Further Experiments on Sewage Sludge Air Gasification: Influence of the Non-Stationary Period on the Overall Results

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Abstract

In the present study, the gasification with air of dried sewage sludge (DSS) was experimentally investigated using a bubbling fluidized bed reactor. The experimental work analyzes the effect on the DSS gasification tests of the gas residence time (as a function of both the bed height and the superficial air velocity) and the potential catalytic role of the DSS ashes. Furthermore, the influence of the non-stationary period on the accumulated tar production was analyzed. The results show that a longer gas residence time improves the gas efficiency process. A decrease in tar production and an increase in H_2 content in the raw gas were observed when a mixture of sand and DSS ashes was used as an initial bed. These results confirm the catalytic effect of DSS ashes, which would be able for catalytic applications. The influence of the non-stationary period on the accumulated tar production was minimized when the experiment length was extended. This finding could confirm the higher tar production during this period.

Keywords

Sewage Sludge; Air Gasification; Bubbling Fluidized Bed; Tar Content, Ash

Introduction

In the European Union (EU), production of sewage sludge has been increased as a consequence of implementation of new urban waste-water treatment plants, in agree with the UWWTD Directive (91/271/EEC).[1] It is mentioned also that is necessary to ensure that sludge could be disposed of safely in an environmentally acceptable manner. Sewage sludge contains pathogens, heavy metals and poorly biodegradable organic compounds (persistent organic pollutants) that make difficult their management possibilities.[2]

The main disposal routes are currently application to agricultural land, landfill and incineration. Other ways used in the past, such as the disposal to surface waters, have been phased out as an answer to Directive (91/271/EEC).

Agricultural use is a way for recycling because of it contains of organic substances and nutrimental elements of plants.[2,3] Nevertheless, the EU by means of Directive (86/278/EEC)[4] prohibits the use of untreated sludge on agricultural land unless it is injected or incorporated into the soil. The EU set also limits for concentrations of heavy metals in soil. Applying sewage sludge to soil might provide metals in potentially toxic, and may originate problems of bio-accumulation in plants, animal or man.[2,5] Many countries limit the amount of sewage sludge applied as a precautionary measure.[6]

Another possibility is landfilling, which might be the least expensive way to management sludge; nevertheless the Landfill Directive (99/31/EEC)[7] introduces the reduction of biodegradable municipal waste to landfill.

Incineration reduces the volume of sludge but has the serious disadvantage of popular discord.

To manage the important amount of sludge that would be produced in the next years, it is necessary to develop alternative processes for treatment of this waste material. It is necessary to raise alternatives which eliminate bacteria and viruses, simultaneously that management the sewage sludge in a safe way for the environment.

Gasification presents the advantages of destruction of pathogenic bacteria and volume reduction.[8] This process produces gas with a certain heating value that is enough to generate heat or electricity when burnt.[9] The main drawback to date for this technology is the high tar and dust content of the synthesis gas produced.[10] The relative importance of this problem depends on the final application of the producer gas during the gasification process. If the producer gas is just burned to provide heat, the gas clean-up requirements may be minimal. However, high technology systems, such as gas turbines, require much cleaner fuel gases.[9] Tar can condensate, giving rise tar aerosols, and polymerized to form more complex structures, causing problems in process equipment as well as in engines and turbines used in application of producer gas.[11] Gas used in internal combustion engine operation has been postulated as less than 100 mg/Nm^3 for tar concentration.[12] Another author mentioned a limit of 50 mg/Nm^3 when gas is used in gas turbine cogeneration facility in order to prevent compressor problems[9].

In case of sewage sludge, its high ash content could be an advantage for tar removal. Many of gasification reactions are catalyzed by alkali metals salts (specially those containing calcium, sodium, iron, and potassium) which are excellent promoters of gasification reactions.[8,13,14] In fact, high-ash fuels could be related to catalytic tar cracking.[10,15,16] Sewage sludge char presents higher reactivity in comparison with coal chars. One reason could be the catalytic effect of the ash, which contain large quantities of iron oxides.[17,18] Sewage sludge char is composed of carbon, distributed on a matrix of ash impregnated with catalytically active metals.[19]

In the present study, the gasification with air of dried sewage sludge (DSS) was experimentally investigated using a bubbling fluidized bed reactor. Results from a previous work,[20] performed using the same experimental system, have been encouraging for the application of bubbling fluidized bed gasification with air as an option for energy recovery from sewage sludge. Nevertheless, the high tar content in the raw gas needs to be reduced substantially. In the mentioned study some unexpected trends related to tar generation as a function of the equivalence ratio were found. As has been mentioned in previous works, the high ash content of sewage sludge could represent an obstacle to the gas diffusion[21] (a higher gas residence time could be required to achieve a higher conversion efficiency), at the same time that could constitute an advantage in terms of tar reduction.[10,16,21] This contradictory effect could be a possible explanation for results previously found. In addition, the potential influence of the non-stationary period on the experimental results could reduce reliability of the tar production findings.

As a continuation of the initial study[20] and in order to explain the results previously obtained, the present experimental work analyze the effect on the DSS gasification tests of the gas residence time (as a function of both the bed height and the air superficial velocity) and the potential catalytic role of the DSS ashes.

Experimental Section

BFB Reactor Facility. Experiments were carried out in a laboratory scale plant operating at atmospheric pressure, feeding sewage sludge and air continuously, and with a non-stop ash removal system. The experimental equipment and experimental procedure were based on those used in a previous study[20] although in the current work the following differences were included:

- A heated filter system was used to separate the fine particles from the sample raw gas. The ceramic fiber thimble was located inside a stainless steel tube. The filter system was placed between the reactor outlet and the condensers. The operating filter temperature was fixed at 673 K in order to prevent the tar build-up.[22]

- During the experimental test, the bed height was kept at 300 mm by means of a concentric pipe twice larger than the used in the previous work.[20]

Taking into account that the ratio length/diameter of the fluidized bed reactor has a value of 7.9 (L = 300 mm; D = 38.1 mm), slugging problems could be expected. However, significant variations of bed pressure drop and temperature during all experiments performed in this study were not observed. Moreover, the experimental carbon conversion efficiencies obtained could confirm that the fluidization regime was maintained.

A flow sheet of the experimental system (including the modifications mentioned above) is shown in Figure 1.

Materials Analysis. The samples of dry sewage sludge (DSS) used in the study were obtained as a dried, granulate product from an urban waste-water treatment plant. The sludge was previously treated by anaerobic digestion and thermal drying. Proximate and ultimate analyses of the received sewage sludge are shown in Table 1. The lower heating value (LHV) was also been determined using a calorimeter IKA A-2000 (standard procedure: ISO-1928-89). The value obtained for LHV was 10.26 MJ/kg. All analyses were performed by the “Instituto de Carboquímica” (CSIC, Zaragoza, Spain). The dried sewage sludge was crushed and sieved to provide a feed sample in the size range of 250–500 μm.

Silica sand was used as the initial bed material. Additionally and in order to improve the fluidization, a determined amount of silica sand (20% of the DSS flow rate) was continuously fed to the reactor system. The sand used in both cases was sieved in the range of 250–350 μm.

For specific experiments, a mixture of sand and DSS ashes was used (70% sand and 30% ash) as initial bed material. The mixture was obtained by calcinations of char samples (at 1123 K during 2 hours) produced in previous gasification tests. For this material, the size was delimited in the range 250–500 μm. The DSS ashes produced from oxidation of char (obtained in gasification tests) were analyzed for minor elements (Al, Ca, Fe, K, Mg, Na, Ti) using inductively coupled plasma-atomic emission spectroscopy (ICP-AES). Furthermore, a sample of original DSS ashes was analyzed by the same analytical method. The solutions needed to carry out the analysis were prepared beginning with the addition of HNO_3, HCl and HF. The mixture was heated in a microwave. At last, H_3BO_3 were added in order to reach the neutralization of HF. Analysis of minor elements contents in both DSS ash samples are shown in Table 2.

