO, C02, CH4, H2, 02, N2, H20, HC1, C12, S, H2S, S02, *SOS,* N2, NO. N02, N20, N03, COS, NH3, HCN were present depending on the particular conditions.

Continuing with the study of mass burn, the gas from the combustion chamber is then cooled to 220 C, typical temperature for gases exiting the boilers ot this type of plants. The difference in enthalpy between this gases and those leaving the furnace would be the maximun thermal energy recovered through steam generation.

The second scheme considered is RDF preparation followed by Atmosferic Fluidized Bed combustion In the preparation of RDF, several inert materials such as glass and metals are separated. This does not affect the chemical properties of the waste but allows avoiding damage on the bed and boiler, and also on the mills necessary to reduce the particle size of the RDF feed to the bed within certain range.

In the RDF preparation plant the incoming Waste is also dried to a moisture content in the final product of 10%. The thermal energy for the drying stage is provided by hot flue gases from diesel engines. Drying takes place in a cogeneration mode with power production in the engines and thermal energy reclaimed from the gases to dry the waste. This stage provides additional energy output to the treatment system but was not computed in the calculation as this additional energy comes from diesel fuel and not from waste.

Combustion of the RDF takes place in a Circulating Fluidized Bed. This kind of boiler typically operates holding temperature at a constant level. In MSW combustion, temperature in the furnace must be held at 850 °C or more for more than 2 seconds. In the simulation the operation temperature of the bed was settled at 870 °C. The reactor operates in an isothermal mode with the typical excess air levels (30 to 60 %) encounter in this type of MSW incineration. Once again Gibbs Free Energy minimization provided us gas composition. To keep the 870 °C temperature with a lower excess air than previously, it is necessary to remove heat from the reactor. The amount of energy removed for each excess air level is calculated.. Gas is later cooled to 180 ^{o}C and the amount of Energy removed in this operation is also calculated.

The gasification scheme was necessarily more complicated. It includes the gasificator itself, a gas cooling and cleaning system and a combustor. The gasificator operates at a pressure of and adiabatically. The oxydazing agent is air supplied at ambient temperature (25 °C). The amount of air is variable. To optimize the reactor the curves of temperature, gas composition and HHV of the gas were plotted against the amount of air injected. It is important to avoid char formation in the reactor and to obtain a gas with a high HHV. The optium air supply is the one that avoiding char formation gives the maximun in the amount (gas HHV X gas Volume) as this amount represents the chemical energy stored in the gas.

Both MSW and RDF were initially considered for gasification. Nevertheless the moisture content in MSW is higher, as comented previously, and this leeds to a lower temperature inside the reactor for the same amount of air supply. Then, a higher amount of air injection is needed to avoid char formation as 900 ^{o} C of temperature are necessary. Due to this higher amount of air supply and to the higher moisture content, the gas resulting from MSW gasification is worse than the obtained from RDF. That's why in the following we will refer only to RDF gasification not presenting the results for MSW.

The gas leaving the reactor has a certain amount of thermal energy as a results of its high temperature. Part of this energy is recovered by a heat exchanger placed at the exit of the reactor cooling down the gas to around 600 ^{o}C. The heat recovered might be used in low temperature steam production of water preheating, depending on tube temperature, wich should be low as corroding agents have a concentration higher 2 to 3 times higher than in a normal combustion gas, and so the temperature of the heat exchanger tubes is an important parameter that shoul be kept low to decrease corrosion rate. Anyway the main purpose of this stage is to give an idea of energy that might be recovered or lost with or without this intermediate heat recovery. The purpose is not to define the steam cycle of the plant

After this preliminary cooling, the gas is further cooled through direct water injection in a scrubber. This also cleans the gas. Acid gases or its precursors are removed. In the simulation no reagent was added to water but almost all the HC1 (98 %) was removed. For SH_2 this figure was much lower but with an appropiate reagent will increase. The temperature of the gas leaving the scrubber is of 50 ^{o} C.

The low temperature of the gas also condense volatile compounds such as vaporized chlorides and dioxines. Dioxines, that may be formed during gasification or already present in the RDF, are compounds of a high molecular weight and all tend to condense at 300 ^{o}C. So 50 ^{o}C is a quite suitable temperature for gas cleaning. A negligible amount of HCN, COS and NH_3 remains in the gas, that will later be destroyed in the subsequent combustion. A flowsheet of the plant is displayed in figure 1.

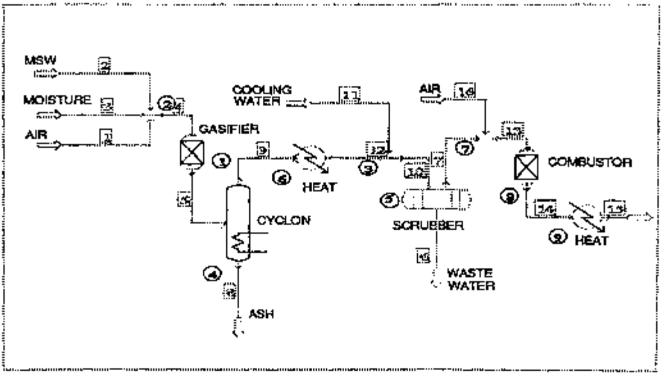


Fig. 1

RESULTS AND CONCLUSIONS

The composition of the gas resulting from combustion for both of the options (mass burn and RDF combustion) varies depending on air supply. Optimun values are 60 % excess air supply for RDF and 110 % for mass burn as contaminants are minimized. A complete mass flow for either case is displayed both in figures 2 and 3.

