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Fluidized-bed gasification of high-ash South African coals: An experimental and modelling study

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Abstract—The gasification of two high-ash South African coals were studied using a pilot-scale bubbling fluidized-bed coal gasifier operated at atmospheric pressure with oxygen enriched air and steam as the gasification agents. Tests were carried out at temperatures between 875 and 975°C and char residence times of between 15 and 55 minutes in the gasifier. The results of the test show that the fixed carbon conversion increases with an increase in temperature and residence time of char particles in the gasifier.

The experimental results were used to calibrate a commercially available fluidized-bed gasifier simulation model (CeSFaMB). The predictive capability of the model was analysed in terms of the degree of variation between experimental and simulated results for each test. The calibrated model was used to design a 15 MW fluidized-bed coal gasifier.

INTRODUCTION

Annually approximately 300 million tonnes of high-ash coal is mined in South Africa supplying 74% of it primary energy requirements. In order to produce coal for the domestic and export markets a large percentage of the mined coal is beneficiated (washed) to produce lower-ash products. The beneficiation of high-ash coal results in the production of 55 million tonnes of discards that have an ash content of between 55% and 70%. The ash content of coal available to the domestic market is expected to increase in future since lower grade coal (high-ash) seams are being mined and coal washing is being scaled down as a result of environmental legislation.

Fluidized-bed combustion and gasification technologies have been identified as potential technologies that can be applied for the utilisation of raw (unwashed) high-ash coals and stockpiled discards with reduced environmental impact. The advantage of fluidized-bed gasification it that gas can be produced for heating applications, Integrated Gasification Combined Cycle (IGCC) power generation and for the production of synthetic fuels and chemicals utilising high-ash coals.

The advantages of fluidized-bed gasifiers compared with fixed-bed and entrained-flow gasifiers are that owing to the lower operating temperature (900–1000°C), oxygen consumption is lower, sensible heat losses in the gas is lower and refractory life is increased. When high-ash coal is used the drainage rate of bed ash is higher which reduces accumulation of coarse particles on the distributor, thereby preventing hotspot formation and clinkering of the bed. The disadvantage of fluidized-bed gasification compared with fixed-bed and entrained-flow gasifiers is that the carbon conversion is lower. The lower carbon conversion is due to lower temperature, attrition and bypassing of char in the reactor and low char reactivity.

Fluidized-bed gasification tests done earlier [1] using air and steam as the gasification agents indicate that the gas calorific value of the gas is low when high-ash, low-reactivity coals are gasified. The objective of the present investigation is to increase the calorific value and fixed carbon conversion in the gasifier by enriching the gasification air with oxygen.

FLUIDIZED-BED GASIFICATION TESTS

A bubbling fluidized-bed gasifier (BFBG) pilot plant at the CSIR was used to investigate the effect of using oxygen-enriched air on the gasification of two high-ash South African coals. A flow diagram and the specifications of the pilot-scale BFBG are given in Figure 1 and Table 1.

Process description

Coal, air, oxygen and steam are the input streams to the process, which produces the output streams: gas and char (ash). Coal is fed to the gasifier by means of a screw conveyor at a height of 1.5 m above the distributor. Steam is generated in an electrode boiler and is mixed with air and oxygen at the inlet to a shell-and-tube heat exchanger. The preheated steam, air and oxygen stream is injected into the gasifier via a nozzle-type distributor. Char (bed char) is removed from gasifier by means of a water-cooled screw conveyor and from the gas (cyclone char) by means of a cyclone which is placed after the gas cooler.

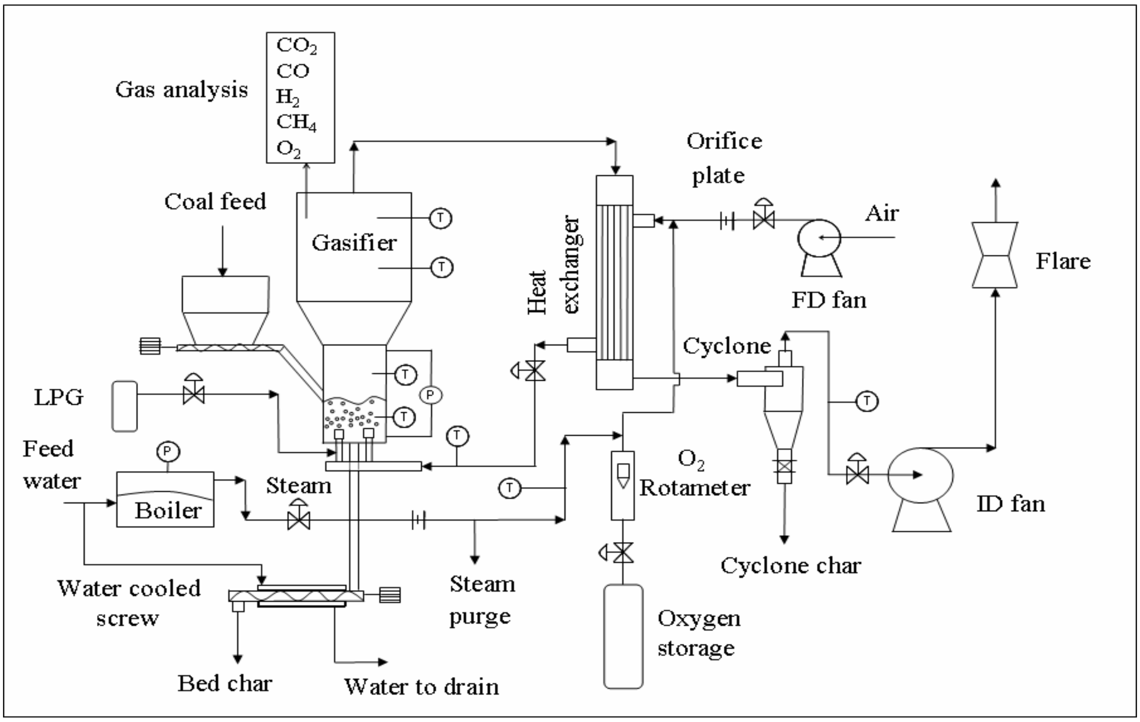

Figure 1. Pilot-scale fluidized-bed gasification plant

The de-dusted gas is combusted (flared) before it is vented to atmosphere. Coal particles that enter the furnace via the coal feed chute drop into the fluidized-bed section and start conversion to gas and char. The char particles move rapidly up and down between the gasification and combustion zones in the bed. The combustion zone is limited to the lower 10–15% of the bed above the distributor and is rich in oxygen. Owing to the fluidizing action of the bed, the char particles experience attrition and break down into smaller particles. When the particles are small enough, they are entrained into the freeboard section (upper part) of the furnace. Owing to the expanded nature of the freeboard, the gas velocity decreases and the particles fall back to the bed, resulting in internal circulation of particles between the bed and the freeboard. Further breakdown of the char particles results in their terminal falling velocity (U_t) being lower than the freeboard velocity and they are elutriated from the furnace. A significant proportion of the char particles (40–60%) are not elutriated from the furnace and these are drained from the bottom of the bed in order to maintain a constant fluidized-bed height.

