Air-steam gasification of sewage sludge in a bubbling bed reactor: Effect of alumina as a primary catalyst

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ARTICLE INFO

ABSTRACT

Numerous references can be found in scientific literature regarding biomass gasification. However, there are few works related to sludge gasification. A study of sewage sludge gasification process in a bubbling fluidised bed gasifier on a laboratory scale is here reported. The aim was to find the optimum conditions for reducing the production of tars and gain more information on the influx of different operating variables in the products resulting from the gasification of this waste. The variables studied were the equivalence ratio (ER), the steam-biomass ratio (SB) and temperature. Specifically, the ER was varied from 0.2 to 0.4, the SB from 0 to 1 and the temperature from 750 °C (1023 K) to 850 °C (1123 K). Although it was observed that tar production could be considerably reduced (up to 72%) by optimising the gasification conditions, the effect of using alumina (aluminium oxide, of proven efficacy in destroying the tar produced in biomass gasification) as primary catalyst in air and air-steam mixture tests was also verified. The results show that by adding small quantities of alumina to the bed (10% by weight of fed sludge) considerable reductions in tar production can be obtained (up to 42%) improving, at the same time, the lower heating value (LHV) of the gas and carbon conversion.

1. Introduction

Sewage sludge is the liquid or semi-liquid waste generated in wastewater treatment plants. The amount of sludge produced in Spain and Europe has increased in last years. This increase is expected to continue [1] as a result of population growth and more stringent quality standards for waste waters [2].

The most widely used management alternatives for sewage sludge are agricultural use, incineration and landfilling, but in recent years, pyrolysis, gasification and wet oxidation have generated growing interest [3,4]. European legislation lays down restrictions for tipping organic material in landfills [5]. There is a strong social opposition to waste incineration and there are several constrains in the EU to alternative waste management systems such as direct use for cultivation [6].

Sewage sludge gasification is a technology that leads to a reduction in the volume of waste, the elimination of pathogens and fixes the heavy metals, thereby meeting the requirements imposed by the social and legislative framework [7]. During the process, gases are produced that can be used to generate energy or used as raw material in chemical synthesis processes. The substoichiometric conditions of the process reduce the formation of nitrogen oxides and sulphur oxides, heavy metal emissions, flying ash, dioxins and furans [4]. However, one of the main problems to be solved is tar production.

Tars are complex mixtures of condensable hydrocarbons, aromatic compounds, oxygen-containing hydrocarbons and polycyclic aromatic hydrocarbons (PAH) [8]. These compounds can condense in pipes and filters and obstruct them, can affect the fuel line and the injectors of internal combustion engines and condense in the compressors or transfer lines, making the use of gases impossible for many applications [9,10].

Tar elimination can be achieved by different methods, such as optimising the conditions of gasification [8], using physical systems (scrubbers, filters, cyclones, etc. [11]), thermal cracking [12,13] or catalytic cracking [14]. Tar removal technologies can be classified into primary (when carried out inside the gasifier) or secondary (when carried out outside the gasifier) [8].

A large amount of research has studied the effect of the various primary methods on the composition of the products of biomass gasification on a fluidised bed, specifically focusing on tar production [9,15–18]. These works have demonstrated the performance of calcined dolomite and olivine as catalysts for eliminating tar in gasification with air [15], steam [16] and steam-O_2 mixtures [9]. Although olivine is more resistant to wear than calcined dolomite in fluidised bed processes, calcined dolomite appears to be more active in the destruction of tars [17,18].

The effect of secondary methods of tar reduction in biomass gasification has also been studied [19–21]. A comparison has been made of the capacity of nickel and calcined dolomite, calcite and magnesite catalysts for cleaning raw hot gas from biomass steam gasification [19]. With a three-stage system comprising a fluidised bed biomass gasifier with air, a fixed bed reactor of calcined dolomite, and

a commercial nickel catalyst reactor, tar concentrations of 1–2 mg/ Nm^3 have been obtained [20]. Tar reductions of up to 88% have also been obtained in gases from a biomass fluidised bed gasifier with air using four different types of calcined dolomite. In these experiments it was seen how the reductions increased on increasing Fe_2O_3 content in the catalyst [21]. Although the catalytic action increases with the calcination of the catalyst, the materials become more friable [22].

This paper deals with the effect of some gasification parameters, such as temperature, ER and SB, on sewage sludge gasification products. The results analysis has focused on the composition, production and LHV of the gases, on tar production, cold gas efficiency and carbon conversion. The results were compared with those existing in the bibliography for biomass gasification in order to know if the different characteristics of these materials have any influence on the products obtained, since the properties of the fuel play an important part during the conversion process [23].

Besides this work, the effect of in-bed use of alumina was also tested. The catalytic activity of this material (similar to that of dolomite) is related to its complex mixture of aluminium, oxygen and hydroxyl ions that produces both acid and base sites [24]. His porous structure increases the residence time of the tars in the bed by promoting the cracking and steam reforming reactions of the tars [25,26]. As the main innovation of this work, a set of tests with alumina were carried out using not only air but also air-steam mixtures as gasifying agents.

2. Materials and methods

2.1. Materials

The dried sludge samples came from urban wastewater treatment plants and consisted of spherical aggregate of approximately 2–5 mm diameter. The water content, ash content and other characteristics of the sludge are shown in Table 1. The lower heating value (LHV) of the sludge was 11.5 MJ/kg and was calculated from the modified Dulong's formula. The sludge was crushed and sieved to obtain a particle of between 300 and 500 μm, the same as for the alumina on tests with catalyst. The alumina (Spheralite 505) was supplied by Axens Procatalyse Catalysts & Adsorbents with specifications as described in Manyà et al. [26]. Silica sand was used as bed material.

