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Techno-economic performance of HCl and SO2 removal in waste-to-energy plants by furnace direct sorbent injection

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3 # **Graphical Abstract**
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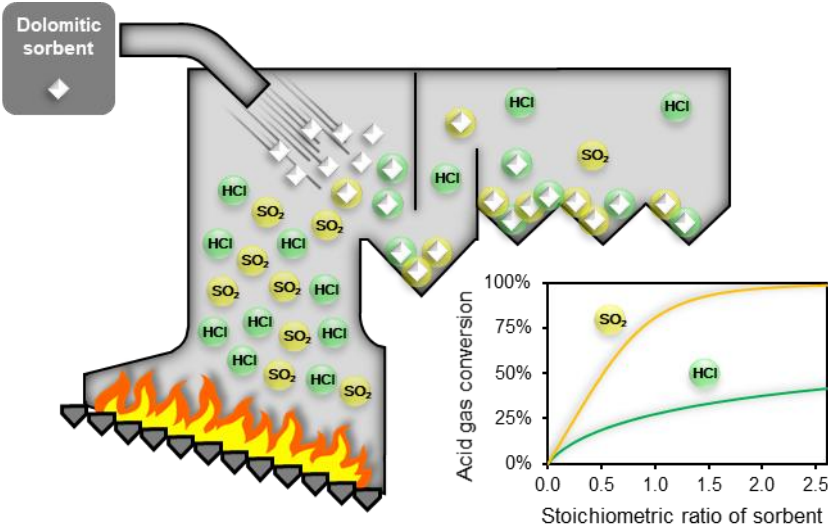
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HIGHLIGHTS

Techno-economic performance of HCl and SO$_2$ removal in waste-to-energy plants by furnace direct sorbent injection

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- Furnace injection of a dolomitic sorbent in WtE plants was tested
- Full-scale test runs were carried out to assess sorbent performance
- A model of acid gas removal efficiency was calibrated
- The dolomitic sorbent reacts at high temperature with acid gases
- The cost-effectiveness of dolomite sorbent injection was assessed

<u>Abstract</u>

With the impending release of Best Available Techniques (BAT) conclusions on waste incineration, existing European waste-to-energy (WtE) plants will be required to achieve a higher efficiency in the removal of several target pollutants, such as acid gases (above all, HCl and SO_2). The direct injection of a sorbent in the furnace as a primary deacidification stage may be a cost-effective option to achieve the required performances. The present study investigated the furnace injection of a specific dolomitic sorbent, with the aim of identifying the techno-economic optimum for the sorbent feed rate considering different scenarios of flue gas composition. A full-scale test run campaign was carried out on two WtE plants and a phenomenological model linking HCl and SO_2 conversion to reactant feed rate was developed. The model allowed assessing the cost-effectiveness of dolomitic sorbent furnace injection for WtE plants equipped with either a single or a two-stage acid gas treatment system. The addition of dolomitic sorbent resulted particularly suitable for WtE plants equipped with a single stage treatment coping with a high SO_2 concentration in the flue gas, where the injection of dolomitic sorbent can increase the redundancy of flue gas treatment with a reduction of the total operating costs.

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Revised Version

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Abstract

With the impending release of Best Available Techniques (BAT) conclusions on waste incineration, existing European waste-to-energy (WtE) plants will be required to achieve a higher efficiency in the removal of several target pollutants, such as acid gases (above all, HCl and SO₂). The direct injection of a sorbent in the furnace as a primary deacidification stage may be a cost-effective option to achieve the required performances. The present study investigated the furnace injection of a specific dolomitic sorbent, with the aim of identifying the techno-economic optimum for the sorbent feed rate considering different scenarios of flue gas composition. A full-scale test run campaign was carried out on two WtE plants and a phenomenological model linking HCl and SO₂ conversion to reactant feed rate was developed. The model allowed assessing the cost-effectiveness of dolomitic sorbent furnace injection for WtE plants equipped with either a single or a two-stage acid gas treatment system. The addition of dolomitic sorbent resulted particularly suitable for WtE plants equipped with a single stage treatment coping with a high SO₂ concentration in the flue gas, where the injection of dolomitic sorbent can increase the redundancy of flue gas treatment with a reduction of the total operating costs.

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Keywords

Waste-to-energy; Flue gas treatment; Acid gas removal; Sorbent; Furnace injection.

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1. Introduction

Acid gases such as HCl and SO_2 are typical pollutants formed in waste combustion, whenever chlorine and sulphur are present in the waste feed [1]. Control of their emission in waste-to-energy (WtE) facilities is performed by reaction with neutralising agents, supplied either as absorbent solutions in wet scrubbing equipment or as powdered alkaline sorbents in dry injection systems [2-6].