BFBG start-up and control

The BFBG is started up by adding 15 kg of silica sand (0.4–0.85 mm) to the furnace. The silica sand is fluidized by starting the forced-draught (FD) and induced-draught (ID) fans. LPG is injected into the fluidized silica sand bed via the distributor nozzles. The LPG is ignited by means of a pilot flame which is inserted through the furnace door and directed down towards the bed. When the temperature reaches 850°C, the pilot flame (lance) is removed and the furnace door is closed. The temperature is further increased to 930°C by increasing the LPG flow. The furnace is operated with LPG at 930°C for 6 hours to allow thermal soaking of the refractories and heating of the freeboard. After 6 hours, coal, steam and oxygen are added to the furnace and the temperature of the bed is controlled at the required set-point by varying the oxygen flow. The furnace is operated for a further 6 hours to allow the bed carbon content and freeboard temperature to stabilize. Once stable conditions have been achieved, operating data are recorded and samples are collected for a period of 3 to 4 hours.

Table 1. Specifications of the BFBG pilot plant

Operating pressure	atmospheric
Bed dimensions (m)	0.2 × 0.2 (square)
Freeboard dimensions (m)	0.40 × 0.40 (square)
Furnace height (m)	4 (2 m bed & 2 m freeboard)
Fluidized-bed height (m)	< 0.6
Coal feed rate (kg/h)	15–35
Coal particle size (mm) (d_{50})	1. 2–1.9
Coal CV (MJ/kg)	> 10
Air flow rate (Nm^3/h)	15–60
Oxygen flow rate (kg/h)	4–16
Steam flow rate (kg/h)	5–35
Bed temperature (°C)	860–980
Air, steam and oxygen temperature (°C)	155–300
Fluidizing velocity (m/s)	1.2–2.2

Coals selected for BFBG tests

Two high-ash South African coals that are currently being used as feed coals to the Lethabo and Matimba power stations were selected for the fluidized-bed gasification tests. The proximate, ultimate and reflectance analysis of these coals on an air-dried basis are given in Table 2.

Table 2. Analysis of New Vaal and Grootegeluk coals (air-dried)

	New Vaal	Grootegeluk
Calorific value (MJ/kg)	14.84	21.40
Ash content (%)	40.7	31.7
Moisture (%)	5.7	1.9
Volatile matter (%)	20.5	28.3
Fixed carbon (%)	33.1	38.1
Total sulphur (%)	0.84	1.17
Ultimate analysis:		
Carbon (%)	39.25	52.93
Hydrogen (%)	3.45	4.11
Nitrogen (%)	0.90	1.19
Sulphur (%)	0.84	1.17
Oxygen (%)	9.16	7.00
Reflectance analysis:		
Vitrinite random reflectance (%)	0.55	0.71

The analysis shows that the selected coals have high ash and low inherent moistures contents. The vitrinite random reflectance shows that the New Vaal coal has a lower rank than the Grootegeluk coal.

Test results

The operating conditions used during the fluidized-bed gasification tests using Grootegeluk and New Vaal coals are given in Table 3.

Table 3. BFBG operating conditions

Coal	New Vaal	Grootegeluk
Temperature (°C)	875–975	875–975
Residence time (min)	15 –36	30–55
Fluidizing velocity (m/s)	1.5–2.1	1.2–2.2
Oxygen enrichment (%)[1]	35–37	30–38
Absolute pressure (kPa)	95	95
Coal particle size (mm)[2]	1.7	1.6

[1]Oxygen concentration of the enriched "air" stream
[2]d_{50}

To represent the operating conditions given in Table 3, six pilot-scale gasification tests were carried out on New Vaal coal and five on Grootegeluk coal. The results of the tests on New Vaal and Grootegeluk coal are summarized in Tables 4 and 5.

Table 4. Summary of fluidized-bed gasification tests on New Vaal coal

Test number	1	2	3	4	5	6
Mid-bed temperature (°C)	880	944	943	944	940	976
Average char residence time (min)	25.0	18.1	25.4	30.4	30.0	25.8
Coal feedrate (kg/h)	32.2	32.2	32.2	32.2	26.9	32.2
Airflow (Nm^3/h)	24.2	23.8	24.4	23.8	21.9	24.0
Steam flow (kg/h)	16.5	15.6	16.0	16.0	16.5	13.9
Oxygen flow (kg/h)	8.9	9.0	9.2	9.3	8.7	9.9
Oxygen in enriched air (%)[1]	35.7	36.1	35.8	36.0	36.6	36.7
Oxygen in total inlet flow (%)[2]	21.1	21.7	21.5	21.5	21.4	23.3
Total inlet flow temperature (°C)	281	275	299	285	281	285
Coal particle size (mm)[3]	1.80	1.6	1.40	1.75	1.7	1.70
Minimum fluidizing velocity (m/s)	0.24	0.21	0.14	0.21	0.16	0.21
Superficial gas velocity (m/s)	1.64	1.67	1.70	1.69	1.59	1.66
Lower bed temperature (°C)	908	965	958	960	958	997
Gasifier exit temperature (°C)	735	730	762	757	743	757
Dry gas composition (Volume %)						
CO	16.60	19.90	19.40	17.60	18.90	19.10
H_2	24.00	24.3	23.90	22.70	24.20	22.50
CH_4	1.30	1.10	1.0	1.20	1.00	1.0
CO_2	20.70	18.80	19.10	20.80	20.30	20.20
N_2 + others[4, 5]	37.4	35.90	36.5	37.6	35.5	38.6
Gas calorific value (MJ/Nm^3)	5.75	6.12	5.97	5.69	5.94	5.77
Bed pressure drop (Pa)	1905	1185	1868	2356	1839	1896