2.2. Laboratory scale plant

Fig. 1 shows a diagram of the installation where the experiments were conducted. The reactor used was a stainless steel (AISI 316 L) fluidised bed reactor with a total height of 700 mm and an inner diameter of 32 mm. There is a freeboard at the top of the reactor with

Table 1
Characterisation of sludge from wastewater treatment plant (analysis in triplicate).

Parameter[a]		Sludge[b]	Analytical method
Moisture (%)		6.4–7.5	UNE-EN 12880-2001
Organic mat. (%)		53.3–58.7	UNE-EN 12879-2001
Ash (%)		46.7–41.3	UNE-EN 12879-2001
pH		6.9	UNE-EN 12176-1998
Total carbon (%)		25.9–28.7	Elementary micro analyser
Nitrogen (%)		3.7–4.5	LECO CHNS-932
Hydrogen (%)		4.8	
Sulphur (%)		0.9	
Heavy metals (mg/kg)	Cd	1–4.4	UNE-EN 13346-2001
	Cu	387–418	
	Ni	52–64	
	Pb	93–226	
	Zn	1172–1283	
	Hg	1.3–4.2	
	Cr	128–199	

[a] Dry basis values, except moisture and pH.
[b] Value interval for three analytical assays.

an inner diameter of 46 mm. The bed height is kept at 100 mm by a concentric pipe (12.7 mm outer diameter) which goes through the distributor plate (0.1 mm pore size) and lets the overflowing material be collected and stored in a discharge tank. Some stainless steel balls were placed under the distributor plate to pre-heat the gasifying agent before it reached the bed. The sludge was fed by a dosing system comprising a hopper, a controlled screw feeder and a launch screw. The launch screw was inserted into the reactor a few millimetres above the distributor plate by a 12.7 mm outer diameter pipe. This pipe was water-cooled to prevent pyrolysis of the sludge before entering the reactor [27]. The assembly (bed and freeboard) were heated by an electric furnace. The air entering the reactor was regulated using a mass flow controller. By means of two rotameters, part of the gasification air was diverted to the screw feeder to help the sludge enter the reactor while the rest was introduced through the distributor plate. In the air-steam mixture tests, a peristaltic pump was used to introduce water into the reactor.

In order to remove the particles entrained by the gases produced, the equipment was fitted with a cyclone and a micronic filter. Both devices were placed inside a hot box (250 °C; 523 K) to prevent condensation of the tars. The gas leaving the hot box was cooled in 5 condensers containing isopropanol placed over an ice bath (following a similar system to that for tar protocol, CEN/TS 15439:2006 [28]), most of the tars and water being collected. Placed behind the condensers was a water filter, a silica gel filter and a cotton filter to complete the cleaning of the gases. Gas production was measured by a mass flow meter. The gas composition was determined by means of a micro gas chromatograph (Micro-GC, Varian CP-4900) connected at the end of the process line to know the percentage volume of N_2, O_2, H_2, CO, CO_2, CH_4, C_2H_6 and C_2H_4. The time interval between analyses was 5 min. To measure tar production the isopropanol-tar solutions were distilled to eliminate the absorbent (isopropanol). After distillation, the residue (tars) was dried at room temperature until constant weight. Finally, the sample was weighed. The char content was determined according the method used by Rapagná et al. [16].

2.3. Experimental conditions

The tests have been planned to find out the influence of the following parameters: temperature, equivalence ratio, steam and presence of alumina.