Experimental Procedure. The laboratory experiments are grouped into four sections:

- Section I: Experiments performed in order to check the influence of the bed height.
- Section II: Experiments to analyze the effect of the superficial air velocity.
- Section III: Gasification experiments using a mixture of ash and sand as initial bed material.
- Section IV: Experiments performed during 89 minutes.

Experiments related to sections III and IV have been performed in order to check the catalytic effect of the DSS ashes. Until the moment at which the stationary state is reached, the DSS ashes proportion in the bed is variable. This proportion increases with the course of the time, becoming every time more important the catalytic effect of the

DSS ashes. The initial bed (for experiments corresponding to section III) was obtained from oxidation of char resulted in previous tests.

Table 3 shows the scheduled operating conditions for all experiments performed.

All the experiments were performed at the same temperature, 1123 K, and at three different values of the equivalence ratio, λ (25%, 30%, and 35%), defined as the ratio between the actual flow rate of the air and the stoichiometric flow rate required for fuel combustion. The value of λ was changed fitting the feed rate of solid in some cases or the air flow rate in other cases.

The reproducibility of experimental results obtained in this laboratory scale-plant was analyzed in a previous work.[20]

The average experiment length was 45 min, except for tests performed in section IV (89 min).

Results and Discussion

The results presented in this paper show the influence of the operating conditions on gasification efficiency. The DSS gasification performance has been evaluated for the total experimental time elapsed for each test. Tables 4–6 report measured and calculated data for each experiment performed: experimental equivalence ratio, experimental DSS flow rate, average gas composition (dry basis), specific yield to gas obtained (y_{gas}), lower heating value of the product gas (LHV), average cold gas efficiency defined as the ratio of the LHV of the produced gas to the LHV of the DSS fed (η)[†], average percentage of carbon in the biomass recovered in the gas (y_{carbon}), and mass products distribution. Products distribution is shown based on the biomass fed and also on the

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[†] The average cold gas efficiency is an estimated value because the external heat provided by the electric furnace is not considered for calculations. In this way, the η values are only appropriate for determining the gasification yield.

DSS weight plus the oxygen mass fed during the experiment, giving an experiment closure balance of nearly 100%.

Effect of the bed height

In this section, the results[20] obtained from experiments 1–3 were compared with those performed by Manyà et al. at the same conditions, except for the bed height value (150 mm).

Figure 2 displays the average cold gas efficiency, the feed carbon to gas conversion, and the specific yield to gas obtained as a function of equivalence ratio and bed height. From analyzing the results, it is reasonable to assume that an increase of the bed height improves the efficiency of the DSS gasification process. One possible explanation for these results could be related to the fact that reactivity of carbonaceous particles appreciably decreases at a late stage of their conversion, as remaining carbon is "diluted" in the ashes and internal diffusion of gaseous reactants becomes more difficult. In this sense, the high ash content of DSS represents an obstacle to gas diffusion; as a consequence, the intrinsic kinetics of the particle conversion could be lower than for low-ash fuels. In conclusion, a longer gas residence time would be required for the complete conversion of fuel particles in the bed.

Regarding the quality of the produced gas, Figure 3 shows the average gas composition of the exit gas (in a dry basis) as a function of bed height. The evolution of the LHV of the product gas is also displayed in Figure 4. The increase of LHV with bed height (observed for each value of equivalence ratio analyzed) can be attributed to steam reforming (generation of H_2 and CO) and steam cracking reactions (generation of H_2 and light hydrocarbons). In this sense, results displayed in Figure 3 (including the decrease of water content as bed height increases) support this argument.

Figure 5 shows tar production (expressed as percentage of DSS fed) as a function of both the bed height and equivalence ratio values. The unexpected high value of tar production at $\lambda = 35\%$ (especially for the bed height value of 150 mm) can be related to the fact that an increase of the experimental value of λ implies a reduction of the solid flow rate (for experiments performed in this section). Thus, a decrease of the DSS flow rate reduces the amount of ash, which could promote catalytic cracking of tar species[10,15] in the regions of the bed and freeboard (see Table 2 for details concerning DSS ashes composition). In other words, a lower DSS flow rate value could delay the steady-state achievement in the bed (replacement of initial sand for a mixture of sand and char with a high percentage of ash).

On the other hand, the observed increase of tar production for experiments performed at a higher bed height (except for run performed at $\lambda = 35\%$) could be explain by an increase in the time needed to attain the bed stabilization. During this period of time (non-stationary period), the tar released could be very high due to the small quantity of ashes deposited into the bed and, consequently, the reported value (accumulated tar production) corresponding to a given experiment can be overestimated.

Effect of the superficial air velocity

Assuming that an increase of the bed height has been proven effective, it is interesting to analyze the influence of the superficial air velocity on the gasification process of sewage sludge. In this way, Table 4 reports the results obtained from experiments 1–8, performed at different values of u_f.

Figure 6 displays the values obtained for the average cold gas efficiency, the feed carbon to gas conversion, and the lower heating value of produced gas as a function of the ratio u_f/u_{mf} for different values of λ. The explanation of observed trends from the data is not an easy task. The results were different depending on the equivalence ratio value. For experiments conducted at λ = 25%, all parameters analyzed, including the tar production (see Table 4), indicate an improvement when reducing the ratio u_f/u_{mf} from 8 to 6.5. This expected effect would be attributed to an increase in the gas residence time, as it was happening when the bed height was increased. However, this trend was not observed for the rest of gasification tests performed at different equivalence ratio values. In fact, and unexpectedly, an improvement in the gasification results as the superficial air velocity was increased can be deduced for experiments performed at λ = 35% (see Figure 6). This apparent contradiction could be related to the DSS flow rate value. As it has been stated previously, a lower value of solid flow rate could delay the steady-state achievement in the bed. For this reason, results obtained from tests carried out at low values of Q_{DSS} (high values of λ and/or low values of u_f) could be altered significantly because of high time required to achieve the bed stabilization.

Figure 7 compares the tar production obtained for experiments performed using different ways to set the values of λ: fixing the value of Q_{DSS} (3.75 g/min) or fixing the value of Q_{air} (3.10 Ndm^3/min). As can be deduced from Figure 7, experimental results (in terms of tar generation) measured at a fixed value of DSS flow rate seem to be better than those obtained fixing the air flow rate value. This fact confirms that the effect of a lower solid flow rate is very relevant for these 45 min-long experiments. Consequently, the influence of gas residence time needs to be evaluated minimizing the effect of Q_{DSS}.

Effect of the use of DSS ash as initial bed material

In this section, experiments performed using —as initial bed material— a mixture of ashes (obtained from combustion of char) and sand are analyzed. The composition of the initial bed was fixed at approximately 70% sand and 30% ash. Table 5 shows the results obtained from experiments 9–11. These experiments were carried out at u_f/u_{mf} = 6.5.

Figure 8 compares the products distribution results obtained from experiments 9–11 with those obtained from experiments performed at the same conditions except for the initial bed material (tests 4–6). Overall, the effect of the use of DSS ash as initial bed material on the products distribution was moderate. Nevertheless, a significant decrease in tar production was observed for experiments performed at higher values of λ. This result could be related to the Q_{DSS} value and its influence on the steady-state achievement. In this sense, the decrease in tar content (using DSS ashes as initial bed material) becomes more pronounced as the Q_{DSS} value is decreased. A similar behavior was observed when analyzing the evolution of the average cold gas efficiency as a function of the equivalence ratio value and the initial bed material (see Table 5): the increase of the η value (by addition of DSS ashes in the initial bed) was more significant for experiments performed at higher values of λ.