Table 4—continued

Ash extracted from the bed (kg/h)	6.1	5.64	4.81	6.78	4.40	6.63
Carbon in bed ash (%)	13.7	7.2	4.4	2.92	3.21	2.50
Bed ash particle size (mm)	0.90	0.81	0.65	0.79	0.69	0.81
Ash elutriated to cyclone (kg/h)	9.04	8.91	9.66	7.92	7.27	6.63
Carbon in cyclone ash (%)	24.7	23.24	22.66	17.62	19.78	14.9
Cyclone ash particle size (μm)	60	62	62	50	62	40
Ash elutriated (%)	59.72	61.24	66.76	53.87	62.28	54.10
Fixed carbon conversion (%)	73.40	78.54	79.25	85.05	83.65	87.54
Total carbon conversion (%)	78.37	82.54	83.12	87.40	86.68	89.49
Cold gas efficiency (%)[6]	60.4	65.5	65.3	61.4	62.8	63.1

[1]Oxygen concentration of combined air and oxygen flow stream
[2]Total inlet flow consists of air, steam and oxygen
[3]d_{50} – 50% of the coal mass is less than the d_{50} size
[4]Others are < 0.4% and include H_2S, NH_3, HCN and C_2^+
[5](N_2 + others) by difference
[6]Energy in the cooled gas as a percentage of the energy in the coal

Table 5. Summary of fluidized-bed gasification tests on Grootegeluk coal

Test number	1	2	3	4	5
Mid-bed temperature (°C)	979	983	976	943	876
Average char residence time (min)[7]	52.5	40.4	32.8	34.5	33.9
Coal feedrate (kg/h)	16.0	19.6	23.0	23.0	23.0
Airflow (Nm^3/h)	15.9	20.9	22.1	22.7	24.5
Steam flow (kg/h)	11.5	15.0	18.3	20.0	17.8
Oxygen flow (kg/h)	6.5	7.6	9.5	7.2	4.9
Oxygen in enriched air (%)	37.5	36.0	37.5	34.4	30.0
Oxygen in total inlet flow (%)	21.9	20.9	20.6	18.0	16.7
Total inlet flow temperature (°C)	247.7	273.8	293.9	287.4	255.7
Coal particle size (mm)[3]	1.7	1.4	1.4	1.7	1.7
Minimum fluidizing velocity (m/s)	0.09	0.13	0.16	0.21	0.29
Superficial gas velocity (m/s)	1.19	1.54	1.75	1.76	1.63
Lower bed temperature (°C)	998	1003	999	958	908
Gasifier exit temperature (°C)	752	768	808	810	747
Dry gas composition (volume %)					
CO	16.6	17.0	16.1	13.6	10.5
H_2	20.9	20.7	18.6	18.6	14.8
CH_4	1.5	1.4	1.7	2.2	2.9
CO_2	22.8	20.0	21.2	22.3	20.5
N_2 + others	38.1	40.8	42.3	43.2	51.2
Gas calorific value (MJ/Nm^3)	5.48	5.46	5.46	5.09	4.51
Bed pressure drop (Pa)	1816	1816	1816	1964	1964

Table 5—continued

Ash extracted from the bed (kg/h)	3.27	3.10	5.05	5.90	8.85
Carbon in bed ash (%)	4.50	10.80	21.71	28.24	36.3
Bed ash particle size (mm)	0.52	0.70	0.88	1.05	1.35
Ash elutriated to cyclone (kg/h)	2.88	5.0	5.18	5.12	3.23
Carbon in cyclone ash (%)	32.4	30.7	35.51	40.10	49.0
Cyclone ash particle size (μm)	67	67	70	72	70
Ash elutriated (%)	46.86	61.54	50.69	46.49	26.74
Fixed carbon conversion (%)[8]	82.26	75.08	66.51	57.56	45.23
Total carbon conversion (%)	87.23	82.06	75.89	69.48	60.57
Cold gas efficiency (%)	57.43	51.36	45.11	44.44	41.02

[7] The average char residence time was calculated using the coal feed rate and bed pressure drop
[8] The fixed carbon conversion is calculated using the—
- Coal feed rate
- Fixed carbon in coal
- Flows of bed and cyclone ashes
- Carbon contents of bed and cyclone ashes

The effect of temperature on the fixed carbon conversion for New Vaal and Grootegeluk coals is given in Figure 2.