- Temperature. Tests were performed at temperatures of 750 °C (1023 K), 775 °C (1048 K), 800 °C (1073 K), 825 °C (1098 K) and 850 °C (1123 K).
- Equivalence ratio (ER). The ER, defined as the ratio between the flow rate of the air introduced into the reactor and the stoichiometric flow rate of the air required for a complete combustion of the sludge, was modified for each temperature. The values were 0.2, 0.3 and 0.4.
- Steam. A set of tests were carried out setting the ER at 0.3, modifying the temperature (750 °C (1023 K), 800 °C (1073 K) and 850 °C (1123 K)) and introducing steam until a steam-to-biomass ratio (S/ B), defined as the flow rate of steam fed to the reactor divided by the flow rate of sludge (dry and ash free), of 0, 0.5 and 1 was reached. It was decided to set the ER at 0.3 because after the first set of tests it was observed that under these conditions, tar production was relatively low (11 mg/g daf) and the LHV of the gases remained at acceptable levels (4 MJ/Nm^3).
- Alumina. Different quantities of catalyst (alumina) were placed in the gasifier (0%, 10% and 15% by weight of fed sludge), the temperature was modified (750 °C (1023 K), 800 °C (1073 K) and 850 °C (1123 K)) and the ER was set at 0.3.
- Alumina and steam. Tests were performed adding alumina (10% by weight of fed sludge) at a temperature of 800 °C (1073 K), ER 0.3 and modifying the S/B ratio (0, 0.5 and 1). The tests were carried out at 800 °C (1073 K) because problems were found to stabilise the
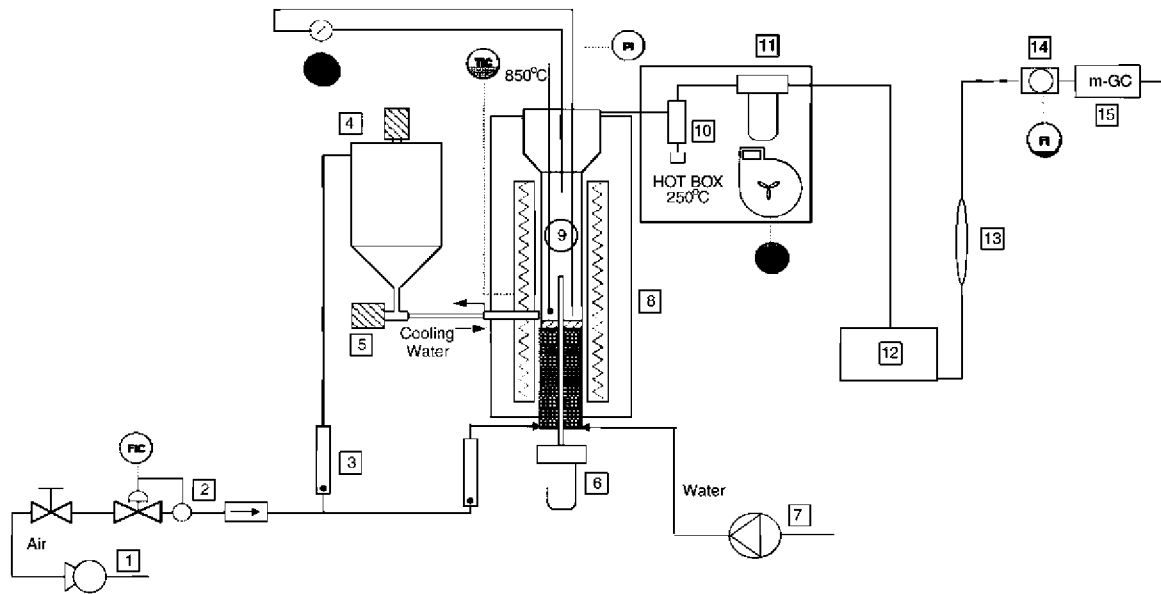
Fig. 1. Diagram of the laboratory scale plant: (1) compressor; (2) mass flow controller; (3) rotameter; (4) feed hopper; (5) screw feeder; (6) ash hopper; (7) peristaltic pump; (8) furnace; (9) reactor; (10) cyclone; (11) hot filter; (12) condensation train; (13) water filters, silica gel and cotton; (14) mass flow meter; (15) micro gas-chromatograph.

temperature at 850 °C (1123 K) with steam in the tests with no catalyst.

The conditions and results of the tests carried out are shown in Tables 2 and 3. Prior to each test 80 g of silica sand (or sand/catalyst mixture in the proportion as reported in Table 3) were placed in the gasifier in the right proportions. Once the temperature of the test had been reached, the gasifier was continuously fed with sludge and a specific sand-catalyst mixture, as reported in Tables 2 and 3. In order to avoid the effect of the transient period [29], tests were running along 30 min in order to reach stable conditions, and after that, the tests were continued for 50 min.

3. Results

3.1. Influence of temperature and equivalence ratio in air gasification

Fig. 2 shows the evolution of the average gas composition as a function of temperature for different ER in the tests performed with air and without addition of alumina. Higher temperatures favour hydrogen production. The concentration of CH_4 increases slightly with temperature because higher temperatures produce more intense volatilisations [30] and the cracking reactions (generation of H_2 and light hydrocarbons) prevail over reforming reactions (generation of H_2 and CO from light hydrocarbons).

Table 2

Results of gasification experiment without catalyst. Effect of temperature, equivalence ratio (ER) and steam-biomass ratio (SB).

Parameters	Units	Test number														
		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
Temperature	K	1023	1073	1123	1023	1073	1123	1023	1073	1123	1023	1073	1123	1023	1073	1123
	°C	750	800	850	750	800	850	750	800	850	750	800	850	750	800	850
u/u_{mf}		3.6	3.8	4	3.6	3.8	4	3.6	3.8	4	3.6	3.8	4	3.6	3.8	4
ER		0.2	0.2	0.2	0.3	0.3	0.3	0.4	0.4	0.4	0.3	0.3	0.3	0.3	0.3	0.3
S/B											0.5	0.5	0.5	1.0	1.0	1.0
Sludge	g/min	2.2	2.2	2.2	1.4	1.4	1.4	1.1	1.1	1.1	1.4	1.4	1.4	1.4	1.4	1.4
Sand	% fed sludge	20	20	20	20	20	20	20	20	20	20	20	20	20	20	20
WHSV	h^{-1}	2.4	2.4	2.4	1.6	1.6	1.6	1.2	1.2	1.2	1.6	1.6	1.6	1.6	1.6	1.6
Composition	Dry basis															
H_2	%	9.8	14.0	16.4	7.6	10.4	12.1	3.2	8.5	9.7	9.9	11.0	15.5	10.0	12.0	16.1
N_2	%	60.9	52.2	47.7	65.1	60.8	57.7	74.9	65.3	63.3	60.8	59.2	53.8	60.4	58.5	53.1
CH_4	%	3.7	4.9	5.7	2.7	3.0	3.3	1.0	2.4	2.1	2.8	3.1	3.2	2.9	2.7	3.0
CO	%	7.2	10.7	12.6	6.9	8.0	10.1	4.3	6.9	8.2	7.7	8.4	10.4	7.8	8.6	10.1
CO_2	%	14.2	13.6	13.0	14.1	14.1	13.1	14.1	13.8	13.9	14.8	14.8	14.1	15.6	15.2	14.7
C_2H_6	%	0.15	0.10	0.07	0.14	0.09	0.04	0.09	0.07	0.03	0.07	0.04	0.03	0.06	0.03	0.02
C_2H_4	%	2.6	3.2	3.3	2.1	2.2	2.4	0.8	1.5	1.4	1.9	2.0	1.5	1.8	1.6	1.5
LHV gas	MJ/Nm^3	3.4	4.8	5.6	2.8	3.3	3.9	1.3	2.7	2.9	3.2	3.5	4.2	3.2	3.5	4.2
HHV gas	MJ/Nm^3	3,7	5,2	6,1	3,0	3,6	4,2	1,4	3,0	3,2	3,4	3,8	4,6	3,5	3,8	4,6
Y_{gas}	Nm^3/kg sludge, daf	1.9	2.0	2.0	2.5	2.5	2.6	2.9	3.1	3.1	2.4	2.6	2.8	2.5	2.6	2.8
C_{tar}	g/Nm^3	17.3	14.1	10.2	11.8	7.6	4.3	6.6	4.8	3.0	11.2	7.2	4.0	10.9	6.9	3.9
Y_{tar}	mg/g sludge, daf	32.3	28.1	20.7	29.0	19.2	11.2	19.5	14.7	9.1	27.3	18.6	10.9	26.9	18.2	10.8
GMB	%	96.7	97.2	95.3	98.9	96.9	96.9	97.8	97.6	99.4	96.5	99.7	97.3	95.4	97.5	95.0
Xc	%	54.0	67.3	72.8	66.2	71.3	77.9	61.6	79.5	89.1	72.2	79.7	86.7	74.7	79.8	87.6
X_{H2O}	%										4.1	4.1	22.2	10.0	13.3	23.3
GE $_{LHV}$	%	28.7	43.2	50.7	30.4	37.4	45.5	16.6	38.1	42.3	35.2	41.0	53.1	35.6	41.1	53.0
GE $_{HHV}$	%	28,9	43,4	51,6	30,8	38,0	46,1	16,9	38,9	43,3	35,8	41,9	54,6	36,4	42,3	54,6
Char	g/kg daf	87.3	69.8	71.1	88.3	56.0	37.4	45.4	38.6	12.8	37.2	16.5	13.2	34.7	14.9	14.9