To compare emmissions on the same basis, calculations were computed for one Ton of MSW that goes into the system. Despite emmissions show strong dependance on waste composition it is clear that mass burn provides worse conditions for waste burning. Also as it operates with higher amounts of excess air it requires bigger cleaning systems and the concentrations of contaminants in the exit gases will be lower, affecting the efficiency of pollutants removal.

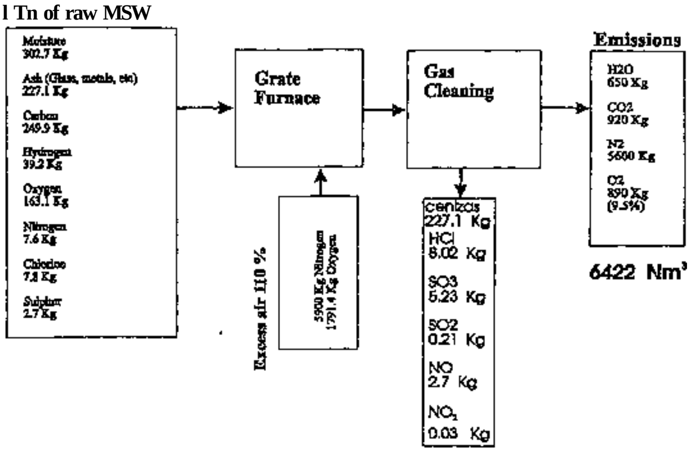


Figure 2

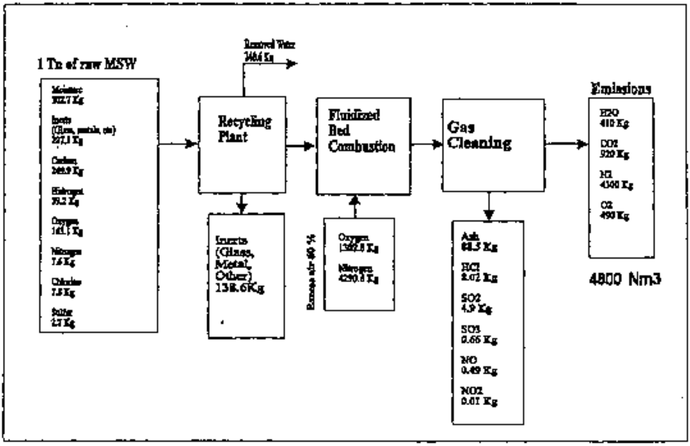


Figure 3

In our gasification scheme pollutants are removed in an intermediate stage after the gasification reactor and previous to the combustion of the gas. The variation in the composition of the gas as a function of the percentage of air injected is shown in figure 4.

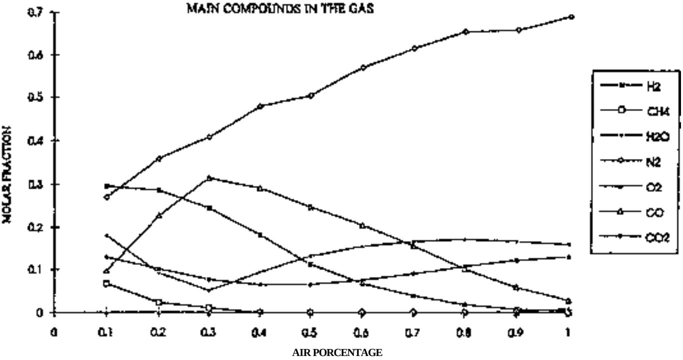


Figure 4

There is also a remaining amount of Carbon, minimized to less than 2 % when 40 % of the stoichometric air or more is reacts with the RDF. In that point the temperature of the adiabatic system is of 956 °C. The amount and HHV of the gas is also affected by the amount of oxydant (Figure 5).

HHV & AMOUNT OF GAS

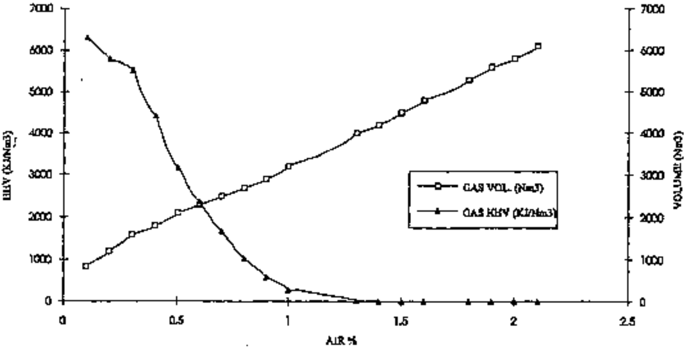


Figure 5

It is anyway clear that operating in the conditions mentioned above apart from minimizing the amount of Carbon residue is possible to obtain a gas with a Calorific Value higher than 4000 KJ/Nm3.