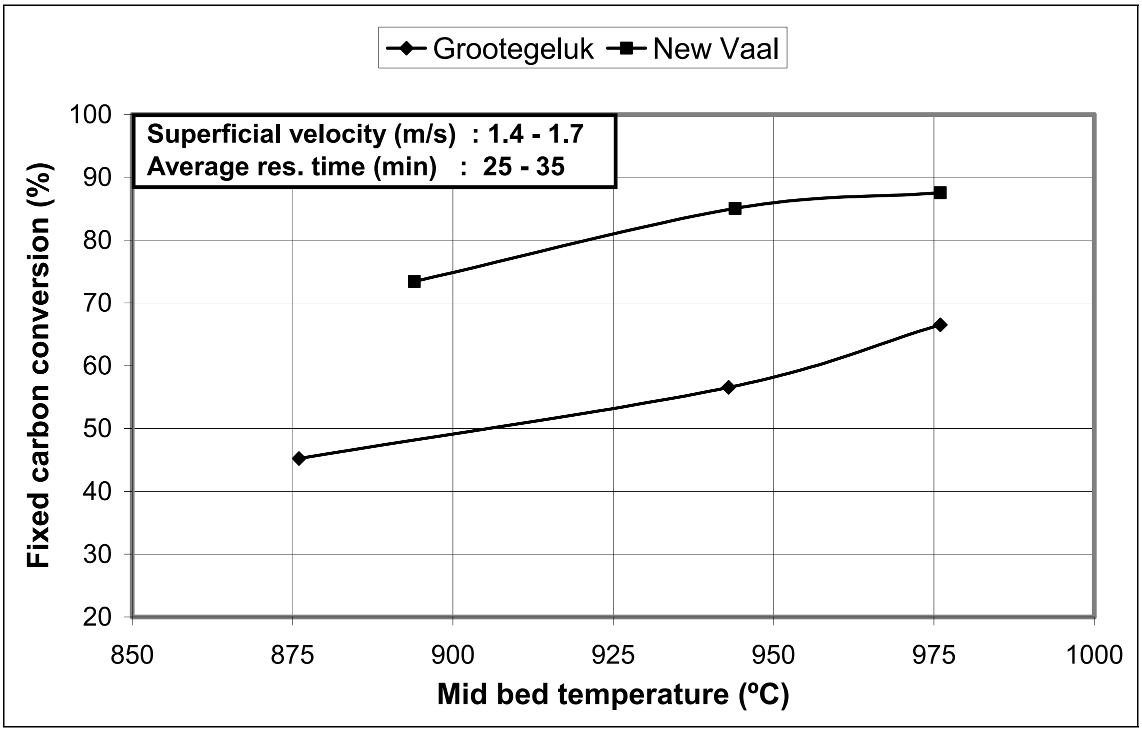

Figure 2. Fixed-carbon conversion as a function of temperature

Figure 2 shows that the fixed carbon conversion achieved in the BFBG pilot plant increases with temperature and that owing to the higher reactivity of the New Vaal coal the fixed carbon conversions were higher than those obtained for the Grootegeluk coal. The effect of residence time on fixed carbon conversion is given in Figure 3.

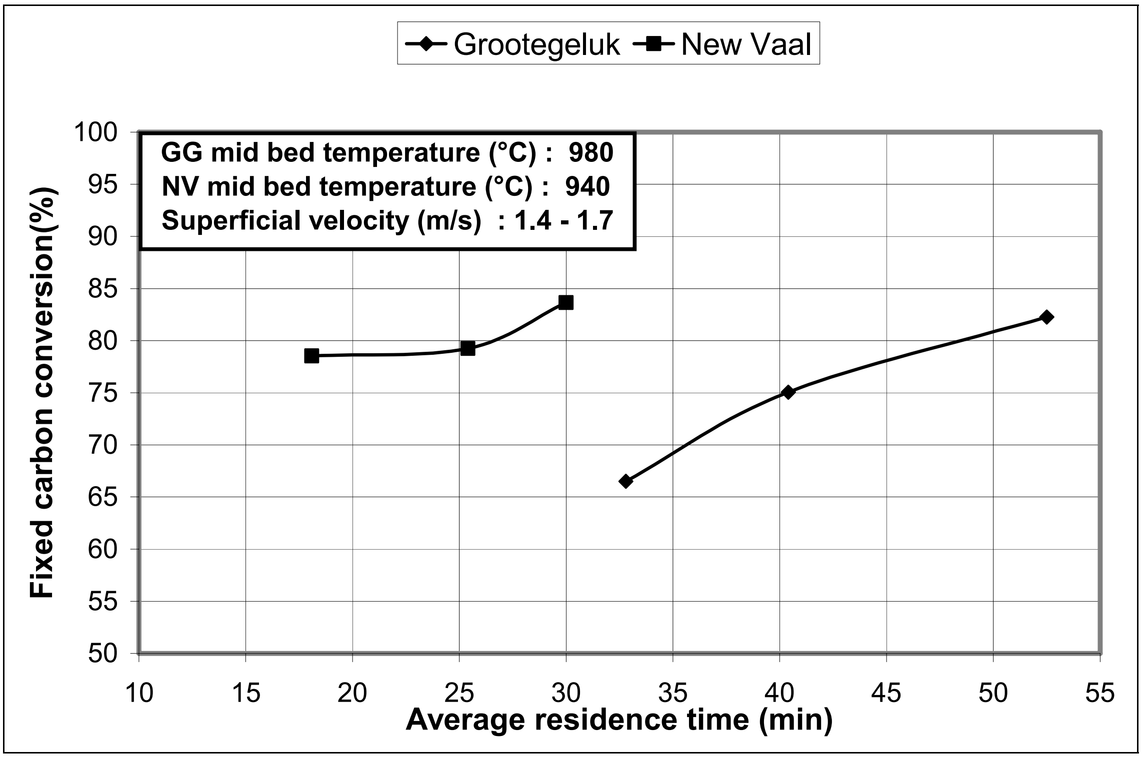

Figure 3. Fixed-carbon conversion as a function of residence time

Figure 3 shows that the fixed-carbon conversion in the BFBG pilot plant increases when the residence time of char particles in the gasifier are increased. The fixed carbon conversion of the lower reactivity Grootegeluk coal can be increased to over 80% if the residence time is increased to 52.5 min. The higher fixed-carbon conversion is however achieved at the expense of plant thermal output since the residence time of char in the gasifier is increased by decreasing the coal feedrate.

The effect of temperature on the calorific value of the gas is given Figure 4.