Table 3
Results of gasification experiment with catalyst. Effect of temperature, amount of catalyst and steam-biomass ratio (SB).

Parameter	Units	Test number							
		16	17	18	19	20	21	22	23
Temperature	K	1023	1073	1123	1023	1073	1123	1073	1073
	°C	750	800	850	750	800	850	800	800
u/u$_{mf}$		3.6	3.8	4	3.6	3.8	4	3.8	3.8
ER		0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3
S/B								0.5	1.0
Sludge	g/min	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.4
Alumina	% fed sludge	10	10	10	15	15	15	10	10
Sand	% fed sludge	10	10	10	5	5	5	10	10
WHSV	h^{-1}	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6
Composition	Dry basis								
H$_2$	%	8.0	12.3	15.0	8.8	12.1	14.2	13.3	15.2
N$_2$	%	64.9	60.1	55.1	64.4	59.8	56.5	58.2	56.1
CH$_4$	%	2.3	2.8	2.8	2.3	2.8	2.5	2.4	2.6
CO	%	7.6	8.5	12.2	7.7	8.3	12.2	8.4	8.1
CO$_2$	%	14.2	13.5	12.5	14.1	12.6	12.2	14.9	15.3
C$_2$H$_6$	%	0.06	0.05	0.02	0.05	0.05	0.02	0.03	0.03
C$_2$H$_4$	%	1.5	1.3	1.0	1.3	1.4	1.0	1.3	1.3
LHV gas	MJ/Nm^3	2.7	3.5	4.3	2.8	3.5	4.1	3.5	3.7
HHV gas	MJ/Nm^3	3,0	3,8	4,7	3,1	3,8	4,5	3,8	4,0
Y$_{gas}$	Nm^3/kg sludge, daf	2.5	2.7	2.8	2.5	2.8	2.8	2.7	2.7
C$_{tar}$	g/Nm^3	7.4	4.3	2.3	4.7	2.6	1.4	3.4	3.5
Y$_{tar}$	mg/g sludge, daf	18.3	11.8	6.4	11.7	7.2	3.9	9.1	9.5
GMB	%	99.0	100.3	100.9	100.1	102.8	101.0	99.9	96.4
Xc	%	68.4	73.1	85.5	69.4	71.4	83.3	79.0	80.8
X$_{H2O}$	%							14.0	18.9
GE $_{LHV}$	%	30.6	41.1	53.5	32.3	41.3	51.1	42.4	45.6
GE $_{HHV}$	%	31,3	42,3	55,3	33,2	42,6	52,8	43,8	47,3
Char	g/kg daf	49.6	57.9	18.2	51.3	84.3	16.5	26.5	29.8