Special attention was focused on the effect of the DSS calcined ashes on the evolution of the mean gas composition as a function of the equivalence ratio. From results shown in Figure 9, it can be observed a significative increase in H_2 composition and a subsequent decrease in light hydrocarbons composition when DSS ashes were used in the initial bed. This effect could be explained by the assumption that the ashes promote the reforming reactions. The catalytic potential of calcined DSS ashes would be able to allow their use as a bed additive and/or any other catalytic application.

Figure 10 displays the time evolution of hydrogen and ethylene composition during test 5 (initial bed: sand) and test 10 (initial bed: ash/sand), both experiments performed at $\lambda = 30\%$. The analysis of dynamic response suggests that the use of DSS ashes as initial bed material does not reduce the non-stationary period (the trends observed from Figure 10 were very similar for the rest of tests of this section). This finding could be due to lower DSS ashes weight percentage in the initial bed (30%). The estimation of the ash composition equilibrium value is approximately a 60% of ashes in the bed (taking into account the fresh silica sand flow rate and the fixed carbon quantified from the final bed). This value is substantially higher that the one corresponding to the bed initially charged.

Effect to set the experiment length to 89 min

Results obtained from experiments performed at 89 min (three gasification tests carried out at u_f/u_{mf} = 6.5) have been satisfactory in terms of bed and process stabilization.

For these experiments, the initial bed material was sand and, consequently, the evolution of char (carbon plus ashes) concentration in the bed could vary with time in an exponential manner to a final equilibrium value. However, this assumption has not been validated experimentally, because of the system's incapability to evaluate the bed composition by instant values. Nevertheless, the operational time of 89 min may be enough in order to reach a higher time proportion of "stabilized" experiment.

As a consequence of extending the experiment length, the effect of the non-stationary period (basically as a function of the Q_{DSS} value) becomes diluted. In this way, the comparison of gasification results obtained under different conditions can be more accurate.

Figure 11 shows the products distribution results obtained from tests 12–14. Results obtained from experiments 4–6 (performed at 43 min) are also displayed for comparison. A significant decrease in tar production was obtained (for all values of λ) when the operational time was extended. This finding could confirm the higher tar production during the non-stationary period. As can be deduced from results shown in Table 6, tar decrease was accompanied by an increase in H_2 composition (except for experiments performed at λ = 35%, where average hydrogen content remained unchanged) and a subsequent decrease in light hydrocarbons content as a consequence of the major role played by the reforming reactions.

Data shown in Figure 12 illustrate the importance of reducing the results uncertainty related to the non-stationary period. For experiments performed at 89 min, the evolution of tar production as a function of equivalence ratio can be considered as expected.

Conclusions

This work complements a previous study where the influence of temperature, equivalence ratio, and fluidizing velocity over sewage sludge gasification in a bubbling fluidized bed were analyzed. In that work some unexpected results related to tar generation as a function of the equivalence ratio were found.

At the present study, the bed height has been analyzed resulting that a longer gas residence time improves the gas efficiency process. Nevertheless, as in the mentioned previous work, some unexpected results have been found when low Q_{DSS} are used. In these cases, a lower DSS flow rate value could delay the stationary state achievement in bed (replacement of initial sand for a mixture of sand and char with a high percentage of ash, which could promote catalytic cracking of tar).

Contradictory results (linked also to lower DSS flow rate values) have been found when the superficial air velocity has been analyzed. This finding shows the importance of minimizing the influence of Q_{DSS} in order to provide results more accurate. For this reason, a mixture of sand and ash (obtained by calcination from DSS char) has been used as initial bed material. In this case, the potential catalytic effect of DSS ashes has been present from the beginning of gasification. As it was expected, a decrease in tar production and an increase in H_2 production have been observed. This fact confirms the catalytic effect of calcined DSS ashes, which would be able for catalytic applications (further studies in this subject will be interesting). Nevertheless, a decrease of the non-stationary period length has not been observed when DSS ashes are used as initial bed. This behavior could be due to the fact that the initial bed of sand and ashes doesn't represent the real composition of the bed when the stationary period has been reached (when the stationary period is achieved, some carbonaceous particles are present into the bed).

The influence of the non-stationary period over the experimental tar production has been diluted when the experiment length has been extended. This finding could confirm the higher tar production during this period. For experiments performed at 89 min, the evolution of tar production as a function of equivalence ratio can be considered as expected.

In this work is shown that a disadvantage of sewage sludge, such as their high ash content, could become in an advantage by their catalytic effect.

In order to make future investigations, the influence of the non-stationary period over the experimental results must be quantified. Thus, the results concerning to DSS gasification performance would be more accurate.

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Nomenclature

BFB = bubbling fluidized bed

DSS = dried sewage sludge

daf = dry and ash free basis

LHV = lower heating value of the produced gas, dry basis

Ndm^3 = litter, normal conditions (273 K, 101 kPa)

Q_{air} = air flow rate used in an experiment, Ndm^3/min

Q_{DSS} = mass flow rate of sewage sludge, g/min

$Q_{DSSreal}$ =experimental mass flow rate of sewage sludge, g/min

T = bed temperature, K

u_f = superficial gas velocity at the inlet of the gasifier bed, cm/s

u_{mf} = minimum fluidization gas velocity (gasifier bed conditions), cm/s

y_{carbon} = average percentage of carbon in the biomass recovered in the gas

y_{gas} = specific yield to gas obtained, (Ndm^3 of dry gas/kg of daf DSS fed)

Greek Symbols

η = average cold gas efficiency based on LHV of gas and biomass, %

λ = equivalence ratio, defined as the air-to-fuel ratio used in the reactor divided by the air-to-fuel ratio for the stoichiometric combustion, %

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Table 1. Analysis of dried sewage sludge samples

Proximate	Analytical standard	% by weight
Moisture	ISO-589-1981	8.92
Ash	ISO-1171-1976	42.12
Volatiles	ISO-5623-1974	42.30
Fixed Carbon	By difference	6.66

Ultimate (organic fraction)	Analytical instrument	% by weight (daf)
Carbon	Carlo Erba 1108	55.33
Hydrogen	Carlo Erba 1108	6.70
Nitrogen	Carlo Erba 1108	8.15
Sulfur	Carlo Erba 1108	1.75
Oxygen	By difference	28.07

Table 2. Ash analysis

	SA[‡](mg/g ash)	SB[§](mg/g ash)
Ca (mg/g)	112.96	117.65
Al (mg/g)	92.33	94.62
Fe (mg/g)	42.91	54.09
K (mg/g)	24.39	29.98
Mg (mg/g)	23.46	19.44
Na (mg/g)	9.78	12.09
Ti (mg/g)	5.42	6.36

[‡] Ash in received DSS

[§] Ash resulting from oxidation of char produced in gasification

Table 3. Summary of operating conditions for the experiments

Test no	Temperature[**] (K)	Initial bed material	Test duration (min)	λ (%)	Q_{DSS} (g/min)	Q_{air} ($\mathrm{Ndm^3/min}$)	u_f/u_{mf}
1	1123	sand	43	25	4.50	3.75	8.0
2	1123	sand	43	30	3.75	3.75	8.0
3	1123	sand	43	35	3.23	3.75	8.0
4	1123	sand	43	25	3.75	3.10	6.5
5	1123	sand	43	30	3.12	3.10	6.5
6	1123	sand	43	35	2.67	3.10	6.5
7	1123	sand	43	35	3.75	4.34	9.3
8	1123	sand	43	30	5.20	5.18	11.0
9	1123	ash-sand	43	25	3.75	3.10	6.5
10	1123	ash-sand	43	30	3.12	3.10	6.5
11	1123	ash-sand	43	35	2.67	3.10	6.5
12	1123	sand	89	25	3.75	3.10	6.5
13	1123	sand	89	30	3.12	3.10	6.5
14	1123	sand	89	35	2.67	3.10	6.5

[**] Temperature of the fluidized bed. For all experiments performed in this study, the temperature set-points corresponding to the freeboard and cyclone were 873 K and 623 K, respectively.