Given the recent trends in waste generation and management [7,8], overall Cl and S content in the waste feed of European WtE plants is increasing, due to the increasing amounts of industrial wastes fed to these disposal systems. On the other hand, WtE installations are subject to the strictest emission limits among all combustion facilities [9] and the release of the new Best Available Techniques (BAT) Reference Document for waste incineration [10] will further raise the bar, introducing new, ambitious emission targets.

The adoption of multi-stage treatment systems for the removal of acid gases, *i.e.* the combination of more than one unit for acid gas treatment, will be needed in many situations to obtain the required removal efficiency.

For retrofitting applications, the introduction of an additional acid gas removal stage should ideally require minimal impact on plant layout and limited investment costs. Furnace sorbent injection (FSI), *i.e.* the introduction of a powdered solid reactant for the capture of acid gases directly in the combustion chamber, is a method that meets both criteria [11,12]. The reactant is injected directly in the flow path of the flue gas by pneumatic conveying. Then, the reacted particles are deposited on the surface of the heat exchange equipment, thus a dedicated separation equipment is generally not required [13].

FSI was first introduced as a desulfurisation technique in coal combustion [14,15]. The sorbent, typically limestone, is blended with coal or injected with staged air [16]. Once introduced in the furnace at high temperature, limestone is calcined to calcium oxide (CaO), with the development of a porous structure that increases its reactivity towards acid gases. SO_2 removal efficiencies in the range 40-70% are expected [17,18]. Petrini et al. [19], Mura and Lallai [20] and Partanen et al. [21] demonstrated experimentally the suitability of high temperature reaction with limestone also for HCl capture. However, the reactivity of limestone at high temperature is hindered by pronounced sintering [22]. The onset of sintering for calcium carbonate occurs at T ~ 500 °C [23], whereas the injection near to the combustion area exposes the sorbent at temperatures ranging from 800 to 1200 °C. Upon sintering, the porous structure of the sorbent collapses and the surface area decreases, leading to a reduced sorption activity. Alternative Ca-based sorbents, as organic calcium salts [24,25] or, specifically, calcium magnesium acetate [26,27] have been proposed with the aim of achieving higher acid gas removal efficiency thanks to superior surface areas.

More recently, commercial sorbents derived from dolomite rock have been introduced, with the aim of overcoming the limitations of limestone. The presence of magnesium confers a higher resistance to sintering [28] and reduces the clogging and fouling that the deposition of limestone residues typically causes to boilers [29]. Biganzoli et al. [30] offered a first quantitative assessment of the full-scale acid gas removal performance of a dolomitic sorbent, carrying out a test campaign in four Italian WtE facilities. They observed that a furnace sorbent injection at a feed rate of 6 kg/t_{waste} yielded a reduction of acid gas concentration downstream of the boiler in the range 7–37% for HCl and 34–95% for SO_2.

The aim of the present study is to collect extended and detailed quantitative information on the performance of acid gas removal based on dolomitic sorbent direct furnace injection, quantifying the relationship between reactant feed rate and acid gas removal efficiency. The

82   effectiveness of reactant injection towards HCl and SO_2 was systematically investigated by
83   full-scale test runs carried out in two Italian WtE facilities. The experimental data were used
84   to develop a specific phenomenological model for removal efficiency with respect to sorbent
85   feed and initial acid gas concentrations. The model was used to assess the economics of the
86   process, also considering the other acid gas treatment stages and the composition of the
87   inlet flue gas stream.
88
89

90   # 2. Materials and methods

91
92   ## 2.1 The dolomitic sorbent
93   The solid reactant tested in the experimental campaign, commercially known as Depurcal
94   and supplied by Unicalce SpA (Italy), is obtained via the calcination and slaking of dolomite
95   rock. The reactant, referred to as dolomitic sorbent (DS) in the following, is mainly composed
96   of calcium hydroxide ($Ca(OH)_2$), magnesium hydroxide ($Mg(OH)_2$) and magnesium oxide
97   (MgO). According to patent information [29], an average composition on mass basis of the
98   DS is the following: 58.2% $Ca(OH)_2$, 22.3% $Mg(OH)_2$, 19.2% MgO and 0.3% impurities
99   (SiO_2, Fe_2O_3, Al_2O_3).
100
101   ## 2.2. Reference WtE plants
102   The experimental assessment of DS performance was carried out in two Italian WtE
103   facilities, here referred to as plant A and plant B. The key features of the two plants are
104   summarised in Table 1, while the layout of their acid gas treatment systems is sketched in
105   Figure 1. With respect to acid gas removal, plant A presents a two-stage treatment system
106   (Fig. 1a), consisting in the injection of calcium hydroxide, $Ca(OH)_2$, and sodium bicarbonate,
107    $NaHCO_3$, respectively in two consecutive sections of reaction and filtration. Hence, the
108   introduction of a FSI of the dolomitic sorbent realises a three-stage acid gas removal system.
109   Conversely, plant B is equipped with a single acid gas treatment stage fed with sodium
110   bicarbonate (Fig. 1b). Here, the FSI of the dolomitic sorbent introduces a second stage of
111   acid gas removal.
112   The acid gas treatment lines of the two selected facilities represent two of the most common
113   configurations for WtE flue gas treatment [31] and are both listed among the best available
114   techniques for waste incineration [10]. The reactions between the solid sorbents and the acid
115   gases are illustrated in section S4 of the Supplementary Material.
116
117   **Table 1.** Main features of the WtE plants considered in the study.