CO production increases with temperature, which is due to incomplete combustion reactions and to the Boudouard and CO$_2$ reforming reactions. This assertion is reinforced by the increase in the CO/CO$_2$ ratio with temperature (Fig. 3). However, CO$_2$ production slightly decreases with temperature. This behaviour does not follow such well-defined patterns as those of the H$_2$ and CO. This may be due
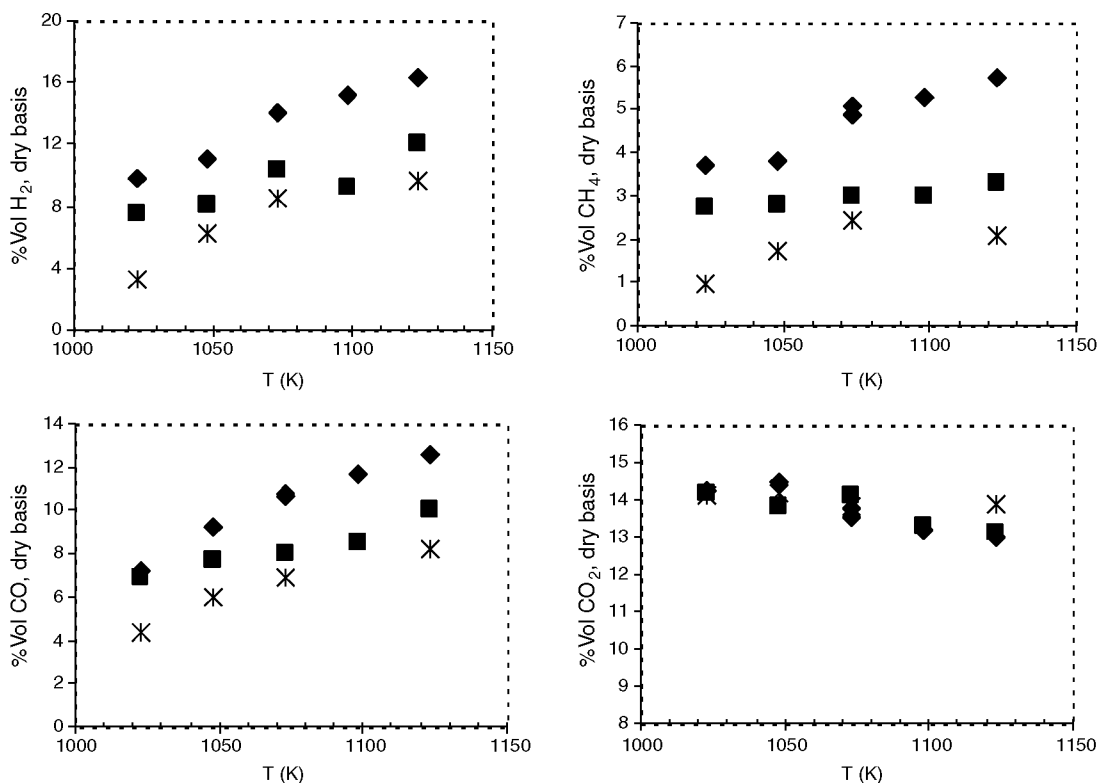

Fig. 2. Evolution of gas composition with temperature for different ER (◆ 0.2; ■ 0.3; ✳ 0.4).
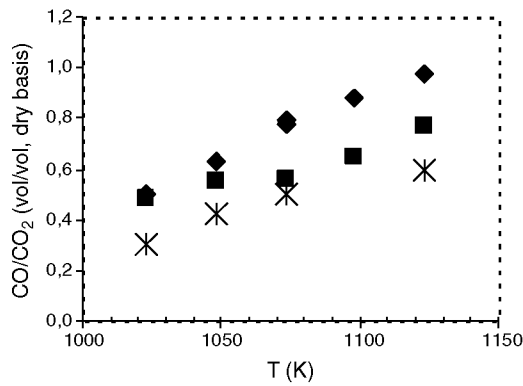
Fig. 3. Variation of CO/CO_2 ratio with temperature and ER (◆ 0.2; ■ 0.3; ✲ 0.4).

Fig. 5. Tar concentration for different temperatures (✲ 1023 K; ◆ 1073 K; ■ 1123 K) and equivalence ratios.

to the fact that CO_2 production is not only a consequence of volatilisation of the organic fraction, but also due to the heterogeneous reactions of CO_2 with char and tars [30].

The dependence of the composition of the gases on the ER for temperatures of 750 °C (1023 K), 800 °C (1073 K) and 850 °C (1123 K) is shown in Fig. 4. H_2 and CO production decreases as ER rises and temperature decreases. When the ER is high, oxidation reactions are favoured due to a higher oxygen content leading to greater amounts of CO_2 and smaller amounts of H_2 and CO [31]. At the same ER, H_2 and CO production decreases at lower temperatures due to the lower importance of the reforming, cracking and water–gas reactions, all of which are endothermic. The production of CH_4 and C_nH_m decreases as ER rises (Table 2) due to partial oxidation reactions [15,32].

Fig. 5 shows the variations in tar concentrations under different gasification conditions. Tar content decreases in every case as temperature rises [8,33], which is a result of cracking of tars and
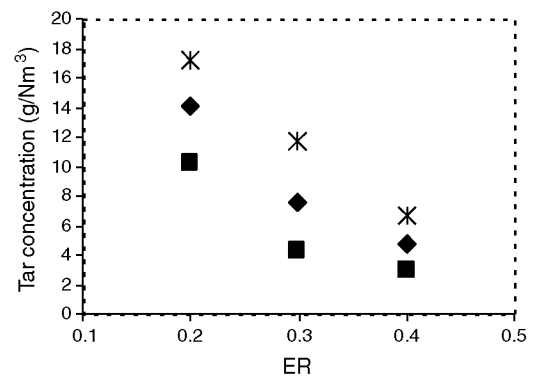
steam and CO_2 reforming reactions [27]. At the same temperature, tar concentration decreases as ER increases due to the oxidation of tars.

Gas production and its lower heating value (LHV) were similar to that reported by other researchers [15,26] (Table 2, tests 1 to 9). Gas production increases with temperature and ER due to a more intense volatilisation of the sludge, the decomposition of the tars and conversion of the char [30]. Higher temperatures favour gasification reactions where combustible gases (H_2, CO and CH_4) are produced. However, increases in the ER involve the oxidation part of these combustible gases resulting in a considerable loss of LHV.