Table 4. Experimental results of gasification tests (section I and II)

Run no.	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**
u_f/u_{mf}		8			6.5		9.3	11
λ (%)	24.40	30.70	34.90	23.80	28.60	32.80	34.10	29.70
$Q_{DSSreal}$ (g/min)	4.64	3.69	3.29	3.93	3.27	2.85	3.84	5.26
Q_{air} (Ndm^3/min)	3.75	3.75	3.75	3.10	3.10	3.10	4.34	5.18
Average gas composition (% vol.dry basis)								
H_2	8.78	6.43	4.91	9.06	6.86	4.96	4.91	5.71
O_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
N_2	59.70	63.31	66.05	58.09	62.06	65.61	65.34	64.21
CO	10.20	8.99	8.47	10.61	9.84	8.90	8.91	9.15
CH_4	2.91	2.66	2.30	2.87	2.68	2.23	2.28	2.53
CO_2	15.42	15.85	15.77	16.32	15.84	15.92	16.04	15.65
C_2H_4	2.43	2.23	2.04	2.52	2.23	1.96	2.03	2.26
C_2H_6	0.17	0.14	0.11	0.18	0.14	0.11	0.11	0.13
C_2H_2	0.13	0.13	0.13	0.13	0.14	0.12	0.15	0.15
H_2S	0.26	0.26	0.23	0.22	0.22	0.19	0.23	0.22
y_{gas} (Nm^3/kg of daf DSS)	2.10	2.49	2.71	2.10	2.37	2.57	2.69	2.37
LHV (kJ/Nm^3)	4886	4255	3766	5020	4418	3749	3821	4176
η (%)	51.0	52.7	50.8	53.0	52.1	47.9	50.9	49.4
y_{carbon} (%)	73.1	76.5	80.3	73.1	76.3	79.1	82.0	72.7
Product distribution based on the sludge fed (% weight)								
Produced Gas	58.84	60.17	63.81	57.20	59.02	64.16	64.79	57.99
Tar	7.81	6.41	5.73	6.62	6.91	5.59	5.20	7.33
Char	43.09	42.79	45.81	45.04	44.26	46.25	44.65	45.14
Water	14.35	16.73	14.97	11.02	13.38	12.77	14.49	13.75
Product distribution based on the sludge and oxygen fed (% weight)								
Gas	47.36	46.1	47.37	46.25	45.94	48.38	48.39	44.76
Tar	6.28	4.91	4.26	5.35	5.38	4.22	3.88	5.65
Char	34.68	32.79	34.00	36.41	34.45	34.87	33.35	34.84
Water	11.55	12.82	11.12	8.91	10.42	9.63	10.82	10.61
Mass balance closure	99.87	96.62	96.75	96.92	96.19	97.10	96.44	95.86

Table 5. Experimental results of gasification tests (section III)

Run no.	**9**	**10**	**11**
u_f/u_{mf}		6.5	
λ (%)	24.50	29.10	34.30
$Q_{DSSreal}$ (g/min)	3.83	3.21	2.73
Q_{air} (Ndm^3/min)	3.10	3.10	3.10
Average gas composition (% vol.dry basis)			
H_2	9.01	8.29	6.06
O_2	0.00	0.01	0.01
N_2	58.83	61.51	65.38
CO	9.75	9.61	8.89
CH_4	2.73	2.48	1.96
CO_2	16.84	15.67	15.57
C_2H_4	2.34	2.01	1.78
C_2H_6	0.19	0.15	0.09
C_2H_2	0.10	0.09	0.08
H_2S	0.22	0.19	0.19
y_{gas} (Nm^3/kg of daf DSS)	2.14	2.43	2.69
LHV (kJ/Nm^3)	4732	4322	3628
η (%)	50.0	52.3	48.6
y_{carbon} (%)	72.9	75.5	79.5
Product distribution based on the sludge fed (% weight)			
Produced Gas	59.25	59.08	65.06
Tar	6.81	5.33	4.75
Char	45.42	46.65	45.81
Water	10.73	12.50	15.59
Product distribution based on the sludge and oxygen fed (% weight)			
Gas	47.66	45.81	48.53
Tar	5.48	4.13	3.54
Char	36.54	36.17	34.17
Water	8.64	9.69	11.63
Mass balance closure	98.32	95.80	97.87

Table 6. Experimental results of gasification tests (section IV)

Run no.	**12**	**13**	**14**
u_f/u_{mf}		6.5	
λ (%)	24.20	29.30	33.80
Q_{DSSreal} (g/min)	3.87	3.20	2.77
Q_{air} (NL/min)	3.10	3.10	3.10
Average gas composition (% vol.dry basis)			
H_2	9.88	7.84	4.63
O_2	0.00	0.00	0.00
N_2	57.70	61.32	65.74
CO	10.68	9.29	7.89
CH_4	2.76	2.33	2.12
CO_2	16.05	16.61	17.16
C_2H_4	2.36	2.08	1.96
C_2H_6	0.16	0.12	0.12
C_2H_2	0.13	0.12	0.11
H_2S	0.28	0.28	0.27
y_{gas} (Nm³/kg of daf DSS)	2.16	2.45	2.65
LHV (kJ/Nm³)	4970	4226	3545
η (%)	53.1	51.5	46.5
y_{carbon} (%)	76.5	81.7	84.0
Product distribution based on the sludge fed (% weight)			
Produced Gas	57.99	62.98	65.12
Tar	6.03	5.33	4.65
Char	45.49	44.88	48.21
Water	10.99	10.37	13.17
Product distribution based on the sludge and oxygen fed (% weight)			
Gas	46.75	48.78	48.74
Tar	4.86	4.13	3.48
Char	36.76	34.76	36.23
Water	8.86	8.03	9.85
Mass balance closure	97.23	95.70	98.30

Caption for Figures

Figure 1. BFB reactor system.

Figure 2. Average cold gas efficiency (η), feed carbon to gas conversion (y_{carbon}), and specific yield to gas obtained (y_{gas}) as a function of equivalence ratio and bed height (■,η; ○, y_{carbon}; ▼, y_{gas}).

Figure 3. Water production (+) and mean gas composition as a function of equivalence ratio and bed height (■, H_2; ●, CO; ▲, CH_4; ▽, C_2H_4).

Figure 4. Evolution of LHV as a function of bed height for the different equivalence ratio values studied.

Figure 5. Tar production as a function of bed height and equivalence ratio.

Figure 6. Average cold gas efficiency (η), feed carbon to gas conversion (y_{carbon}) as a function of equivalence ratio and u_f/u_{mf} (■,η; ○, y_{carbon}; ✳, LHV).

Figure 7. Tar production as a function of equivalence ratio and operational procedure.

Figure 8. Products distribution as a function of equivalence ratio and initial bed material (■, tar; ○, water; ▲, char; ▽, gas).

Figure 9. Mean gas composition as a function of equivalence ratio and initial bed material (■, H_2; ●, CO; ▲, CH_4; ▽, C_2H_4).

Figure 10. Evolution of H_2 (a) and C_2H_4 (b) composition during run 5 (initial bed: sand) and run 10 (initial bed: ash and sand).

Figure 11. Products distribution as a function of equivalence ratio and experiment length (■, tar; ○, water; ▲, char; ▽, gas).

Figure 12. Tar production as a function of equivalence ratio, experiment length and initial bed material for experiments performed at $u_f/u_{mf} = 6.5$.