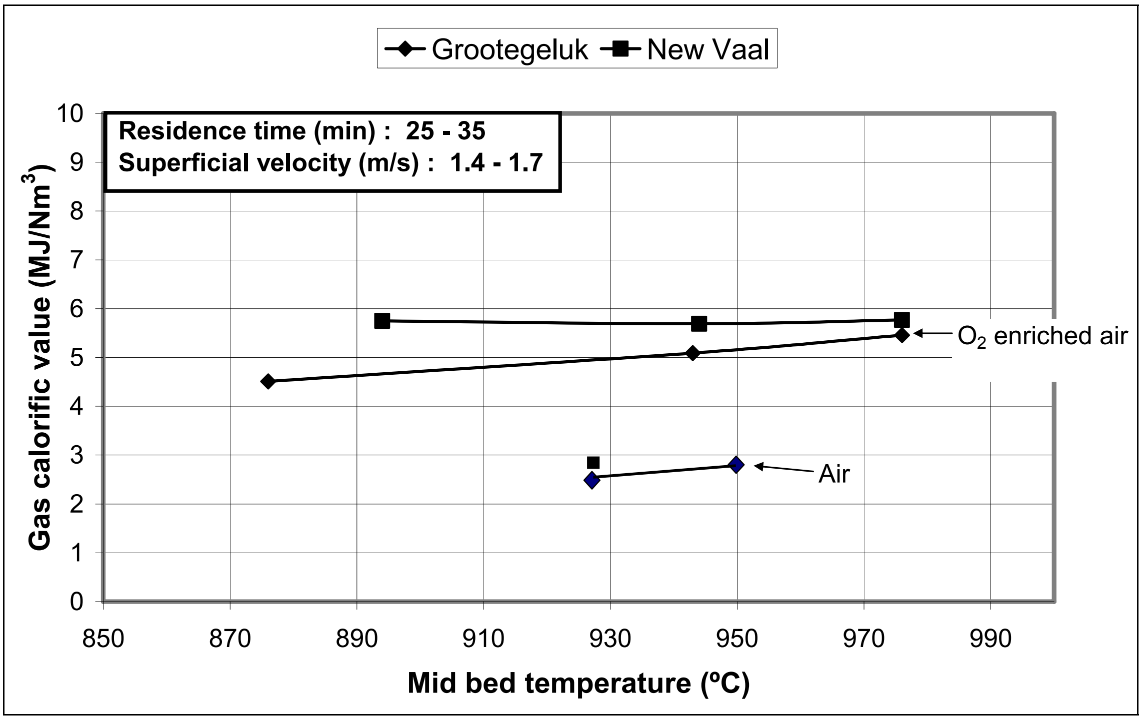

Figure 4. Gas calorific value as a function of temperature

Figure 4 shows that because of the higher fixed-carbon conversion of New Vaal coal a higher calorific value was achieved in the gasifier. Calorific values obtained during earlier air-blown gasification tests [1] are also shown in Figure 4. The oxygen enrichment gasification tests produced a significantly higher calorific value than the air-blown tests since the dilution effect of nitrogen is reduced and the fixed-carbon conversion is higher because of the higher steam concentration in the gasifier.

Owing to thermal fragmentation and attrition of char, fines are produced, which are carried from the gasifier. Tables 4 and 5 show that the elutriated chars has a higher carbon content than the bed char because of the shorter residence time of the elutriated char.

FLUIDIZED-BED COAL GASIFIER MODELLING

Background

The objective of fluidized-bed coal gasifier modelling is to predict the performance of the gasifier based on given input conditions. The input conditions usually include—

- Coal feed rate and analysis
- Flow rate, temperature and pressure of air, oxygen and steam
- Fluidized-bed height
- Gasifier design parameters such as diameter, height and thickness of thermal insulation

The performance parameters include—

- Gasifier temperature
- Carbon conversion
- Gas flowrate and composition

Fluidized-bed gasifier models can be used for—

- Design, optimisation and scale-up
- Assistance with accessing start-up and shutdown conditions
- Adaptive control
- Trouble shooting

Fluidized-bed gasifier models

Fluidized-bed gasifier models that have been used (with increasing degree of complexity) are mass and energy balance models, equilibrium models and kinetic models.

Mass and energy balance models

Mass and energy balance models are based on the conservation of mass and energy. Carrying out mass and energy balances over the gasifier results in more unknowns (output variables) than equations, meaning that more than one possible output condition can satisfy the mass and energy balance equations. In this case assumptions regarding some of the output variables, such as the fixed carbon conversion, methane concentration and carbon monoxide composition of the gas are required in order to calculate the values of the remaining output variables.

Equilibrium models

Equilibrium models add equilibrium equations to the mass and energy balance equations, which reduces the difference between the number of variables and equations to one and only an assumption regarding the fixed carbon conversion is required in order to calculate the values of the other output variables. However, owing to the short gas residence time in a BFBG the gases are seldom in equilibrium which reduces the predictive capability of equilibrium models. Equilibrium models are useful in that they can indicate output conditions that are not attainable owing to equilibrium limitations. This can assist in the estimation of the methane and carbon monoxide concentrations that are required in the mass and energy balance model.

Kinetic models

Kinetic models include the rates of the heterogeneous (gas-solid) and homogenous (gas-gas) reactions that take place in the gasifier. A hydrodynamic model that describes the flow pattern of gases and solids in the gasifier is also included. A hydrodynamic model is required since the flow pattern of gases and solids affects the reaction rates and rates of heat and mass transfer in the gasifier. Reaction models (sub-models) are used to describe the rates of the heterogeneous and homogenous reactions. Reaction models have parameters that depend on the reactivity of the coal and the catalytic effect of the ash components in the coal. The model parameters can be determined by means of laboratory TG measurements and pilot-plant tests. The advantage of kinetic models is that the fixed-carbon conversion and the approach to equilibrium of the gas-gas reactions are calculated based on chemical kinetics and therefore no assumptions regarding output values are required.

Comprehensive simulation of fluidized and moving beds (CeSFaMB)

Several kinetic models have been developed for the fluidized-bed coal gasification process [2–10]. The model developed by De Souza Santos [10] is available commercially and was obtained by the CSIR and North West University under an academic licence. Development of the model started in 1987 at the University of Sheffield in the UK and is described in the PhD thesis of De Souza-Santos [11]. As detailed in the thesis, correlations are given to describe the rates of chemical reaction, hydrodynamics, mass transfer and heat transfer rates and elutriation. These are summarized Table 6.

Table 6. CeSFaMB sub-models and correlations

Sub-model	Correlation used by CeSFaMB
Hydrodynamics	
Minimum fluidizing velocity (U_{mf})	Wen and Yu [12]
Bubble diameter (d_b)	Mori and Wen [13]
Bubble rise velocity (m/s)	Davidson and Harrison [14]
Bubble fraction (-)	Davidson and Harrison [14]
Reaction rates	
Gas-solid (combustion and gasification)	Yoon *et al* [15], Johnson [16]
Gas combustion	Villenski and Hezeman [17]
Water gas shift	Franks [18]
Mass transfer coefficients	
Bubble - emulsion	Sit and Grace [19]
Emulsion- solid	La Nauze *et al* [20]
Heat transfer coefficients	
Bubble - emulsion	Kunii and Levenspiel [21]
Emulsion- solid	Kunii and Levenspiel [21]
Elutriation	Wen and Chen [22]

Simulation of the pilot-scale BFBG using CeSFaMB

The pilot-plant tests given in tables 4 and 5 were used to calibrate CeSFaMB for New Vaal and Grootegeluk coals respectively. Calibration of CeSFaMB for high-ash coal gasification involves adjusting the pre-exponential factors of four reactions:

$$C + H_2O \rightarrow CO + H_2 \quad (1)$$

$$C + CO_2 \rightarrow 2CO \quad (2)$$

$$C + 2H_2 \rightarrow CH_4 \quad (3)$$

$$CO + H_2O \rightarrow CO_2 + H_2 \quad (4)$$

The reaction model used by CeSFaMB for reactions (1) to (3) is—

$$\frac{dX}{dt} = k_{oi} \exp(\frac{-E_i}{RT}) F(X) P_j^{\ n} \tag{5}$$

for i = 1, 2 and 3.

$F(X)$ is the exposed core particle conversion model and P_j is the partial pressure of H_2O, CO_2 and H_2 in the gasifier.