The cold gas efficiency and carbon conversion in Fig. 6, calculated as in Olivares et al. [9] and Cao et al. [34] respectively, are comparable to those found by other authors for biomass and sludge [26,34,35]. The cold gas efficiency varies between 17% (at 750 °C (1023 K) and ER 0.4) and 51% (at 850 °C (1123 K) and ER 0.2) while carbon conversion varies between 54% (at 750 °C (1023 K) and ER 0.2) and 89% (at 850 °C (1123 K) and ER 0.4).
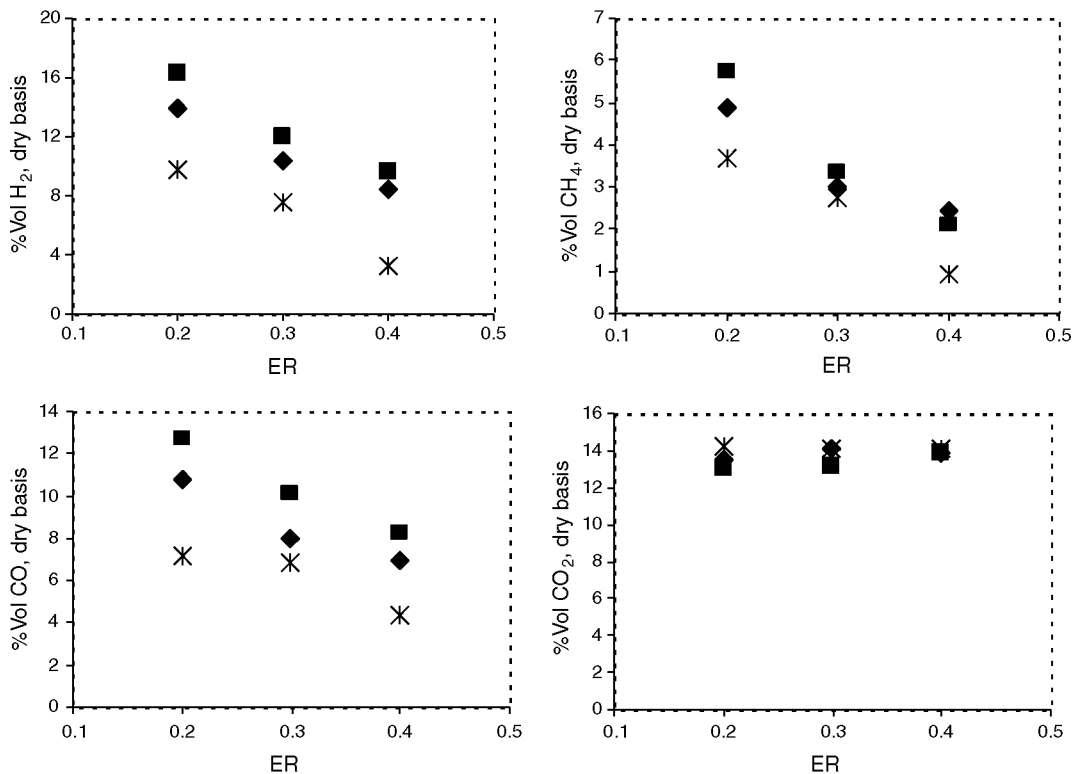

Fig. 4. Evolution of gas composition according to the variation of ER and temperature: ✲ 1023 K; ◆ 1073 K ; ■ 1123 K.
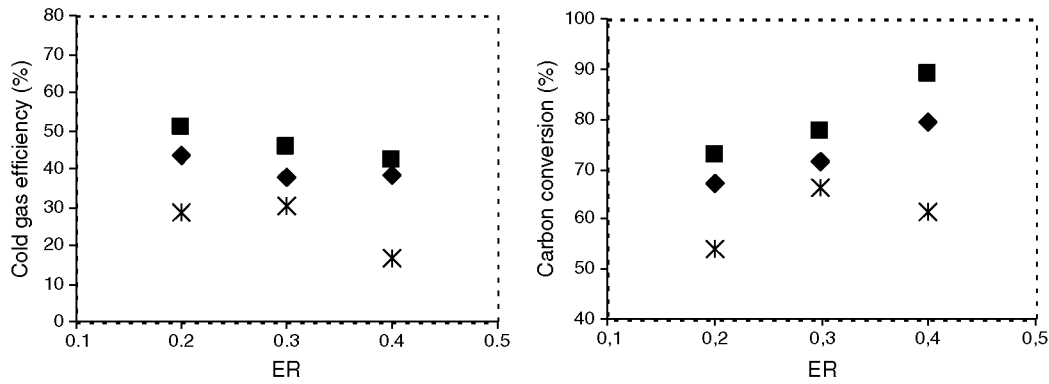
Fig. 6. Cold gas efficiency and carbon conversion for different ER and temperatures (✳ 1023 K; ♦ 1073 K; ■ 1123 K).

3.2. Influence of steam as gasifying agent

The results of the gasification tests conducted with steam are shown in Table 2, tests 10 to 15. Fig. 7 shows the change in gas composition at different temperatures and ratios SB, being $ER = 0.3$.

The results are similar to those found by Campoy et al. [36], the addition of steam increases hydrogen production due to the water–gas-shift, steam reforming and water–gas (at the highest temperatures) reactions. This finding agrees with the great reduction in the production of light hydrocarbon C_nH_m (Table 2) and with the increase in CO_2 production. The CO production is slightly higher in the steam tests than in the air tests. These results differ from what has been observed by other authors [15,32,36,37], who found lower CO contents as SB increased due to the water–gas-shift reaction. This behaviour might be due to the prevalence of the reforming and gas–solid reactions (Boudouard and water–gas), which are promoted by the gasification conditions and the gasifier design.

