Compressor Issues for Hydrogen Production and Transmission Through a Long Distance Pipeline Network

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A hydrogen energy system will require the production of hydrogen from coal-based gasification plants and its transmission through long distance pipelines at 70 – 100 bar. To overcome some problems of current gasifiers, which are limited in pressure capability, two options are explored, in-plant compression of the syngas and compression of the hydrogen at the plant exit. It is shown that whereas in-plant compression using centrifugal machines is practical, this is not a solution when compressing hydrogen at the plant exit. This is because of the low molecular weight of the hydrogen. It is also shown that if centrifugal compressors are to be used in a pipeline system, pressure drops will need to be restricted as even an advanced two-stage centrifugal compressor will be limited to a pressure ratio of 1.2. High strength steels are suitable for the in-plant compressor, but aluminium alloy will be required for a hydrogen pipeline compressor.

Keywords: hydrogen production, coal gasification, CO_2 capture, H_2 purity & pressure

The introduction of hydrogen in the energy system as an energy alternative carrier is drawing much interest in Europe, offering significant advantages including reduction of greenhouse gas emissions, enhancing energy suppling security and improving economical competitiveness. Even at the present time, hydrogen has a major use in the petrochemical and oil refining industries, where it is mainly produced from natural gas and oil [1]. In the future, coal based gasification however, is likely to play a key role in European large-scale hydrogen production [2]. These will be based on entrained flow gasification as this type of gasifier maximises hydrogen production and facilitates the capture of carbon as CO_2, whereby it can be stored in geological reservoirs or used for enhanced oil recovery.

A condition for the successful penetration of hydrogen in the energy system is its supply from the production facility at high pressure, as this will enable hydrogen to be transmitted via long distance pipelines [3]. The hydrogen transmission pressure should be similar to that in the natural gas pipeline network, i.e. in excess of 70 bar. For a natural gas pipeline network, this pressure level is kept reasonably constant using booster stations with centrifugal compressors, which are set at about every 50 – 100 km and raise pressures by a factor of 1.2 – 1.4 at each compressor station. In general these compressors are of single or two stage type and are driven by an aeroderived gas turbine, fuelled by natural gas from the pipeline. The operating speed of the gas turbine and the pipeline compressor are similar, simplifying the drive mechanism.

For the same energy throughputs as a natural gas system, a hydrogen pipeline will be subjected to similar or even slightly higher pressure drops. Although the low density and viscosity of hydrogen will reduce pressure drops for the same flow rate as natural gas, because hydrogen has to be transmitted at 3 – 4 times the flow rate, to compensate for its much lower calorific value (10.8 MJ/Nm^3 compared with 35 – 40 MJ/Nm^3 for natural gas) pressure drops will be in fact greater. Unfortunately the compression of hydrogen using centrifugal machines is extremely difficult because of its low density. This is caused by the low molecular weight of hydrogen, requiring the use of multistage compression. The standard approach for hydrogen compression is that of reciprocating compressors. This is likely to be a more bulky and costly solution given the large throughputs. In addition operating speeds of reciprocating compressors are low which would seem to preclude the use of gas turbines as drivers, or if these have to be used, it will be necessary to use a large step down gear box [4].

Obviously it is advantageous if the delivery from the hydrogen production plant is at the highest possible pressure, close to 70 bar. However only one type of entrained flow gasifier that is currently available can do this. This utilises a water-based slurry feed and because of this feature, which enables it to operate at high pressures, its hydrogen production efficiency is about 2 % lower than dry feed gasifiers. The dry feed designs are limited to outlet pressures of 30 – 40 bar and to obtain a high pressure suitable for delivery to a pipeline two options are explored, that is the use of a compressor within the plant, partly through the processing route, and the use of a compressor after the purified hydrogen leaves the plant.

Obviously the use of a compressor results in an increased energy demand, the impact of which needs to be determined by process flow modelling. Some of this compression energy appears as temperature increase in the compressed gas, which in some cases can be used to replace other methods of heating. But as noted, the most important factor in compressor design is the molecular weight of the gas, as this will determine the pressure rise that can be achieved at each stage in the compressor.