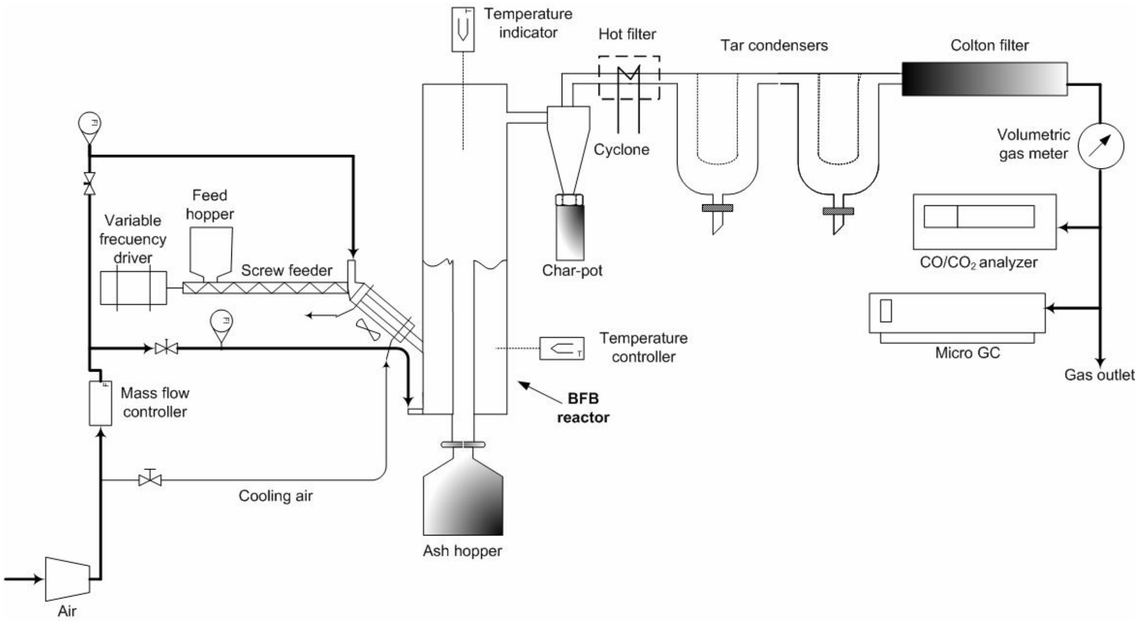

Temperature indicator
Hot filter
Tar condensers
Colton filter
Volumetric gas meter
Cyclone
Feed hopper
Variable frecuency driver
Screw feeder
Char-pot
CO/CO$_2$ analyzer
Micro GC
Temperature controller
Gas outlet
Mass flow controller
BFB reactor
Cooling air
Ash hopper
Air

Figure 1

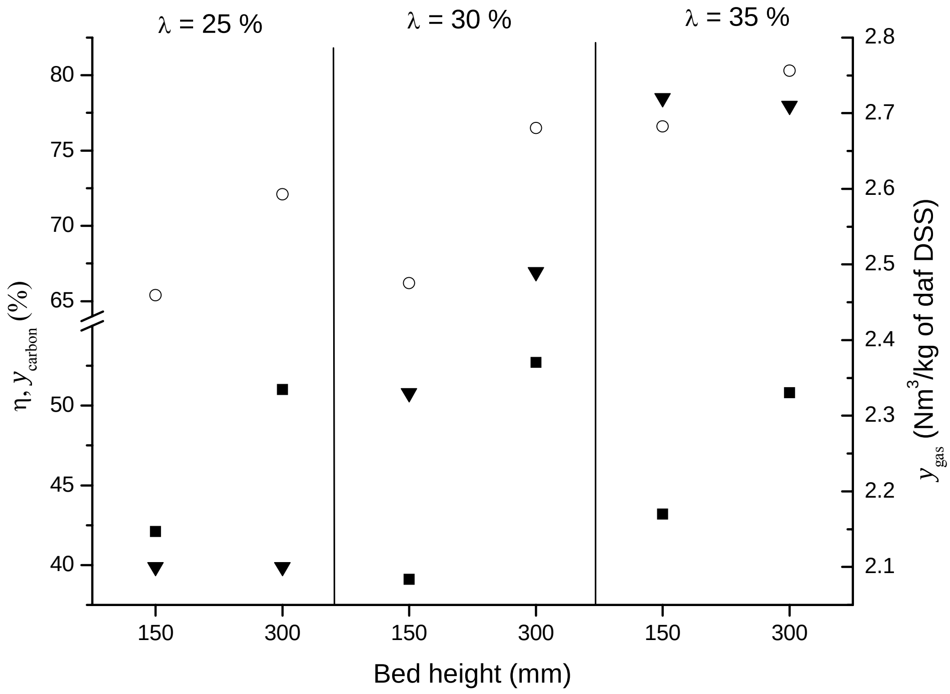

λ = 25 %
λ = 30 %
λ = 35 %
80
75
70
65
50
45
40
2.8
2.7
2.6
2.5
2.4
2.3
2.2
2.1
η, y_carbon (%)
y_gas (Nm³/kg of daf DSS)
150
300
150
300
150
300
Bed height (mm)

Figure 2

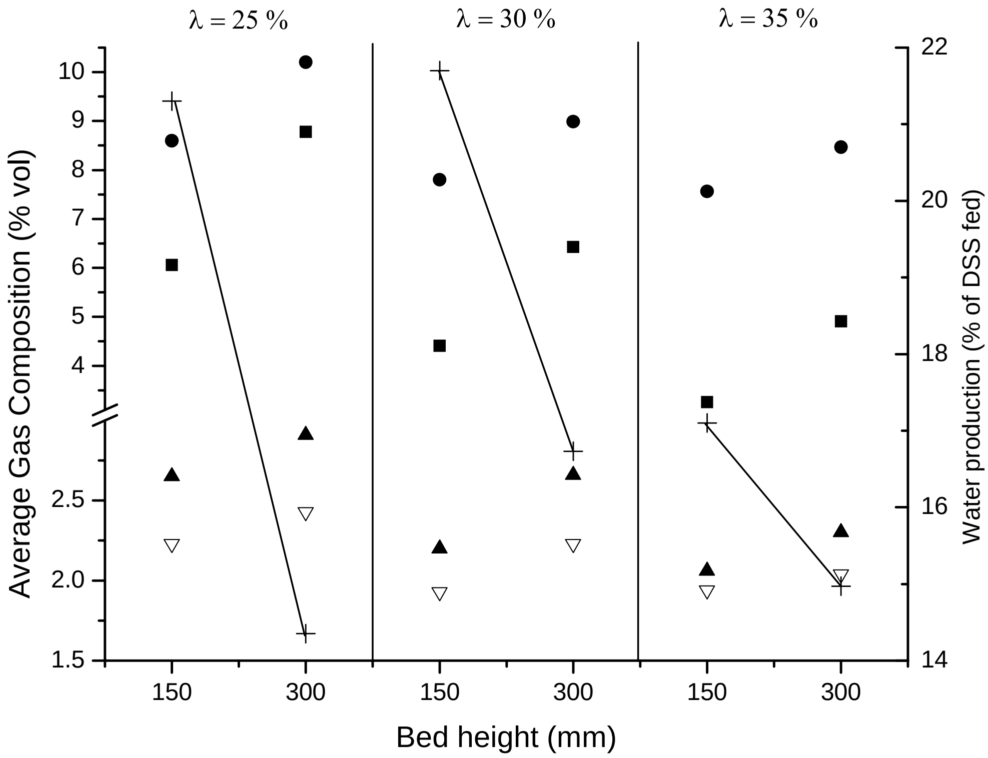

λ = 25 %
λ = 30 %
λ = 35 %
Average Gas Composition (% vol)
Water production (% of DSS fed)
Bed height (mm)
10
9
8
7
6
5
4
2.5
2.0
1.5
22
20
18
16
14
150
300
150
300
150
300