Tar production decreases in the presence of steam due to steam reforming reactions [15]. This fact, combined with the water–gas reaction between steam and char produces slight increases in CO and H_2 production [10]. The smaller amount of char found in these tests (Table 2) strengthens this assertion. The variation in gas composition and more specifically the increase in H_2 content, give rise to increases in the LHV of up to 15%.

The addition of steam leads to a mean improvement of 14% in cold gas efficiency and 12% in carbon conversion. The maximum values for these parameters (53% and 88%, respectively) are found at 850 °C (1123 K) and $SB = 1$.

3.3. Effect of adding alumina to the bed

Table 3 shows the results of the gasification tests with alumina as primary catalyst. In air gasification (tests 16 to 21), the alumina increases the production of H_2 compared to the tests without catalyst for the whole range of temperatures, reaching contents of 15% by volume at 850 °C (1123 K). There is a tendency for CO production to rise, particularly at higher temperatures (850 °C; 1123 K), and for CH_4, CO_2, and C_nH_m, production to decrease, which agrees with Manyà
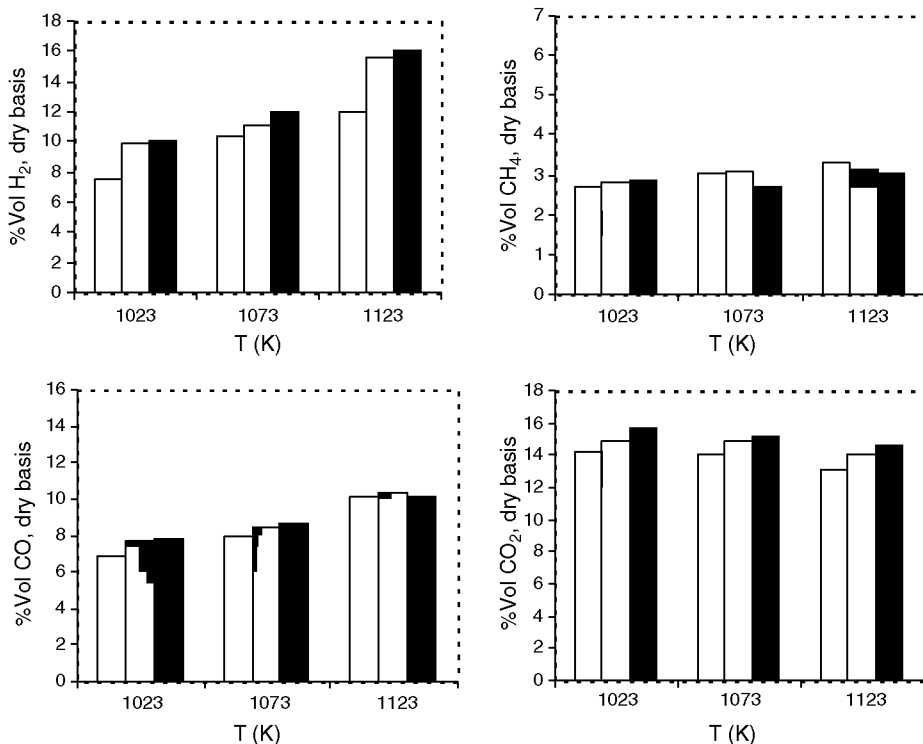

Fig. 7. Comparison of the composition of gasification gases at different temperatures, ER 0.3 and different SB values (☐ 0; ■ 0.5; ■ 1).
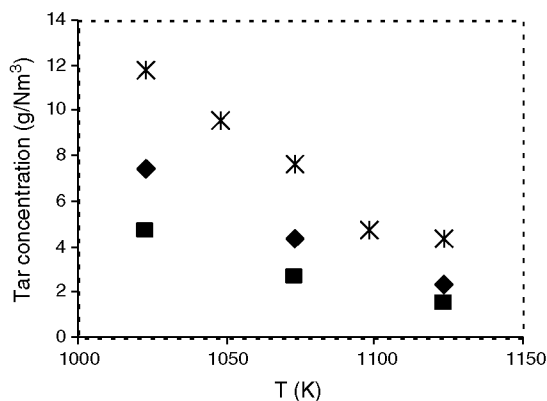
Fig. 8. Tar concentration for different temperatures and amounts of alumina in the feed
(✳ 0% no catalyst; ◆ 10% fed sludge; ■ 15% fed sludge). ER = 0.3.

et al. [26]. The results indicate that the presence of alumina favours cracking and CO_2 reforming reactions, since, in these tests, where the water only comes from the sludge, the influence of water–gas-shift and steam reforming reactions is limited. Gas composition hardly varies when the amount of catalyst placed on the bed is modified.

Fig. 8 shows the decrease in tar content in gases using alumina. In the presence of alumina, tar reduction varies between 37%, at 750 °C (1023 K) with 10% of catalyst in the feed and 65% at 850 °C (1123 K) with 15% of alumina in the feed. In previous works, tar production reduction close to 44% has been found using 8% by weight of alumina in the feed at 850 °C (1123 K) [26]. Similar results have been reported here (43% at 850 °C (1123 K) with 10% alumina). However, in contrast with Manyà et al. [26] where larger amounts of catalyst did not reduce tar production, this research has achieved additional reductions by increasing the catalyst added to the bed.

Cold gas efficiency and carbon conversion are higher in the presence of alumina due to a slight increase in gas production and in its LHV. On average, and compared to the tests without catalyst, cold gas efficiency is 10% higher and carbon conversion 5% higher.