The gas already obtained carries contaminants in the following amounts per Ton of MSW processed:

Table 2. Contaminants in the Gas

HCl	H_2S	HCN	COS	NH_3
8000 gr	2800 gr	0,178 gr	231 gr	12,4 gr

After the cooling and cleaning of the gas the amounts of this compounds change and are now:

Table 3. Contaminants after cleaning with water

HCl	H_2S	HCN	COS	NH_3
174 gr	889 gr	0,173 gr	228 gr	0,2 gr

The percentage for HC1 removal is 98 % while the remaining HCN, COS and NH_3 are destroyed in the combustion of the gas. The only problem is then, H_2S removal, but this is due to the low solubility of this compound in water and with an appropiate reagent would be solved

The removal of this compounds, contaminants and at the same time corrosive have other adventages. It is possible for example to cool the combustion gases to a lower temperature than in normal combustion without damage on the tubes of water preheaters. Anyhow to compare the amount of recoverable Energy we assumed gas cooling to 180 °C as in the case of Fluidized Bed Combustion. The comparison of this parameter for the three cases is shown in figure 6.

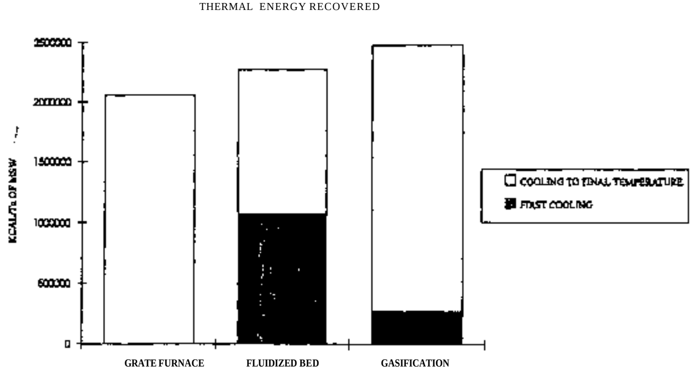


Figure 6

In the figure the first cooling represents the energy removed from the bed in FBC and from the gas between 956 ^{o}C and 600 in gasification. Figures are in terms of KCal/Tn-MSW for a proper comparison and it is clear the advantage of gasification. It is interesting that the amount of energy recovered in the first cooling in gasification is low, 15-18 %, of the total.

To wash the gas the amount of waster needed is of 500 liters per Tn of MSW.

Advanced Continuous In-Line Gasification and Vitrification of Solid Waste -

A New Technology in Practice

R.Stahlberg, B.Calaminus, C.Spadacini, F.Steiger

Introduction

In Europe during the late 1980's and through the early 1990's, the public opinion became very concerned about the problem of dealing properly with the amounts of waste every person produces throughout the year. Landfilling has been identified as an inappropriate, not sustainable solution especially in longer terms with regard to the heritage of future generations. Avoiding and recycling of waste helps reducing the quantity of some specific waste fractions like for example glass and paper; but despite these efforts the remaining overall production of waste per inhabitant in the range of 200 to 400kg/year claims for other alternatives. As mass burn - a technique used to reduce the volume and the weight of the waste - has been found as ecologically not satisfying, emission regulations on municipal waste incinerators were tightened drastically in Central Europe. In order to meet the emission limit values, very big investment is required. The solid combustion residues (about 30% of the input) still represent a burden making post-treatment and special landfilling necessary.

In this situation an innovative thermal waste treatment process was developed combining well known and proved technical items to a closed, continuously working system transforming waste completely into directly usable products. The process steps comprise compaction, degasification, high temperature gasification and melting. The use of oxygen instead of air for the gasification process allows to reach very high temperatures in short time. So, equilibrium is attained in the gas phase as well as in the molten phases before leaving the reaction chambers. By using rapid cooling at both - gaseous and molten phase - system outlets, homogeneous phases are obtained corresponding to high temperature equilibrium states. "De-novo"-synthesis of organic compounds (like PCDD/F) and formation of nitrogen oxides is avoided, production of ash and filter ash is excluded. From one ton of waste and about 500 kg of oxygen the following stable end-products are obtained from the THERMOSELECT process without any further treatment:

- 890 kg synthesis gas [energy carrier for conversion into electricity; chemical synthesis]
- 350 kg water [for cooling purposes instead of drinking water]
- 230 kg vitrified minerals [various industrial application like road building, sand blasting]
- 29 kg ferrous metal alloy granulate [re-introduction into the iron industry]
- 10 kg salt [sodium chloride; industrial quality]
- 3 kg non ferrous metal concentrate [zinc and lead]
- 2 kg sulphur [industrial quality]

The following paper describes the process with special emphasis on the degasification, gasification and melting part. Experiences obtained are presented.

1. The THERMOSELECT Process

1.1 Compression

Municipal, industrial and other kinds of waste are left untreated and compacted to about one fifth of their original volume by means of an armoured hydraulic press, and then pushed into an indirectly heated degasification channel (wall temperature about 600°C). The high degree of compaction greatly densifies the waste and reduces the residual air content, the air has no insulating effect (thus significantly increased heat transfer into the plug is achieved), nitrogen ballast does not need to be heated and subsequently cleaned, and the compressed plugs relax sealing the channel against the press. During compression, liquids flow into the remaining cavities, and bulky material is crushed.

The press is equipped with state-of-the-art wear-resistant materials. Heat transfer from the coupled degasification channel is prevented by a temperature-stabilised element. A mechanical lock (hydraulically operated compression unit) and a slightly conical expansion in the cross section of the channel prevents trash packets from sliding back.