Gasifier characteristics

The dry feed type of entrained flow gasifier, is at the present time the most efficient process for producing hydrogen from coal. In such a plant, nitrogen is used to transport the coal to the gasifier. It is well known that this reduces oxygen consumption compared to slurry feed gasifiers and is one reason why the dry feed type of design is a good option for the standard type of Integrated Gasification Combined Cycle (IGCC) plants which just produces electricity [5]. It is less well recognised that the use of slurry feeding reduces the amount of hydrogen that

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can be produced from a given amount of coal, since a greater proportion of the carbon in the coal is converted to CO_2 rather than CO, as the CO_2 reaction produces the heat needed to heat the water in the slurry to the reaction temperature. As a result less CO is available for the subsequent shift reaction to produce hydrogen.

Using nitrogen to transport the coal to the gasifier improves hydrogen production, but results in more purification and hence additional costs. More important, nitrogen transport limits gasifier pressures currently to about 30 – 40 bar [5].

Gasifier process route alternatives

All systems analyzed in this paper produce hydrogen only, the steam raised in the plant and the purge gas coming from the hydrogen purification stage are used to produce the power needed to run the ancillary equipment (i.e. the plant has no net power output). Some of the ancillary power can be generated from the steam produced in the gasification train, but as this is not sufficient, about 20 % of the hydrogen from the plant has to be used to generate extra power. A recent paper showed that this is best done using a small Combined Cycle Gas Turbine (CCGT) [6].

In slurry feed entrained flow gasifier, the hot raw syngas from the gasifier is cooled down to a suitable temperature using a water quench [5]. The raw syngas is then cleaned of particulates and passed through a series of CO shift converters, where CO reacts with steam to produce hydrogen and carbon dioxide. H_2S and CO_2 are then separately removed in an absorption system (e.g. Selexol®). The hydrogen, if necessary, is further purified in a Pressure Swing Adsorption (PSA) unit. The outlet pressure from the gasifier is around 75 bar. This concept is referred as Case 1 (fig. 1).

The dry feed entrained flow gasifier can also use a water quench to cool the hot raw gas from the gasifier. However the system modelled here utilizes a gas quench system in which cool gas is recycled back from further down the process to bring the raw gas temperature to below 800°C. The hot gas can then be used for steam raising. The main disadvantage of this system is that steam must be added later to promote the shift reaction. This concept is considered as Case 2 fig. 2).

In the alternative concept presented here, to overcome the pressure limitation associated with dry feed gasifiers, an in-plant compressor is positioned before the shift converter (see Figure 2) [7]. The temperature and the pressure of the gas at this point are 150°C and 35 bar. After the compression, the temperature and the pressure of the gas is raised to 250 – 270°C and 72 bar. The heat generated during compression of the syngas is used to promote the shift reaction and to raise steam, this having a positive influence on the overall plant efficiency. The gas is dry at the inlet of the in-plant compressor. As noted steam must be added to the syngas at the compressor outlet to promote the shift reaction. This concept is Case 3 (fig. 2).

Process flow simulation results

In all the plant configurations analysed in this paper, the thermal energy of the coal input was considered to be the same (2200 MW_{th}); the thermal energy of the fuel input and the hydrogen output are expressed taking into consideration the lower heating values – LHV (28.13 MJ/kg for coal and respectively 10.795 MJ/Nm^3 for hydrogen).

The heat and power balances (gas turbine and steam turbine generated power, ancillary power, hydrogen output), along with plant efficiency and CO_2 capture rate for the plant configurations analysed in this paper are presented in table 1. The hydrogen delivery pressure was considered in all cases to be 65 bar. All plant concepts were simulated using ChemCAD software.

For gasification processes assessed in this paper (coal – water slurry feed and dry coal feed gasifiers with and without in-plant compression), raw syngas compositions are presented in table 2. This shows the CO content of the syngas from the slurry gasifier is lower than that from the dry feed gasifiers, with the CO_2 content being correspondingly greater. As noted earlier this was an expected result and is in line with comments in the literature [5].