Fig. 9 illustrates the gas composition and tar concentration found in tests with air but without alumina (test 5), tests with air and alumina (test 17) and tests with steam and alumina (tests 22 and 23). H_2 production increases with steam-alumina up to 46% compared with the tests with air and without catalyst. Contrary to the tests without catalyst, CO production decreases with steam due to the water–gas-shift reaction. The predominance of this reaction not only explains the increase in the content of H_2 and the decrease in CO, but also the increase
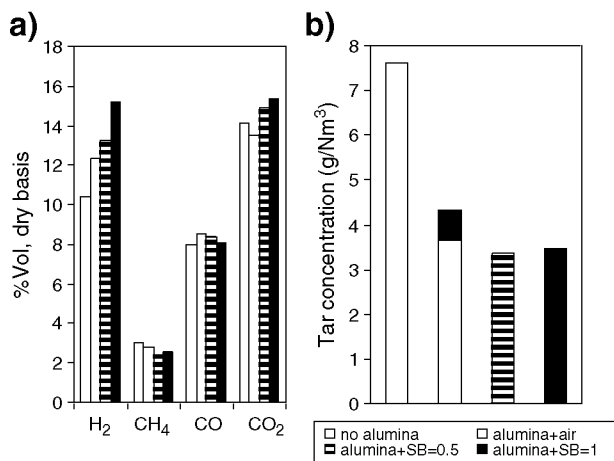
Table 4
Effect of alumina and steam on the production of gases, lower and higher heating value, cold gas efficiency and carbon conversion.

Parameter	Units	Air without alumina	Air with alumina	Alumina and steam (SB = 0.5)	Alumina and steam (SB = 1)
LHV gas	MJ/Nm³	3.3	3.5	3.5	3.7
HHV gas	MJ/Nm³	3.6	3.8	3.8	4.0
Y_{gas}	Nm³/kg sludge, daf	2.5	2.7	2.7	2.7
Xc	%	71.3	73.1	79.0	80.8
GE_{LHV}	%	37.4	41.1	42.4	45.6
GE_{HHV}	%	38.0	42.3	43.8	47.3

in the concentration of CO_2. The CH_4 production decreases in the steam catalysed tests mainly due to the steam reforming reactions.

The reduction in tar content in the gasification tests with steam-alumina is 20% compared with the tests with air-alumina and 55% compared with the tests with air and without catalyst (Fig. 9-b). As Table 4 shows, the presence of catalyst increases gas production compared with the tests without catalyst, but adding steam to the process does not seem to have a significant effect. Something similar occurs with the LHV, although the combined presence of catalyst and steam leads to increases in H_2 at higher SB, it also causes a reduction in CO and CH_4 content due to water–gas-shift and steam reforming reactions, respectively. The final results are small variations in LHV when steam is added to the process. For these reasons, cold gas efficiency and carbon conversion tend to be slightly higher (between 8 and 9%) at higher values of SB reaching a maximum increase of 11% at SB = 1.

4. Conclusions

In this work the combined influence of steam and alumina (as primary catalyst) on the products of gasification has been studied. Interest has focused on the composition, production and LHV of the gases, tar production, cold gas efficiency and carbon conversion.

In the absence of a catalyst, the introduction of steam into the gasification atmosphere increases hydrogen production and enhances the lower heating value of the gas, in addition tar production is reduced and the production of gases is slightly increased. The combination of alumina and air favours the production of H_2 and CO especially at high temperatures (850 °C; 1123 K), as well as a reduction in the production of CH_4, CO_2 and C_nH_m. It also significantly reduces tar content.

If steam is introduced in the presence of alumina, tar production is reduced even more. Generally speaking, steam has a more qualitative (it improves the composition of the gas from a higher hydrogen production point of view and a lower tar content) rather than quantitative influence as no significant changes are observed in gas production, in either LHV (the rise in H_2 concentration is offset by a fall in CO and CH_4), in cold gas efficiency or in carbon conversion.

As future lines of research we will consider conducting tests catalysed with steam at 850 °C (1123 K) and a study of the effects of the different primary catalysts on the gasification products, with the goal of producing a multi-purpose quality gas.

Nomenclature

|	
daf	dry and ash-free
ER	equivalence ratio, defined as the air-to-fuel ratio divided by the air-to-fuel ratio for the stoichiometric combustion, %
LHV	lower heating value of the produced gas, MJ/Nm³, dry basis
HHV	higher heating value of the produced gas, MJ/Nm³, dry basis
SB	steam-to-biomass ratio, defined as the flow rate of steam fed to the reactor divided by the flow rate of sludge (daf).
X_{H2O}	water conversion in the gasifier (%)
u	superficial gas velocity in the gasifier bed, cm/s

Fig. 9. Combined influence of steam and alumina (10% catalyst, ER = 0.3, T = 1073 K) in products of gasification: (a) gas composition; (b) tar concentration.

u_{mf}	minimum fluidisation gas velocity (gasifier bed conditions), cm/s
WHSV (weight hourly space velocity)	((kg sludge as received/h)/kg bed in the gasifier)
Nm^3	cubic meter, normal conditions (0 °C, 101 kPa)
Y_{gas}	gas yield, Nm^3 dry gas/kg sludge, daf
C_{tar}	tar concentration, g/Nm^3
Y_{tar}	tar yield, mg/g sludge, daf
GMB	global mass balance, %
X_c	carbon conversion, weight of carbon in the produced gas divided by weight of carbon in the sludge introduced in the gasifier
GE (cold gas efficiency) $= GE_{LHV}$	LHV of gas divided by the LHV of sludge; GE_{HHV}: HHV of gas divided by the HHV of sludge

Acknowledgements

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