1.2 Degasification - Gasification - Melting

Technical conditions for the closed loop process

As the waste plugs are pushed down the channel in an air-free environment, water is evaporated and the organic components in the refuse are partially degasified and converted into a coal-like product as the temperature increases. The walls of the degasification channel are tempered to over 600°C by the circulation of hot gas from the combustion of synthesis gas. After about 2 hours dwelling time, the waste packets are conditioned for the subsequent processes in the high-temperature reactor (HTR) [1-6]. Long years of operation and analytical studies have shown that dense layers of carbonised material protect the steel walls of the channel safely from penetrating corrosion. Secondly, a sharply reduced friction coefficient allows the compressed waste to slide smoothly through the degasification channel. Mechanically damaged layers are cured in the short run by materials within the contents of the advancing trash. The steel materials selected and the type of construction result in service lives of over 10 years [7,8]. In addition, the heating channel system acts as a second sealing level, in which any potential degasification products would be burned up. This would be monitored by the operation monitoring system analysing the gaseous composition in the channel heating combustion chamber.

The fact that the degasification channel is permanently filled with solid waste and degasification products is another safety relevant characteristic. In contrast to rotary kilns for example, no larger gas volumes can be created. A purely hypothetical entry of surrounding air and the formation of combustible mixtures is thus made impossible not only by the sealing of the system that was mentioned, but also by the fact that there is only approximately 300 hPa (0.3 bar) of excess pressure in the system.

To allow the degasification products to be carried smoothly out of the channel into the high-temperature area, regardless of the composition of the waste, a system of internal gas-diversion holes and a conical enlargement of the cross section of the channel of approximately 30% were designed.

In summary, the basic difference from interrupted processes, which for example use rotary kilns as pyrolysis reactors, is the direct connection of the high-temperature reactor to the degasification channel, which rules out risky, expensive handling of tar-like products and degasification residues containing hazardous substances.

In the transitional area between the channel and the high-temperature reactor (HTR), the advancing, partially degasified material forms an acclivity, which is heated even more before it slides onto the pile in the lower part of the HTR by convection and radiation of the hot reaction gases from the high temperature reactor.

Temperatures reached are below the softening temperatures of the inorganic portions - for example glass - and prevent it from sticking to the inside surfaces of the channel.

Temperature measurements and changes detected by metallography in the texture of the channel material 13CrMo4.4 (hardening texture with heating features up to over 750°C, [7]), lead to the design and installation of a cooling system in this zone.

The high-temperature reactor (HTR) is the crucial reaction chamber of the Thermoselect process allowing the thermal conversion of the organic portion of the waste into synthesis gas (> 1200^{o}C) using pure oxygen and water vapour and of the inorganic components into liquid melt phases (> 2000°C).

The degasification products slide into the lower third of the HTR from the degasification channel, which is connected directly to it.

The pressure-resistant steel construction of the HTR, which is also secured against excess pressure by a water lock, has inside wall temperatures clearly over 100°C when in operation, so that dew-point corrosion cannot occur. The high heat capacity of the refractory is sufficient compensation for large short-term changes in the heat as a result of the fluctuating composition of the waste. Moreover, the refractory protects the outer metal sheathing from the direct effect of high temperatures and provides protection from attack by corrosive media. The refractory is state-of-the-art and has temperature sensors, which are staggered in the areas with the heaviest stress and can indicate damage in the initial stages.

The refractory is made stable and durable by choosing appropriate corundum-based materials and by efficient cooling. Moreover, the modular design and the division of the reactor into segments makes it faster to change individual refractory parts after the designed lifetime. Changing the bottom of the HTR takes roughly 3 days, so that its overall availability is guaranteed to be high.

Inorganic material conversion and homogenisation of the mineral components is effected in a second chamber, called homogenisation reactor directly connected to the HTR. Oxygen and combustible gas are added to reduce the carbon residue level and to maintain the temperature of the melt above 1600°C.

Chemical and physical conditions

Concerning the heterogeneous composition of waste the recovery of reusable materials and energy needs high temperatures [1,2,4] to destroy organic compounds completely and to generate not only the smallest inorganic molecules like hydrogen, carbon monoxide and water vapour as main products of the synthesised gas, but also molten minerals (specific weight > 2.5 g/cm^3) and metals as stabilised phases (specific weight > 7 g/cm^3). The most important chemical reactions are shown in figure 1.

Main Reactions **THERMOSELECT**

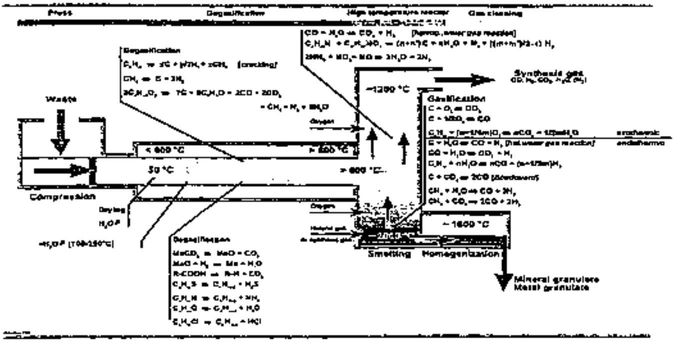


Figure 1 ("⇔" indicates important equilibrium reactions, "⇒"indicates reactions being almost monodirectional)

The residence time for the gas phase and also the molten phases are designed to be sufficient for a reproducible quality of all generated products. The high temperature phases of the gas as well as of the molten material will be frozen" by shock cooling with water.