The pre-exponential factors (k_{oi}) for reactions 1 to 4 were adjusted in order to obtain the best fit between the experimental and simulated results. For the activation energy (E_i) the default values calculated by CeSFaMB were used. For the combustion reaction the value calculated by CeSFaMB was also used since, at the conditions used for the tests (Table 3), the rate of this reaction is not dependent on the reactivity of the coal, but is limited by diffusion of oxygen to the coal particles.

The pre-exponential factors and the comparison between measured and simulated results for New Vaal and Grootegeluk coals are listed in tables 7 and 8.

Table 7. Pre-exponential factors and deviations for New Vaal coal

Test number	1	2	3	4	5	6
Pre-exponential factors (s^{-1})						
$C+H_2O \rightarrow CO + H_2$: k_{01}	15000	15000	15000	15000	15000	15000
$C + CO_2 \rightarrow 2CO$: k_{02}	200	200	200	200	200	200
$C + H_2 \rightarrow CH_4$: k_{03}	0.8E-7	0.3E-7	2.5E-7	5E-07	5E-07	0.1E-7
$CO + H_2O \rightarrow CO_2 + H_2$: k_{04}	80	65	80	80	80	80
Deviations between measured and simulated results [1]						
Mid-bed temperature (°C)	1	-42	-18	-20	19	-18
Gasifier exit temperature (°C)	-22	-27	-62	-6	-50	-9
Dry gas composition (vol. %)						
CO	-0.92	0.46	-0.9	-0.18	-3.2	-1.4
H_2	-6.47	-0.7	-1.5	-2.25	-4	-1.7
CH_4	1.03	-0.26	0.2	0.1	0.27	-0.68
CO_2	1.67	-0.5	0.6	0.05	2.02	0.7
Fixed carbon conversion (%)	-4.1	0.96	-0.55	-0.7	-4.45	-6.79

[1]A positive value indicates that the value predicted by CeSFaMB is higher than the measured value and a negative value indicates that the predicted value is lower than the measured value.

Tables 7 and 8 shows that, because of the higher reactivity of New Vaal coal, the reaction rate constants for reaction (1) and (2) are significantly higher than the values obtained for Grootegeluk coal. Owing to the higher volatile matter content of Grootegeluk coal more CH_4 is present in the syngas and higher values of k_{03} were required to fit the experimental data. The higher rate of the water gas shift (k_{04}) obtained for New Vaal coal could be attributed to better catalytic activity of the ash compared with Grootegeluk coal.

For New Vaal coal satisfactory deviation between measured and simulated results were obtained. However for Grootegeluk coal, CeSFaMB produced significant deviations for the gasifier exit temperature and the hydrogen in the gas. This could be attributed to the fact that for Grootegeluk coal, CeSFaMB produced higher elemental mass balance non-closures. For Grootegeluk coal the simulation results show that 13% less carbon leaves the gasifier with the gas and the ash in comparison to the carbon that enters the gasifier with the coal. For oxygen the elemental mass balance deviation is 5%, with more oxygen leaving the gasifier in comparison to the oxygen in the reactants and coal feed streams. The mass balance non-closures affect the gasifier syngas composition and the gasifier exit temperature.

Table 8. Pre-exponential factors and deviations for Grootegeluk coal

Test number	1	2	3	4	5
Pre-exponential factors (s^{-1})					
$C + H_2O \rightarrow CO + H_2$: k_{01}	49	49	49	49	49
$C + CO_2 \rightarrow 2CO$: k_{02}	6	6	6	6	3
$C + H_2 \rightarrow CH_4$: k_{03}	4E-07	4E-07	4E-07	1.5E-05	1E-06
$CO + H_2O \rightarrow CO_2 + H_2$: k_{04}	25	25	25	7	25
Deviations between measured and simulated results					
Mid-bed temperature (°C)	-21	9	21	-2	-38
Gasifier exit temperature (°C)	-116.3	-46	-43.3	-79	-136
Dry gas composition (vol. %)					
CO	1.7	-2	-0.4	0.8	1.6
H_2	-4.7	-7.7	-4.6	-8.9	-3.7
CH_4	0.1	0.1	-0.9	-0.3	-0.7
CO_2	-1.8	2.7	2.4	0.01	0.4
Fixed carbon conversion (%)	0.19	-1.78	0.99	-2.56	3.37

FLUIDIZED-BED GASIFIER SCALE-UP USING CESFAMB

A 15 MW thermal fluidized-bed coal gasifier was designed using CeSFAMB with New Vaal coal as feed. The model parameters used for the design were the average values obtained from the six pilot plant tests given in Table 7. The design and operation was optimized by using a circular and expanded bed (Figure 5), increasing the bed height and increasing the preheat temperature of the air, steam and oxygen to 450°C. An expanded bed was used to accommodate the increase in net gas flow produced by the gasification reactions. The reduced superficial gas velocity in the bed reduces the bubble size, which increases the rates of inter-phase heat and mass transfer. The bed height was increased to increase the residence time of char in the gasifier and the reactants were pre-heated in order to reduce the oxygen consumption. Input conditions are given in Table 9, which is a 100-times scale-up of the BFBG pilot plant.