Figure 3

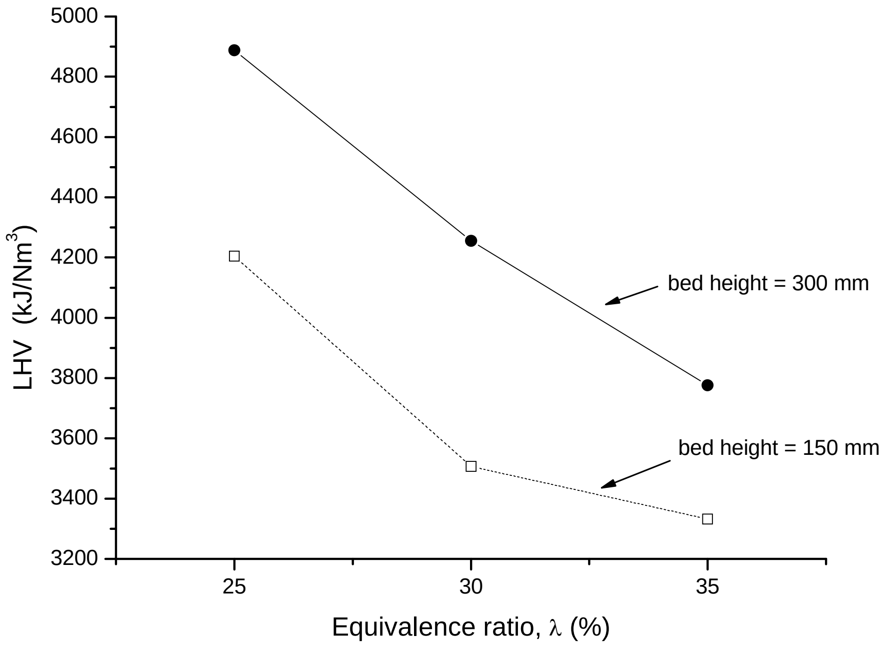

5000
4800
4600
4400
4200
4000
3800
3600
3400
3200
LHV (kJ/Nm3)
25
30
35
Equivalence ratio, λ (%)
bed height = 300 mm
bed height = 150 mm

Figure 4

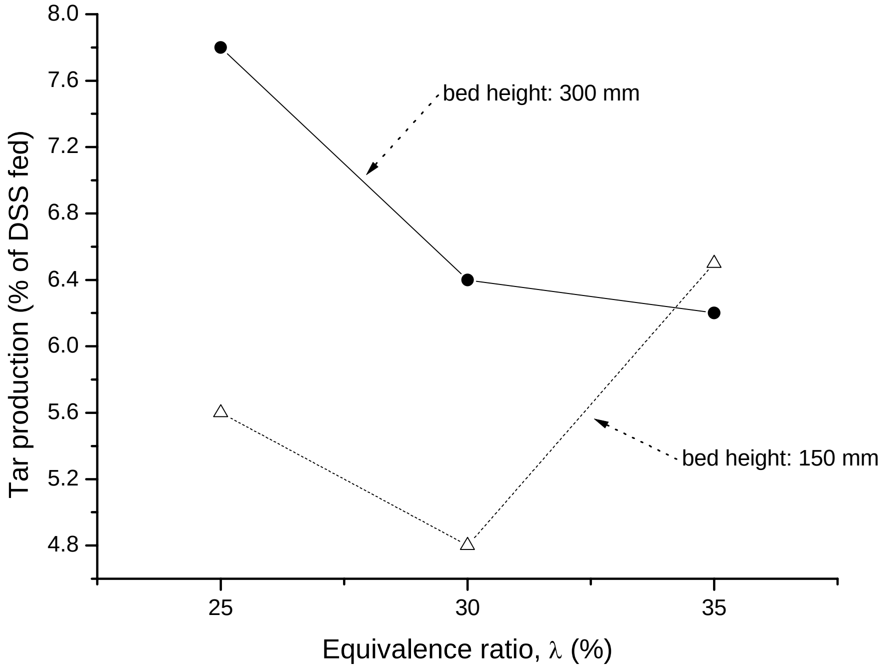

8.0
7.6
7.2
6.8
6.4
6.0
5.6
5.2
4.8
Tar production (% of DSS fed)
25
30
35
Equivalence ratio, λ (%)
bed height: 300 mm
bed height: 150 mm

Figure 5

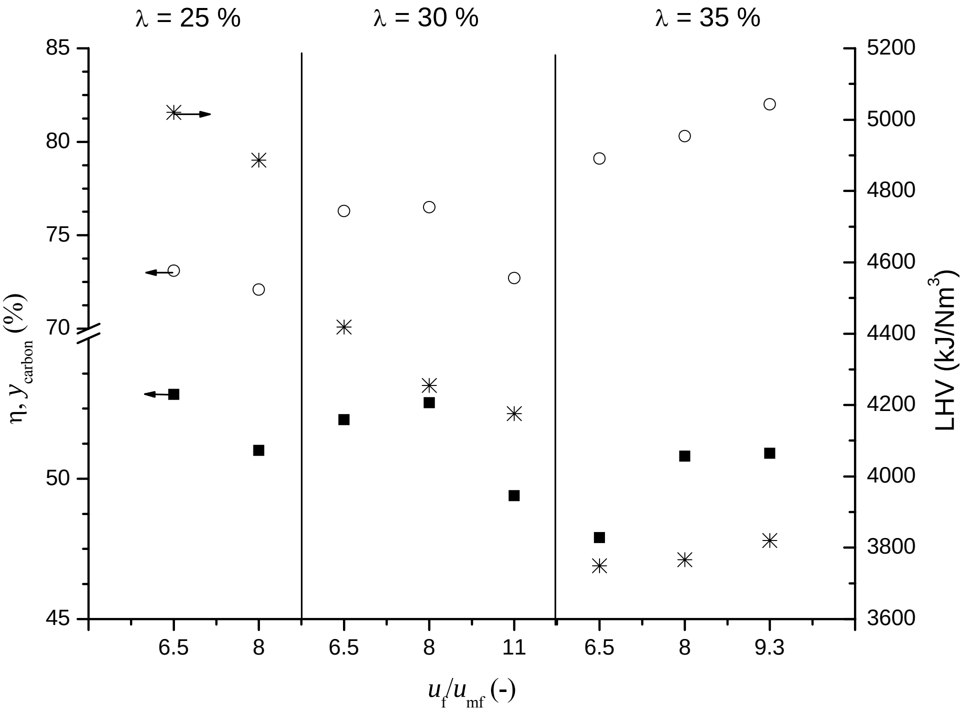
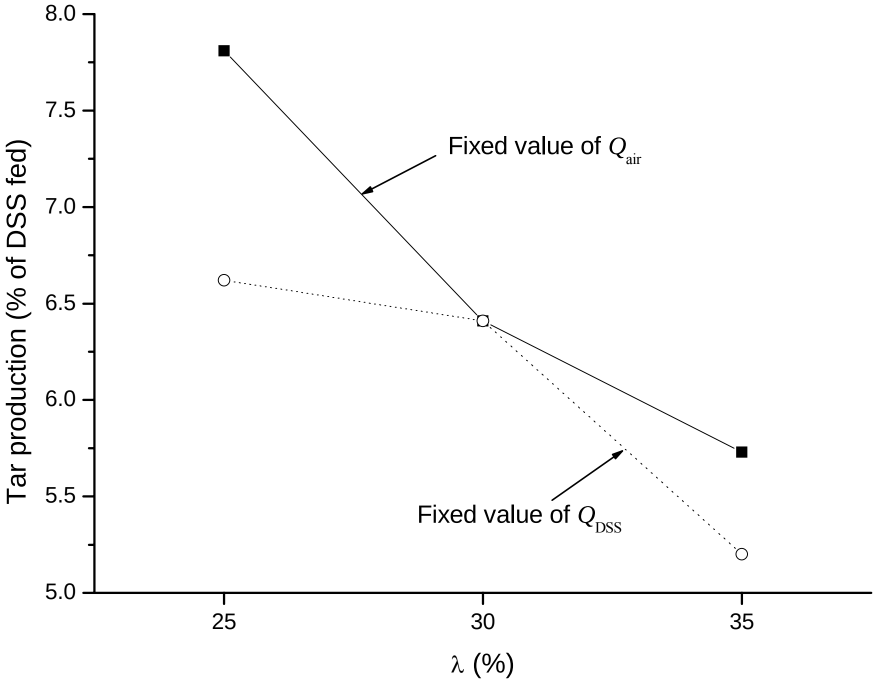