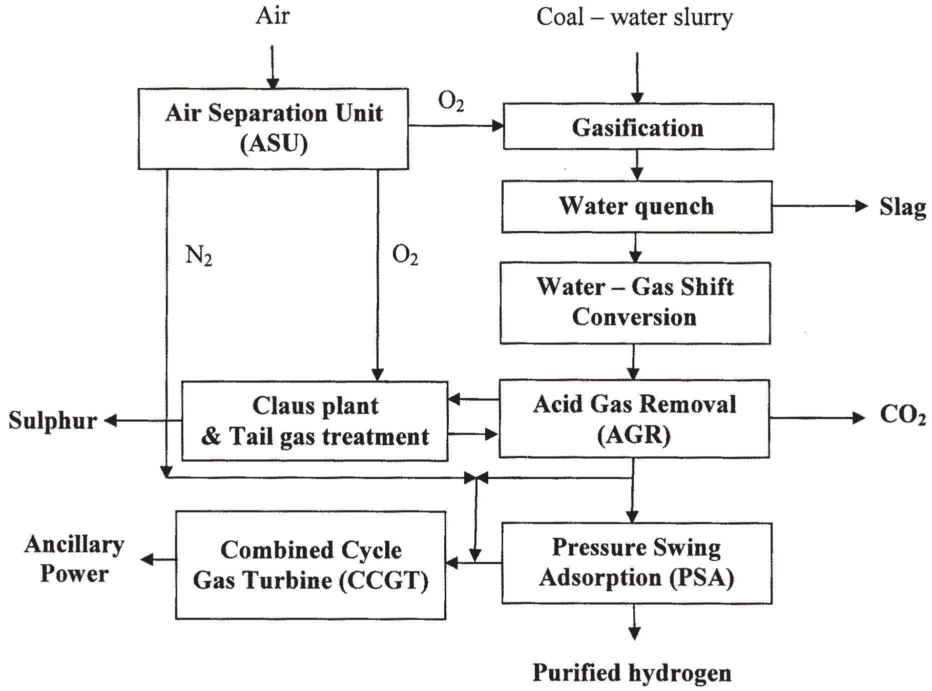

Fig.1. Scheme of hydrogen production based on slurry feed gasification with water quench, shift conversion, CO_2 & H_2S removal and H_2 purification

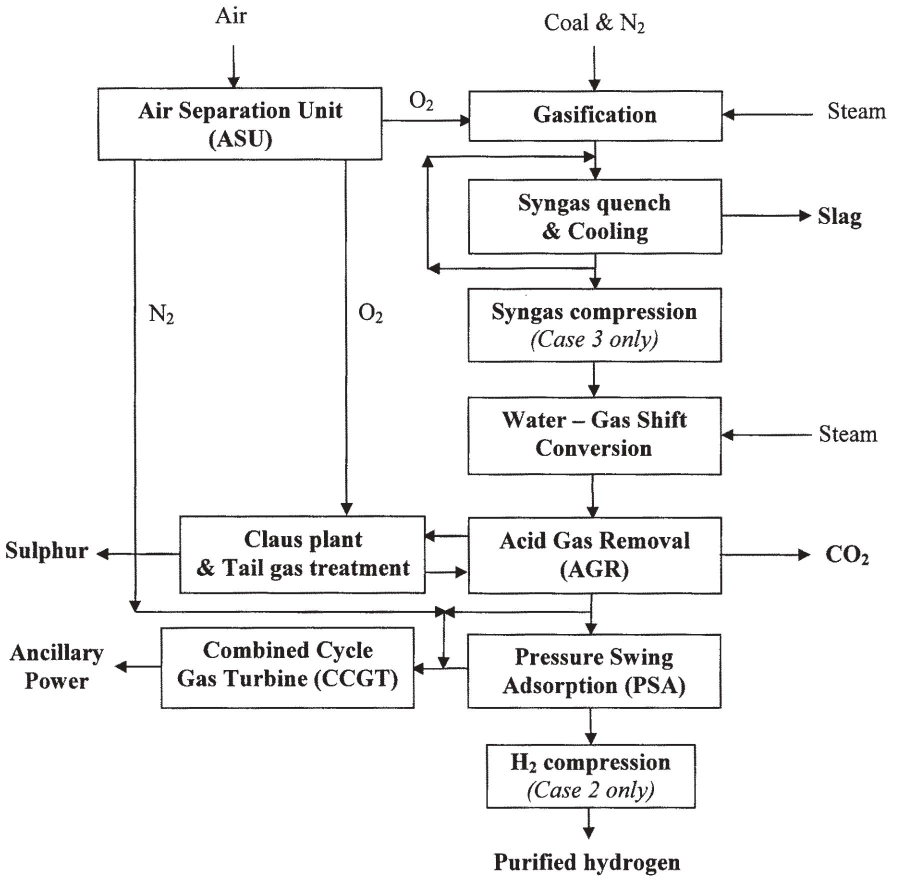

Fig. 2. Scheme of hydrogen production based on dry feed gasification, syngas compression (only Case 3), shift conversion, CO_2 & H_2S removal and H_2 purification