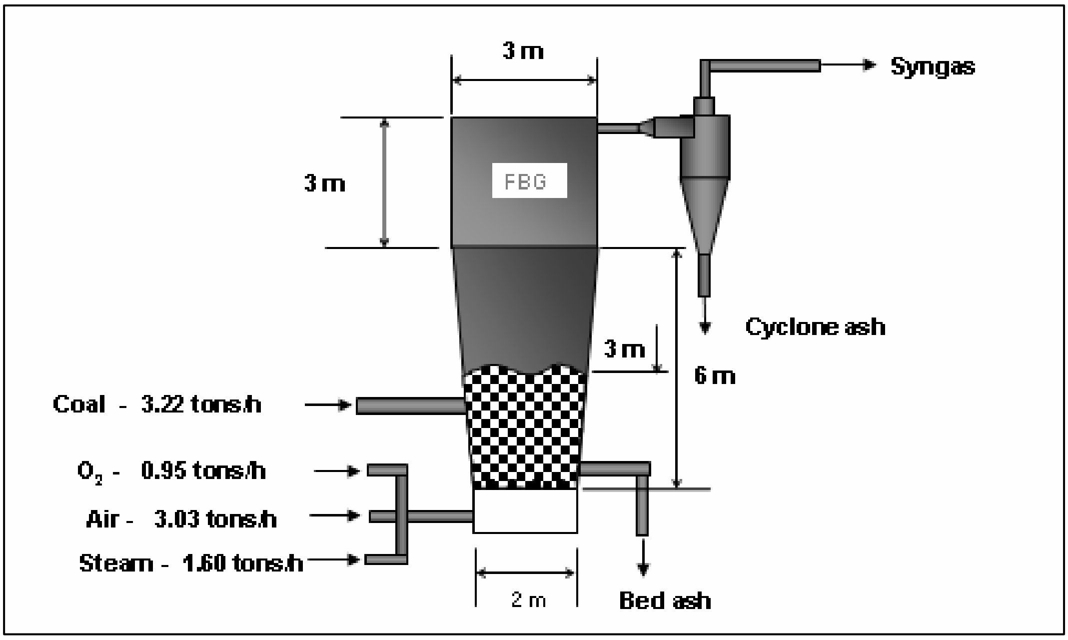

Figure 5. 15-MW fluidized-bed coal gasifier

Table 9. Input values for 15 MW thermal fluidized-bed coal gasifier design

New Vaal coal feedrate (tons/h)	3.22
Steam flow (tons/h)	1.60
Airflow (tons/h)	3.03
Oxygen flow (tons/h)	0.85
Reactant pre-heat temperature (°C)	450
Expanded bed height (m)	3.0
Bed diameter (m)	2.26
Bed section height (m)	6.0
Freeboard diameter (m)	3.38
Freeboard height (m)	3.0
k_{01} (s^{-1})	15000
k_{02} (s^{-1})	200
K_{03} (s^{-1})	77.5
K_{04} (s^{-1})	3.1E-07

Table 10. CeSFaMB output values for the 15 MW thermal fluidized-bed coal gasifier

Mid-bed temperature (°C)	948
Gasifier exit temperature (°C)	931
Dry gas flow (Nm^3/h)	6024
CO (%)	19.80
H_2 (%)	24.23
CH_4 (%)	1.2
CO_2 (%)	18.62
Calorific value of dry gas (MJ/Nm^3)	6.2
Superficial bed velocity (m/s)[1]	1.84
Fixed carbon conversion	98.2
Cold gas efficiency (%)[2]	78.2

[1] At the middle of the bed
[1] Superficial gas velocity = Total syngas flowrate (m^3/s)/bed area (m^2)
[2] Based on energy in coal

A higher fixed carbon conversion is predicted by CeSFaMB for the 15 MW gasifier because of the increased residence time of char (90 minutes), compared with 30 minutes in the pilot-scale BFBG. Increased residence times are achieved because of the increased bed height and lower void fraction of the bed. The void fraction of the pilot-scale BFBG bed was high because of the slugging behaviour of the bed. Slugging occurs when the bubble diameter increases to 60% of the bed diameter. Figure 6 shows that the predicted bubble diameter for the 15 MW BFBG is less than 60% of the bed diameter for the entire height of the bed. CeSFaMB assumes that a single ash stream leaves the gasifier with a carbon content equal to the carbon content of the bed, which could result in an over-prediction of the fixed-carbon conversion when the residence times are extrapolated significantly beyond the value at which CeSFaMB was calibrated.

Owing to thermal fragmentation and attrition of char particles, a fluidized-bed coal gasifier produces bed ash and fly-ash (elutriated char). The higher carbon content of the fly ash (Table 5) reduces the fixed carbon conversion of the gasifier.

Figures 7 and 8 show the concentration profiles of CO_2, CO, O_2, H_2O, H_2 and CH_4 in the bed and freeboard of the 15-MW gasifier. Figure 7 shows that most of the O_2 in the reactant gas is consumed in the lower 15% of the bed. When O_2 has been depleted the concentrations of CO and H_2 increase owing to gasification reactions. The water-gas shift reaction is the dominating reaction occurring in the freeboard of the gasifier.