λ = 25 %
λ = 30 %
λ = 35 %
85
80
75
70
50
45
η, y_{carbon} (%)
5200
5000
4800
4600
4400
4200
4000
3800
3600
LHV (kJ/Nm^3)
6.5
8
6.5
8
11
6.5
8
9.3
u_f/u_{mf} (-)

Figure 6



8.0
7.5
7.0
6.5
6.0
5.5
5.0
Tar production (% of DSS fed)
Fixed value of Q_{air}
Fixed value of Q_{DSS}
25
30
35
λ (%)

Figure 7

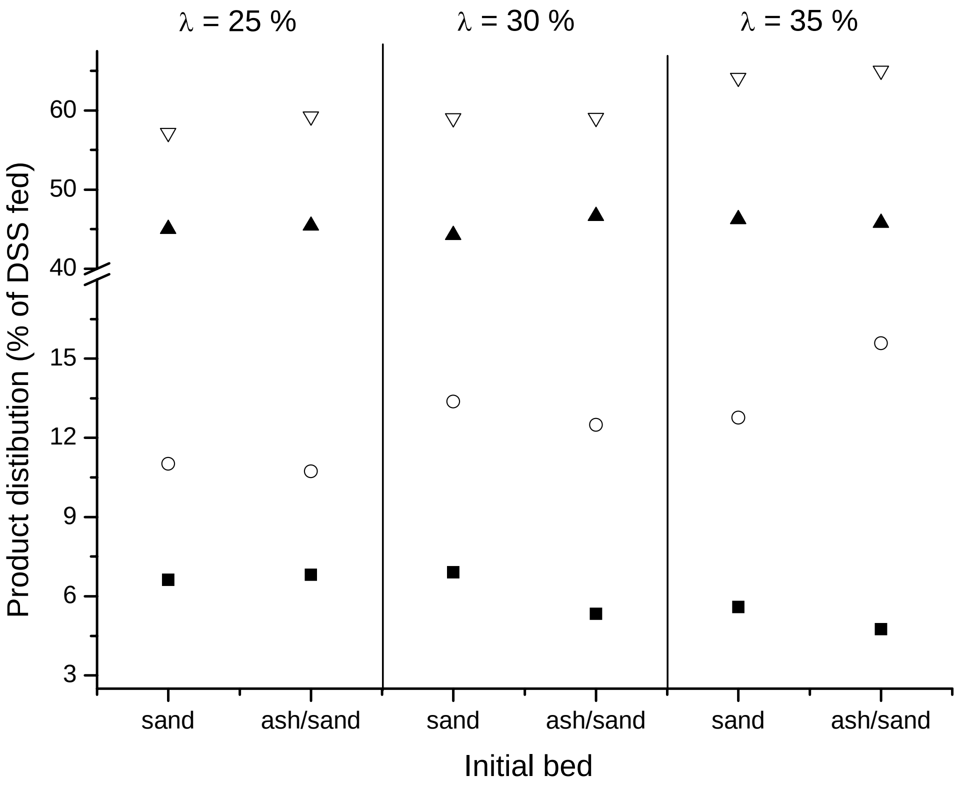

λ = 25 %
λ = 30 %
λ = 35 %
Product distibution (% of DSS fed)
60
50
40
15
12
9
6
3
sand
ash/sand
sand
ash/sand
sand
ash/sand
Initial bed

Figure 8

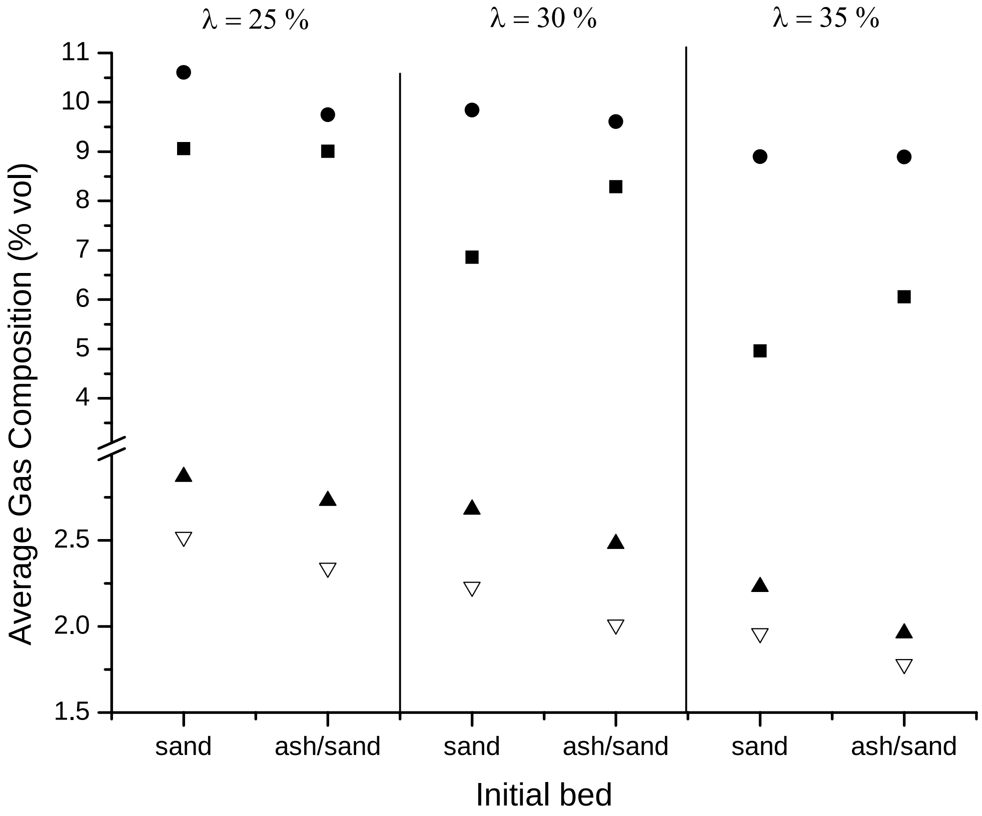

λ = 25 %
λ = 30 %
λ = 35 %
Average Gas Composition (% vol)
11
10
9
8
7
6
5
4
2.5
2.0
1.5
sand
ash/sand
sand
ash/sand
sand
ash/sand
Initial bed