Table 1
PERFORMANCES OF THE INVESTIGATED PLANT CONFIGURATIONS

Property \ Case study	Case 1	Case 2	Case 3
Fuel input (MW_{th})	2200	2200	2200
Power generated (MW_e):			
- Gas turbine	100.60	100.96	102.04
- Steam turbine	144.52	143.48	161.00
Power consumed (MW_e):			
- Air Separation Unit (ASU)	71.68	65.16	65.20
- O_2 compressor	37.04	28.04	27.36
- Gasification auxiliary	22.00	22.00	22.00
- Process pumps	1.36	2.00	2.08
- Cooling water pumps	6.16	5.28	5.20
- Quench gas recirculation blower	0.00	5.12	5,12
- Syngas compressor	0.00	0.00	25.96
- Acid Gas Removal (Selexol®) system	29.96	18.00	30.44
- Purge gas compressor	13.24	16.88	16.04
- CO_2 compressor	63.68	66.6	63.64
- H_2 compressor	0.00	15.36	0.00
Net power output (MW_e)	0	0	0
CO_2 captured (%)	96.46	93.30	96.18
Hydrogen flow (Nm^3/h)	392620	424104	429460
Hydrogen output (MW_{th})	1177.32	1271.72	1287.78
Plant efficiency (%)	53.51	57.80	58.53

The results presented in table 1 indicate that the efficiency of dry feed gasifier with in-plant compression (Case 3) is significantly better, by 5 %, than the slurry feed gasifier (Case 1) mainly because of the higher cold gas efficiency of the dry gasifier and the fact that the heat generated during syngas compression stage is used to promote the shift reaction. Also the results shows that the performances of dry feed gasifiers with an in-plant

Table 2
COMPOSITIONS OF THE RAW SYNGAS

Composition (vol. %) \ Case study	Case 1	Case 2	Case 3
CO	42.24	61.56	61.56
H_2	25.33	26.54	26.54
CO_2	10.61	2.62	2.62
H_2O	20.81	4.58	4.58
CH_4	0.01	0	0
H_2S	0.26	0.31	0.31
COS	0.02	0.03	0.03
N_2	0.64	4.27	4.27
O_2	0	0	0
Ar	0.08	0.09	0.09

compressor (Case 3) is actually slightly better, by 0.73 %, than the conventional dry feed gasifier (Case 2).This is partly due to the heat generated during compression of the syngas, which reduces the need to preheat the gas before shift conversion, but it is also because CO_2 capture is easier, because of the higher pressures in the AGR system.

The CO_2 capture rates for Case 1 and 3 configurations are more than 95 % of the CO_2 quantity generated in the gasification process (percentage of carbon in the input coal that is captured as CO_2). The conventional dry feed gasifier (Case 2) shows a CO_2 capture rate around 93 % because of the lower pressure of the absorption system (28 – 29 bar) compared with the other two cases (66 – 67 bar).

Implications for design of compressors

Calculations and basic data

The main parameters that are needed for the design of centrifugal compressors are the volumetric flow rate, the molecular weight of the gas, its specific heat and the pressure ratio from inlet to outlet. The pressure ratio is governed by the peripheral speed of the compressor disc and also by the density of the fluid. High speed and high molecular weight translate into a large pressure increase, because of the momentum of the gas as it leaves the edge of the disc. Disc speed is limited, however, by the strength to density ratio of the wheel material. Although alloy steels are strong, the high density precludes a peripheral speed higher than about 225 m/s [8]. However since the strength of steel is relatively temperature independent up to 400°C, steels can be used for the in-plant compressor. Compressor discs made out of high strength aluminium alloys can run at peripheral speeds of 400 m/s [9], but they are limited to inlet temperatures close to ambient.

Standard calculation methods for estimating the pressure rise across a centrifugal compressor start by calculating the work done in increasing the momentum of a gas between entering the eye of the compressor and leaving the periphery. This way, the gas density enters into the calculation. The work done is used to estimate the temperature rise, taking into account the specific heat of the gas, and compressor efficiency losses. This temperature rise is then used to calculate the pressure ratio using the usual standard thermodynamic functions involving the specific heat of gas. Whereas the pressure ratio is mainly dependent on the molecular weight, the absolute increase in pressure is dependent on the gas density, which is partly governed by the pressure at the inlet to the compressor.