The shock cooling of the raw synthesis gas prevents the de-novo-synthesis of dioxins, furans and other organic compounds. The gas, having a highly reduced volume per ton of waste due to the absence of nitrogen ballast, can be used after cleaning for heating purposes, to generate electrical and thermal energy, or the chemical synthesis (e.g. methanol). The sensible heat of the quench water can be used for district heating applications.

The vitrified mineral aggregate possesses the quality of natural raw materials suitable for the full range of standard applications. The metal fraction form ferrous alloy pellets which are separated by magnetism for reuse in the smelting industry.

Due to the fact that the amount of material in the closed chemical system is always more than 15 Mg, the residence time of the solids is 1 to 2 hours, of the gas phase 2 to 4 seconds and of the molten phases more than 5 minutes. This is valid at every moment the homogenisation and the stabilisation of the equilibrium take place.

This supposition is also given, if monofractions are introduced with the waste, e.g. large quantities of volatile liquids or portions of PVC. Spontaneous processes are delayed, no abrupt fluctuations in pressure occur. This is another advantage of the Thermoselect process.

Thermodynamic calculations and practical results

The theoretical calculation of chemical equilibrium was developed from the analytical results of the synthesis gas composition using a computer program to address the multitude of components [4,5,6].

The theoretical basis of such a computer program is the chemical equilibrium condition $\mu_i \nu_i$ = 0, where temperature and pressure are constant; (dT=0, dP=0), μ_i is the chemical potential of the component i and vi is the stoichiometric coefficient of the component i for the gross reaction equations. The number of gross reactions is determined by the number of chemical elements and compounds to be considered. As an example, the Boudouard reaction or the water gas equilibrium alone yields only partial information and does not replace the need for detailed thermodynamic calculations.

Figure 2 shows the results of the equilibrium calculations at different temperatures and the resulting purified synthesis gas composition assuming constant oxygen and natural gas consumption. As the figure shows, when in thermodynamic equilibrium, the quantity of CO decreases and that of H_2 and CO_2 increase as the temperature decreases. The quantity of H_2O in the raw synthesis gas at the HTR outlet just prior to shock cooling also decreases (not shown here).

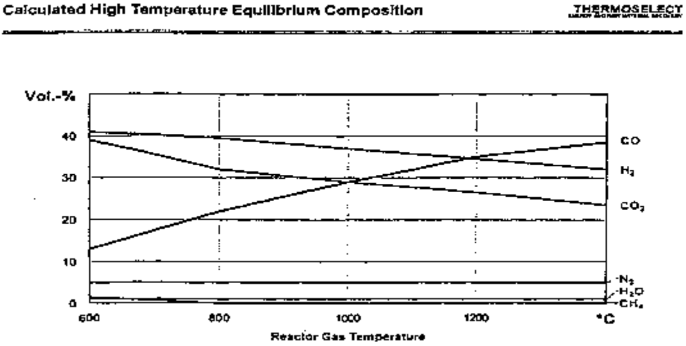


Figure 2

The natural gas component, in the raw synthesis gas and in the purified synthesis gas, rises above 0.0001% in the equilibrium calculations only at temperatures below 800°C. In practice values less than 0.1% were measured.

This essential difference, relative to a pyrolysis process which excludes air, is the result of the addition of gasification using oxygen in the overall process and process-control capability that ensures that the developing gasses receive high temperature exposure with adequate residence time in the upper section of the HTR.' Thus, any concern for the presence of higher natural gas (CH4) content in the synthesis gas when the gas exit temperature exceeds 800°C lacks any material basis.

The sulphur and nitrogen components of the waste were also included in the thermodynamic calculations. The calculated volume quantities for the components: HCN, COS, CS_2 and NH_3, lie far below 0.0001% in the temperature ranges under consideration. These results are in accordance with the experimental data obtained from the plant operation.

The comparison of the average values of the synthesis gas compositions sampled and measured by RWTUV against the calculated values shows good agreement, particularly when one compares the equilibrium composition calculated for T= 1400°C, which demonstrates that sufficiently high temperature and assured equilibrium conditions are reliably present within the HTR. Even the trace quantities of HCN, COS, CS_2 or NH_3, lying within the range of the detectable limits, confirm the theoretical results [5, 9].

Theoretical equilibrium calculations of the generated synthesis gas composition have been compared to values measured over several measurement periods. The agreement of the closest theoretical values and the assumed equilibrium temperature of the synthesis gas, 1400°C when compared to the actual measured synthesis gas exiting temperature of 1200°C, suggests a higher reaction temperature in the upper chamber section of the HTR. Measurements of the synthesis gas composition taken over a considerable time interval differ only slightly from each other and are due to varying waste compositions. Important to note is the very slight trace quantity of natural gas found in the synthesis gas and the absence of oxygen, which are both key indices for the confirmation of equilibrium in the gaseous phase. This further confirms the effectiveness of the process and validates the in-line layout of the thermal line.