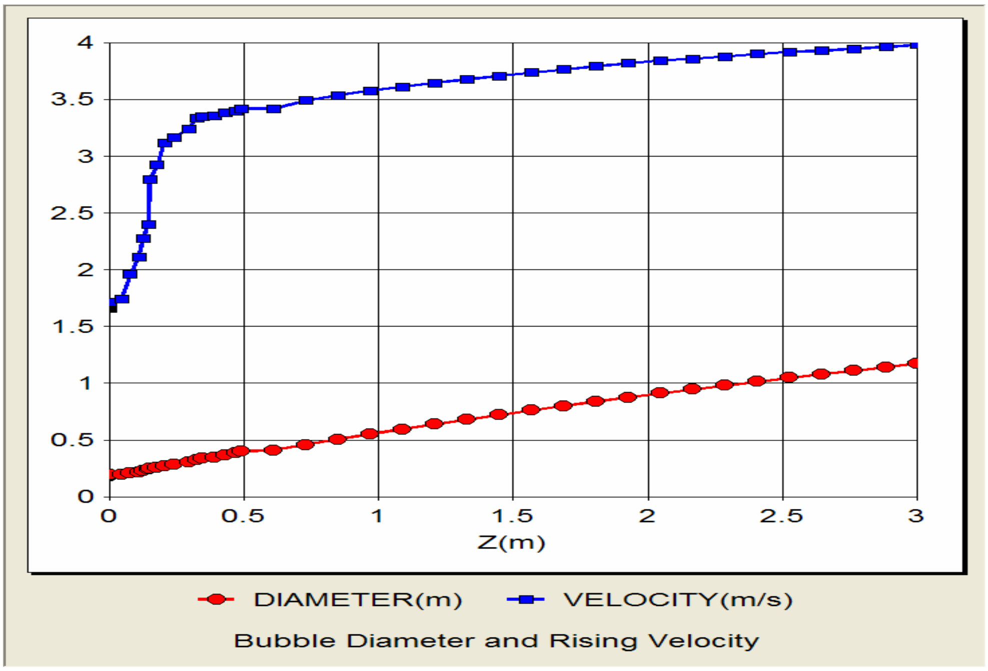

Figure 6. Bubble diameter and rising velocity as a function of bed height

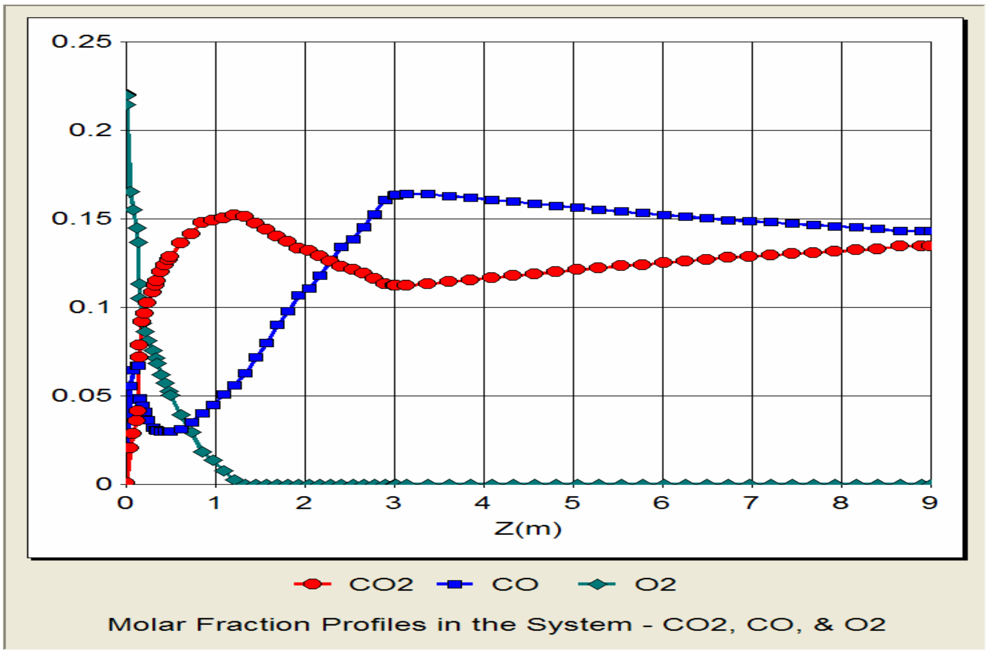

Figure 7. Mole fractions of CO_2, CO, O_2 as a function of gasifier height

CONCLUSIONS

Two high-ash South African coals were successfully gasified in a pilot-scale fluidized-bed gasifier. Compared with air-blown gasification, oxygen-enrichment of the gasification air results in a significant increase in the calorific value of the gas.

The fixed carbon conversion in the fluidized-bed gasifier increases with an increase in coal reactivity, temperature and residence time of char particles in the gasifier.

A fluidized-bed coal gasification simulation model (CeSFaMB) was calibrated using the pilot-scale fluidized-bed gasifier test results. Satisfactory agreement was obtained between measured and simulated results for New Vaal coal. Grootegeluk coal, however, produced significant deviations between the measured and simulated gasifier exit temperature and the gas calorific value. This was attributed to higher elemental mass balance non-closures produced by CeSFaMB in the case of Grootegeluk coal.

The simulation model predicts that a significant increase in performance of the fluidized-bed gasifier can be achieved for a scaled-up 15-MW gasifier compared with the pilot plant. This is possible because of an increase in residence time of char and the absence of bed slugging.

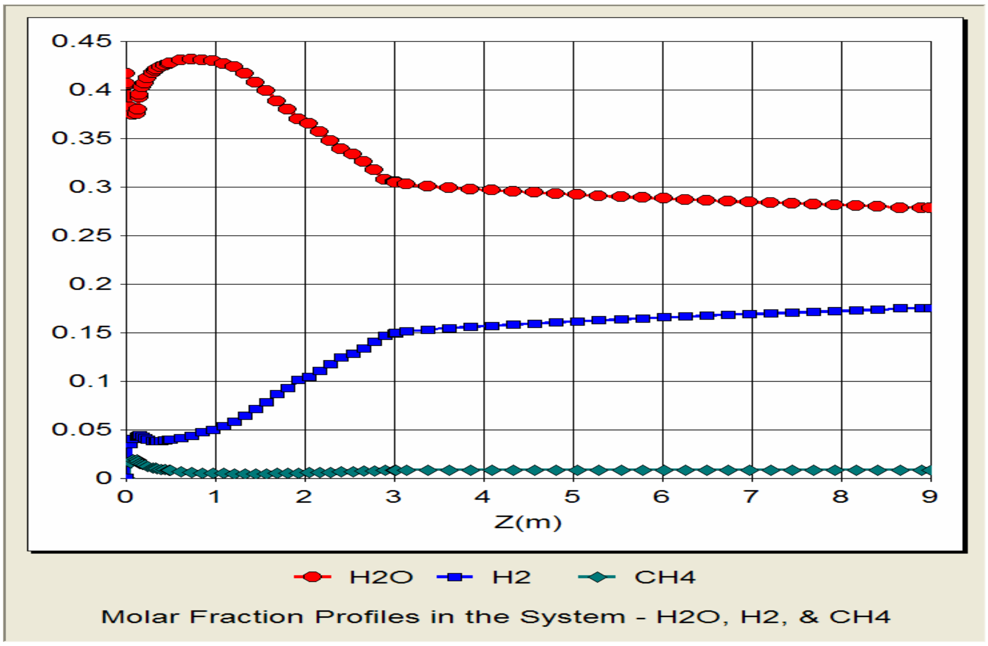

Figure 8. Mole fractions of H_2O, H_2 and CH_4 as a function of gasifier height

NOTATION

d_b	bubble diameter, m
d_{50}	mean particle diameter, mm
d_p	particle size, mm
k_{0i}	pre-exponential factor of reaction i, s^{-1}
E_i	Arrhenius activation energy of reaction i, $kJ.mol^{-1}$
N	reaction order, –
P_j	partial pressure of reactant j, atm
R	universal gas constant, 8.314 $J.mol^{-1}.K^{-1}$
U	superficial gas velocity in the bed, $m.s^{-1}$
U_{mf}	minimum fluidizing velocity, $m.s^{-1}$
U_t	terminal falling velocity, $m.s^{-1}$
X	fractional conversion of fixed carbon in coal, –
$Z(m)$	height above distributor, m

ACRONYMS/ABBREVIATIONS

BFBG	Bubbling Fluidized-bed Gasifier
CeSFaMB	Comprehensive Simulation of Fluidized and Moving Beds
C_2^+	Ethane and higher hydrocarbons
CSIR	Council for Scientific and Industrial Research
FD	Forced draught
ID	Induced draught

IGCC Integrated Gasification Combined Cycle
LPG Liquefied Petroleum Gas
MW Megawatt
TG Thermogravimetry

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