Table 3
COMPRESSOR DESIGN DATA

Parameter	In-plant	Plant outlet	Hydrogen pipeline
Gas	Raw Gas	Hydrogen	Hydrogen
Inlet temperature	150°C	25°C	25°C
Inlet pressure	35 bar	28 bar	58 bar
Outlet pressure	72 bar	65 bar	70 bar
Mass flow	544128 kg/h	38300 kg/h	38300 kg/h
Specific heat	1.4516 kJ/kg*K	14.347 kJ/kg*K	14.347 kJ/kg*K
Ratio of Specific Heats (γ)	1.4045	1.4093	1.4093
Molecular weight	21.1 kg/kmole	2.02 kg/kmole	2.02 kg/kmole
Peripheral speed	225 m/s	225 m/s	393 m/s
Isentropic efficiency	0.8	0.8	0.8
Slip Factor	0.9	0.9	0.9
Power Input Factor	1.04	1.04	1.04
Pressure rise per stage	1.44	1.03	1.1
Number of stages	2	>20	2

The compressor used in this paper is based on a compressor intended for a small jet engine, in which air at atmospheric pressure was the working fluid [9]. Mass flow of this machine was 0.31 mols/sec at 22°C. However the high inlet pressure in the gasifier allows this type of machine to accept a much higher mass flows since the "real" volumetric flow is similar. The overall diameter of the compressor is 0.5 m and the inlet diameter is 0.3 m. This initial design requires an estimate of the compressor isentropic efficiency and the slip factor. The latter is the ratio of gas velocity, leaving the compressor, to the peripheral speed. A common value of slip factor is about 0.9. Table 3 gives the values for the in-plant compressor and for hydrogen compressor at the back end of the plant. The table also includes details of a pipeline compressor.

In-plant compression

The key factors in governing this design are the inlet temperature and the relatively high molecular weight of the gas. The pressure ratio for one stage of an in-plant compressor will be 1.44, at which the speed will be 9000 rpm. Hence a two stage machine will give the required pressure rise of 2. It would be advisable to use a small intercooler between the two stages to compensate any overheating that could result from a degradation of compressor efficiency because of fouling. The intercooler would normally absorb only a relatively small amount of heat. In practice two single stage machines of very similar design could be used in series. Overall power requirement would be just under 26 MW.

Compression of hydrogen at the plant exit

The power requirement of the compressor at the plant exit is 15.36 MW if such a compressor could be built, the required pressure ratio to give an inlet pressure to the pipeline pressure of 65 to 70 bar being 2. This is a very severe requirement for a centrifugal hydrogen compressor. If this compressor was to be made of steel it would require over 20 stages, as pressure rise in each stage would be of the order of about 1.03. Given that the sets of wheels would be 0.5 metres in diameter and the machine would be around 10 metres long, this is clearly an impractical solution. An aluminium alloy compressor, because of its higher operating speed could give a high pressure ratio but even in this case more than five stages would be needed.

Pipeline compression using centrifugal machines

It is clear that a centrifugal machine using a steel compressor would be of little use for pipeline compression. A machine with a disc constructed from aluminium alloys, having a peripheral speed of 393 metres per second will give a pressure ratio of 1.1 per stage. Even this pressure ratio is of no practical use, but a two stage machine would give a pressure ratio of 1.2 which is just about acceptable. But to give the same energy flows as with a natural gas pipeline of the same diameter, compressor stations would have to be 20 – 30 % closer than those of today.

Conclusions

In this paper, different hydrogen production plant configurations based on coal gasification with CO_2 capture were analysed by modelling using ChemCAD software package. The paper aimed at increasing the hydrogen outlet pressure and ensuring the production of purified hydrogen with the least penalty in plant efficiency.

Dry feed coal gasifiers are more efficient than water based slurry gasifiers. However, their operating pressure makes them less suitable for hydrogen plants that would supply hydrogen via long distance pipelines to consumers. This paper has shown that these shortcomings could be overcome by using in-plant compression to raise the pressure to standard natural gas pipeline levels.

The simulation results also show that this change can be made without compromising the hydrogen production efficiency. On the contrary, the efficiency is slightly increased because of the better thermal integration of the plant and because the CO_2 capture penalty decreases due to the increased operating pressure of the acid gas removal (AGR) system.

Some design issues of the in-plant syngas compressor were also discussed in analogy with a hydrogen compressor used at the exit of the plant. The conclusion is that an in-plant compression is definitely a practical and possible solution for overcoming the pressure limitation of the dry feed entrained flow gasifiers used for hydrogen production. The paper also shows that if pipeline compression is required, pressure drops along the system will need to be reduced, since even with a two stage machine with aluminium alloy compressors pressure ratios are limited to 1.2.

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