1.3 Gas Cleaning

As shown on figure 3 as a first step, direct shock cooling of the crude synthesis gas from roughly 1200°C to below 80°C, at a ratio of gas to water flow of roughly 1:25, results in fast cooling so that organic compounds are prevented from re-forming. At the same time, particles of mineral material are deposited leaving the high-temperature reactor as small droplets of liquid melt in the crude gas flow. In addition, the gas quenching protects the equipment carrying the gas from excess thermal stress and causes initial effective particle deposition. Traces of carbon particles that get washed out with the quench water leave the closed water loop via two sinks. First, when the water gets in contact with the hot crude gas (approx. 1200°C), the quench acts like sort of "hot steam reactor," in which organic trace components are disintegrated. Second, constant sludge removal from the quench water also prevents the accumulation in the water circuit. The separated sludge, consisting of precipitated inorganic material (mainly sulphides) and carbon formed in small amounts from "$2\,CO \rightarrow C + CO_2$" is re-introduced into the thermal system.

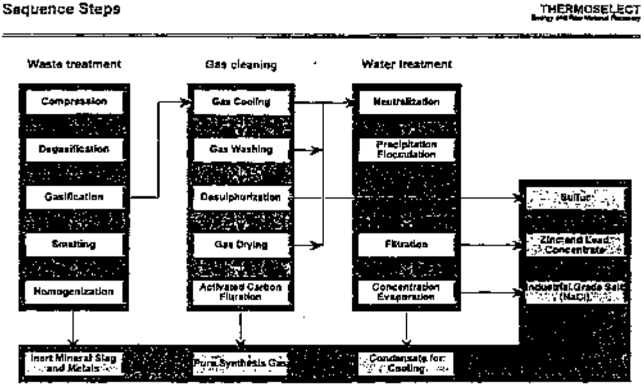


Figure 3 •

Independent quench water circuits equipped with redundant pumping systems guarantee for the safe functioning of the gas quenching.

Trace concentrations of roughly 0.01 ng TE PCDD/PCDF and 0.1 vol.% methane [9] measured in the quenched crude gas support the equilibrium in the high-temperature reactor and the effectiveness of the shock cooling.

The large amount of circulated quench water already absorbs approximately 90% of the hydrogen chloride contained in the gas phase and acts as a buffer when the composition of the waste fluctuates, so that the quality of the synthesis gas is homogenised before it goes into the actual fine gas purification stage.

An acid washing stage and an alkaline washing stage are integrated into the quench unit.

The synthesis gas partially cleaned in the combined quench washer is fed first to the fine particle absorber using glycerine as washing unit before it enters the sulferox purification stage. Oxidation of the pollutant hydrogen sulphide with a Fe-III-complex leads to the formation of hydrogen that goes into the synthesis gas, and elemental sulphur, which is separated from the wash solution. The oxidising effect also depletes other hazardous trace materials. [HS + Fe-III-complex -> 1/2 H_2 + S + Fe-II-complex]

Condensation of the water vapour from the synthesis gas is advantageously coupled with precipitation, if necessary, of acidic trace components that still exist. Doing this, the activated coke adsorption, after slight reheating of the synthesis gas 50_C, is not affected by condensation of water vapour. Because of the low working temperature and the practical absence of oxygen in the gas, build-up of feared "hot spots" with demobilisation of hazardous materials can be excluded.

The purity of the synthesis gas corresponds to that of natural gas. Thermoselect remains well below all standard emission threshold limits as e.g. for dioxins and furans: according to the German Emission Protection Regulations the threshold limit of 0.1 ng/m^3 has to be met. Thermoselect, by comparison, measures only 0.002 ng/m^3.

1.4 Process Water Purification And Use

All process-water streams are combined into one common process-water purification. With a hold-up volume of roughly 500 m^3, there is a quantity of water whose concentrations of contamination are not changed by short-term, even extreme fluctuations in the composition of the waste. Thus, for example, in a plant with a capacity of 20 Mg/h, doubling the chlorine content (e.g. due to high quantities of PVC in the trash or similar reason) causes only a slight increase in the chloride concentrations by 200 mg/litre, i.e., an increase in the concentration in the process water of only 0.02%.

The salt crystallised during the final vacuum evaporation of the process water removal (basically sodium chloride) is sufficiently pure for industrial use. The desalinated, distilled water, which meets drinking water quality requirements, is then available for use as cooling medium. Thermoselect is the only process recovering water from waste and reusing this water.

2 Process Control and Safety Aspects

An important aspect for the smooth operation of a chemical plant or a thermal waste treatment plant is the control of the process.

When the THERMOSELECT plant is startedup cold, the nitrogen atmosphere in the HTR from inertization is gradually replaced by the inert H_2O-CO_2 atmosphere from the stoichiometric conversion of fuel (for example natural gas or propane) with which the system is heated to temperatures over 700°C. Only after this threshold is reached the degasification channel is heated up by further parallel heating of the high-temperature reactor to the operating temperature. After a degasification channel temperature of 600°C is reached and a gas-expulsion temperature of >900°C in the HTR, solid fuel residues from the degasification channel are brought in by starting up the press, so that due to the excess carbon in the resulting pile, combustion with oxygen overall in gasification with oxygen turns into the products CO-H_2-CO_2 In the gas volume above the pile, the operating temperature reaches roughly 1200°C in a short time.

Under normal operation conditions, the process mainly is controlled by two loops. Loop A keeps the temperature of the gas phase measured at the HTR outlet above the set point temperature by introducing small amounts of metered oxygen into the upper HTR section. Setting free thermal energy by the exothermic reaction of O_2 with H_2 and CO allows to rise gas temperature smoothly with short response time.