Figure 9

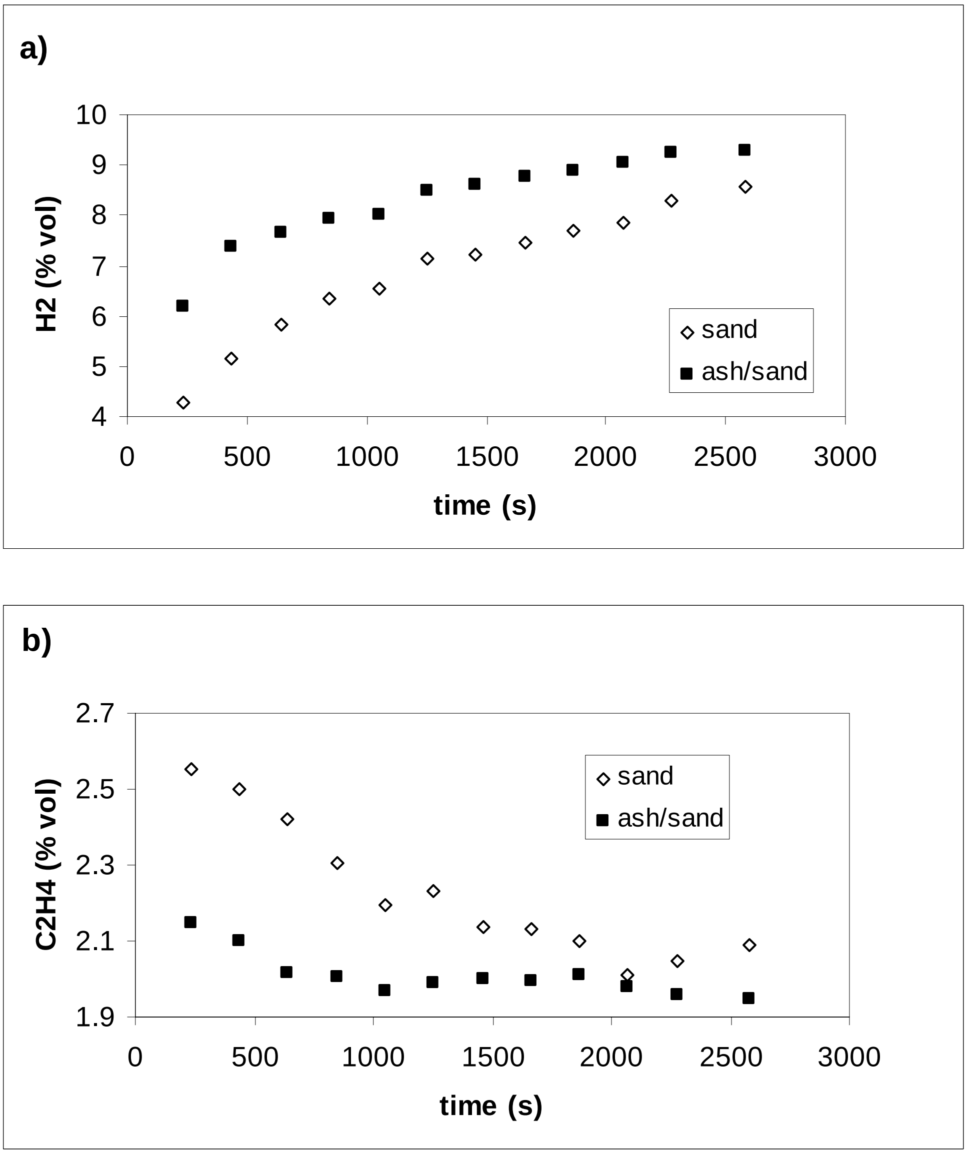

a)
H2 (% vol)
10
9
8
7
6
5
4
0
500
1000
1500
2000
2500
3000
time (s)
◇ sand
■ ash/sand
b)
C2H4 (% vol)
2.7
2.5
2.3
2.1
1.9
0
500
1000
1500
2000
2500
3000
time (s)
◇ sand
■ ash/sand

Figure 10

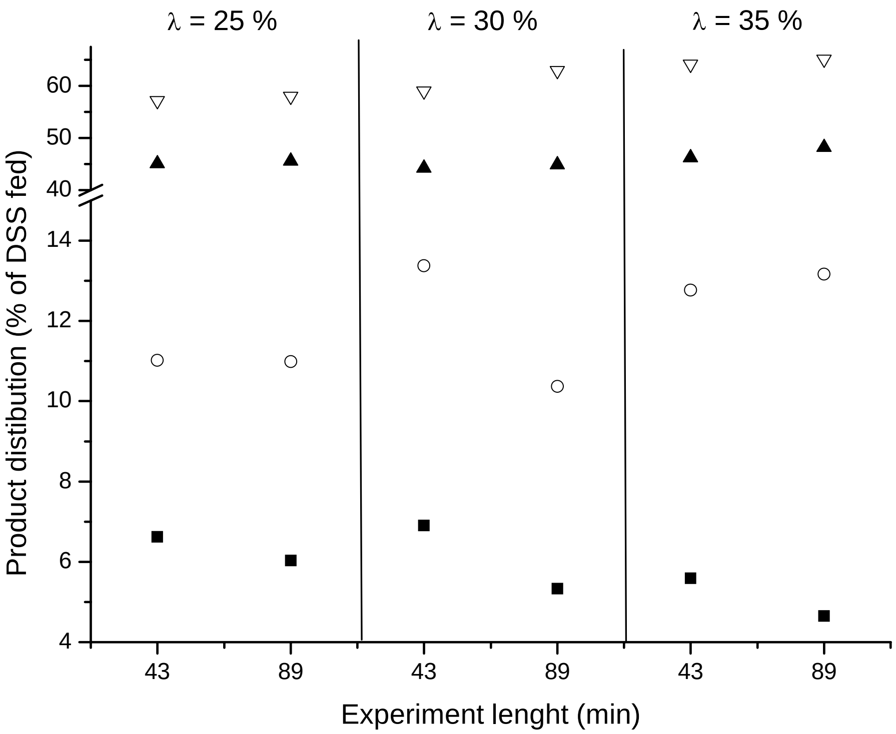

λ = 25 %
λ = 30 %
λ = 35 %
Product distibution (% of DSS fed)
60
50
40
14
12
10
8
6
4
43
89
43
89
43
89
Experiment lenght (min)

Figure 11

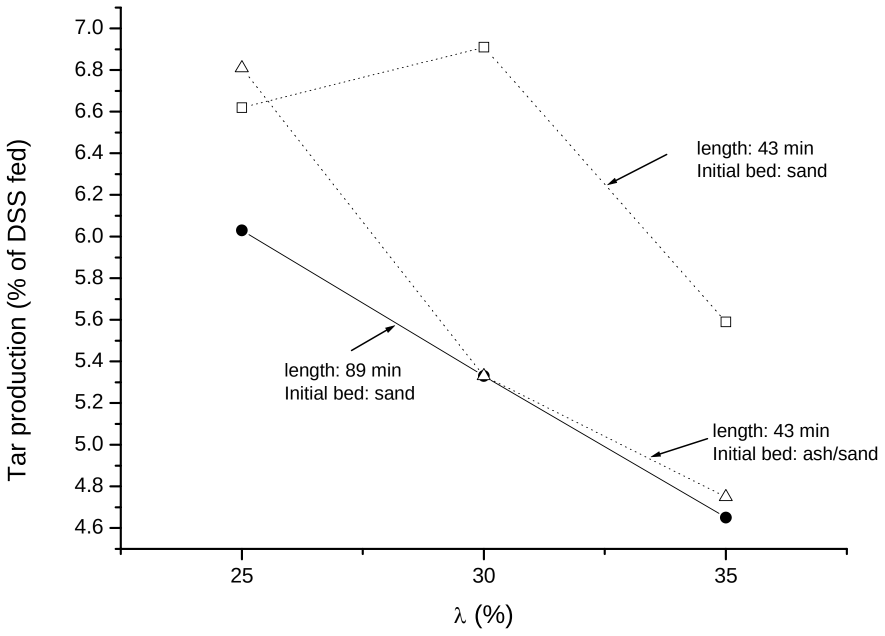

Tar production (% of DSS fed)
7.0
6.8
6.6
6.4
6.2
6.0
5.8
5.6
5.4
5.2
5.0
4.8
4.6
25
30
35
λ (%)
length: 43 min
Initial bed: sand
length: 89 min
Initial bed: sand
length: 43 min
Initial bed: ash/sand

Figure 12