In the second control loop total gas flow is measured against a set flow rate. The production rate of gas is governed by the (uncontrolled) heat value of the waste input and the amount of oxygen introduced into the lower HTR section. Controlling and metering the oxygen flow into the gasification zone, a constant synthesis gas flow is obtained despite the heterogeneous characteristics of the waste composition.

The press (loading system for the degasification channel) works in a fixed cycle. Using a gamma ray beam sent across the lower HTR section at the degasification channel transition point with the HTR, the press cycle is overridden as long as the beam is interrupted by the solid material pile in the gasification zone.

Free oxygen hypothetically present in the synthesis gas above the solid pile would immediately react at the high gas temperatures, which are way above the ignition temperature, with the super-rich supply of oxidizable reactants, so that explosive mixtures cannot occur in the reactor.

Moreover, the oxygen content of the synthesis gas is constantly monitored by means of measurement probes; the alarm threshold is 0.5 vol.%; starting at 1 vol.%, the oxygen supply of the lances is automatically interrupted. Thus the upper explosion limit with a proportion of roughly 4% oxygen cannot be reached, under which ignition of the synthesis gas would even be possible [10,11]. Regardless of these considerations, the reactor is structurally designed (design, welding, wall thickness) and the materials are selected (chrome-molybdenum-steel) so that it maintains inner pressures of at least 8 bar (operating pressure 0.35 bar) without plastic deformation. Large-scale continuous operation has clearly proven the effectiveness of the measures and phenomena described in terms of safe handling of synthesis gas.

Tested, largely standardised plants and types of pipes that meet the pertinent safety regulations *are* used for safe handling of the oxygen used for gasification. The pipelines are state-of-the-art and are laid in compliance with requirements for protection from mechanical and thermal effects [10].

Gas release via Safety Flare

The water lock situated behind the quench washer serves as a quasi-passive safety valve with an immersion depth that corresponds to a counter-pressure slightly above the operating pressure of the reactor. As a pressure-release device in the event of unexpected marginal pressure excesses, it releases the gas route to the flare with its chimney, through a direct line and an activated coke filter. The partially purified synthesis gas is burned up in the flare under controlled conditions, which is designed with a 30% reserve in terms of the total amount of synthesis gas produced; a multi-redundant ignition burner system with a diverse fuel supply guarantees ignition of the gas. Using monitored inertization of the tubing to the combustion chamber prevents the formation of gas/air mixtures capable of exploding or backfiring from the flare into the pipes assigned to the other lines.

If there is a hypothetical total failure of gas purification, the route to the flare can be quickly cleared by draining the water lock. Here it is essential that by immediately stopping the input to the press and stopping the oxygen supply to the high-temperature reactor, the amount of synthesis gas formed is drastically cut, and the result is increased residence time in the high-temperature reactor (by up to a factor of about 5). The water vapour content, which then increases proportionately, promotes the endothermic water gas reaction, which speeds up the temperature drop. Cooling the high-temperature reactor by 100°C reduces the steam pressure of the inorganic components by approximately one order of magnitude. From experience it is known that within one hour, gas production can be expected to fall by roughly a factor of 10.

Excessive gas production due to monofractions whose heat content clearly exceeds the plant-design value is countered by a short stoppage of the input and the amount of oxygen in part of the HTR.

3 Mass and Energy Balances

Raw materials recovery

Calorific value fluctuations in the waste from between 10 MJ/kg to 16 MJ/kg observed during the RWTUV measurement period yield the monitored quantities of synthesis gas of 800 to 1200 Nm^3/Mg.
Figure 4 shows the results of the mass balance relating to an input of 1000 kg of waste. The application of pure oxygen instead of air is closely connected to the quality of all generated products.

Products of the Process **THERMOSELECT**

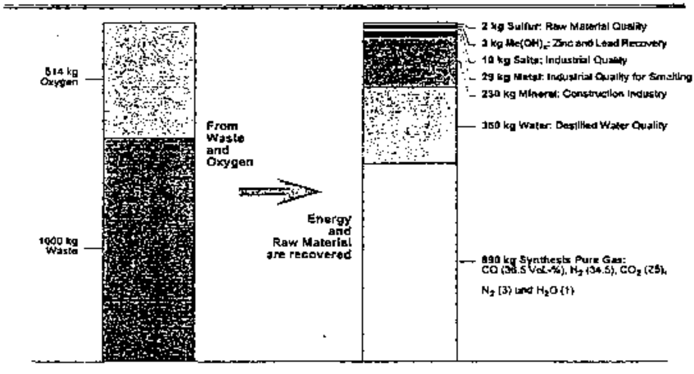
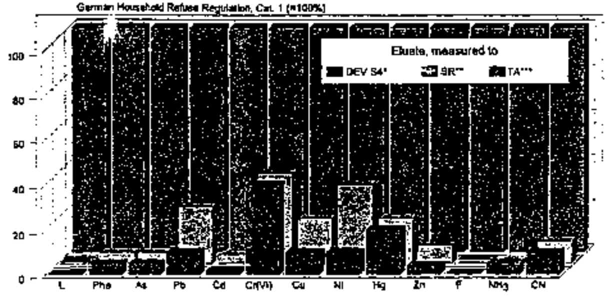


Figure 4

As an example, the following graph illustrates the very low teachability of the vitrified mineral granulate which reprensents one of the major products of the high temperature gasification and homogenisation process. This granulate is used e.g. as an additive in the production of concrete.

The other residues as metals, salt or sulphur are reused as well. Therefore, no further landfill is necessary.

Eluate values of the mineral granulate **THERMOSELECT**



DEV: German Standard Process; SR: Swiss Guideline; TA: German Household Refuse Regulation
L: Electrical Conductivity; Phe: Phenol Index; F: Fluoride; NH4: Ammonium-N; CN: Cyanide (easily liberated)

Figure 5

Flexibility and efficiency in energy recovery

Concerning the waste to energy conversion, the THERMOSELECT process introduces a new, very high degree of freedom. In a first step, the chemical energy content of the waste is efficiently converted into clean fuel gas (see Figure 6) which - in a second step - is then converted into thermal and/or electrical energy. This particular characteristic means flexibility for the choice of the synthesis gas conversion system: gas engines in simple and combined cycle, steam cycles, gas turbines in single or combined cycle, fuel cells, hydrogen engines. The synthesis gas can also be used in an already existing energy conversion plant eventually available on site, or it can be used for chemical synthesis in a nearby chemical plant.

Conversion of the synthesis gas into electricity is realized in the Fondotoce plant using gas engines (Otto cycle) having an efficiency of 34%.

Exergy Diagram (10 ton MSW/h)

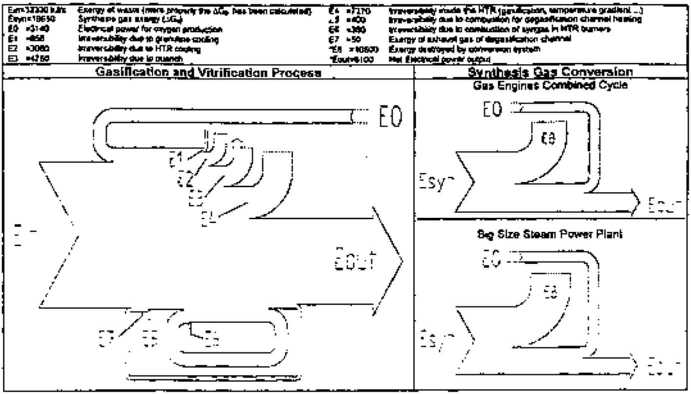


Figure 6 (* data referred to Gas Engines Combined Cycle)

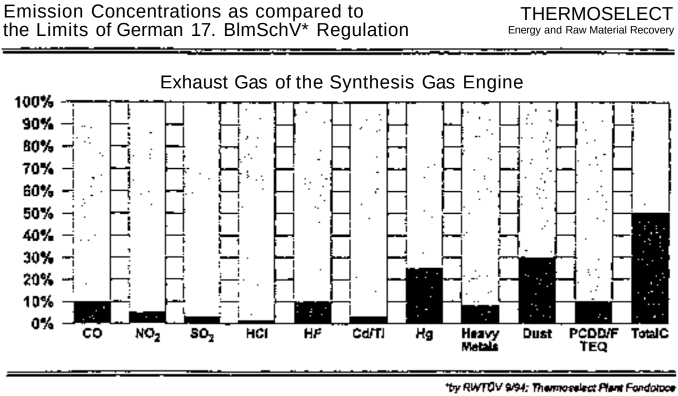
An exergetical balance is made and reported in figure 6. Because of the great flexibility of choice for the conversion system the balance is splitted into two phases: the first is equal for every system, the second is specific for each conversion system, which is reported here only for two cases: gas engines and gas converted in an already existing big size steam cycle.

The choice of the right conversion technique only depends on site specific, technical and economical conditions and requirements.

It is to be anticipated that site specific conditions will impact the mass and energy balances for each individual commercial facility; important factors necessary for efficient design include:

- Solid waste input quantity and calorific value.
- The seasonal fluctuation and expected component composition of the waste.
- The pollutant fraction and the composition of the solid waste input relative to heavy metals concentration, chlorine content and sulphur containing components such as rubber tires.
- The energy recovery technique chosen as already mentioned above.

The measured emission concentrations as compared to the limits of the 17th German Emission Protection Regulations (BImSchV) are shown in figure 7 for synthesis gas conversion in a internal combustion gas engine. These results demonstrate the advanced ecological standard of the THERMOSELECT system.



Emission Concentrations as compared to the Limits of German 17. BlmSchV* Regulation

THERMOSELECT
Energy and Raw Material Recovery

Exhaust Gas of the Synthesis Gas Engine

*by RWTÜV 9/94: Thermoselect Plant Fondotoce

Figure 7

4 Conclusion and Outlook

The THERMOSELECT waste melting process meets the key criteria established for modern thermal treatment techniques:

- ecologically sustainable due to extremely low emissions
- raw material recovery to an extent higher than 99%
- compact, modular, low size and standardised lines which can be multiplied to variable plant capacities
- reliable, simply to handle process coping with varying heterogeneous waste input
- highly efficient energy recovery to be integrated flexibly into site specific needs
- competitive costs, especially when taking into account lowest emission levels and the quality of the end products.

Thermoselect has adapted the natural method of coal formation and coal gasification to create an industrial scale technique for utilizing both the material and the energy content of waste. Thermoselect is the worldwide first closed loop process - realised in an industrial scale plant - integrating a direct melting system for the inorganic components of waste.