	Plant A	Plant B
Nominal capacity (t_{waste}/h)	16	9.5
Average flue gas flow rate (Nm^3/h)	110'000	65'000
Range of HCl concentration in the untreated flue gas (mg/Nm^3)	600-1000	800-1400
Range of SO_2 concentration in the untreated flue gas (mg/Nm^3)	10-100	10-100
Acid gas treatment line	$Ca(OH)_2$ + FF + $NaHCO_3$ + FF	$NaHCO_3$ + FF

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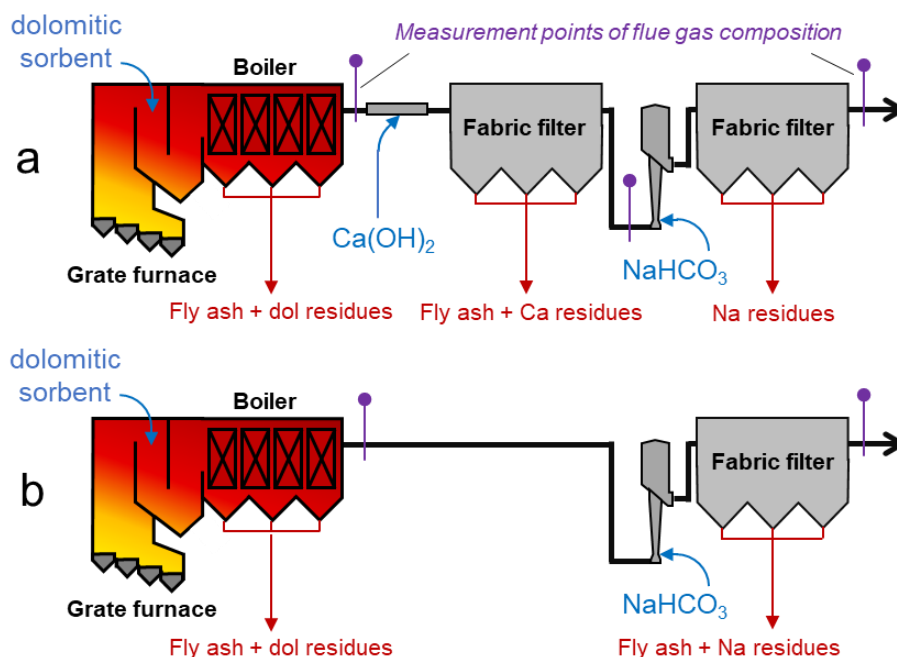
Figure 1. *Configurations of acid gas treatment in the two plants used for the experimental tests: a) two-stage system (CaOH)$_2$ + NaHCO$_3$), which becomes a three-stage system with the furnace injection of dolomitic sorbent, and b) single stage system (NaHCO$_3$), which becomes a two-stage system with the furnace injection of dolomitic sorbent. The points where measurement of flue gas composition is available are highlighted.*

2.3 Test configurations and protocols

Figure 1 shows the feed point of the sorbent in the two plants. The dolomitic sorbent was injected to the top of the combustion chamber by means of a screw feeder that allowed the control of the feed rate. The injected DS reacted with the acid pollutants both while entrained in the flue gas stream and after deposition on the surface of the boiler. The concentration of the acid pollutants (primarily, HCl and SO$_2$) in the flue gas was measured by Fourier-transform infrared (FTIR) spectrometry at the boiler outlet. In both plants, the FTIR device installed at the boiler outlet was part of a continuous emission monitoring system (CEMS) compliant with the Quality Assurance Level 3 (QAL3, see EN 14181 [32,33]) required by the EU Industrial Emissions Directive [34].

2.3.1 Test runs in specific operating conditions

The relationship between DS feed rate and HCl/SO$_2$ removal was analysed quantitatively by dedicated test runs. The test runs were carried out injecting the reactant at incremental steps of constant feed rate for defined time intervals. The steps of feed rate tested were: 40, 60, 90, 100 and 200 kg/h for plant A, ranging from to 0.36 to 1.8 kg/10^3 Nm^3; 40, 120 and 200 kg/h for plant B, ranging from 0.62 to 3.1 kg/10^3 Nm^3. The injection of reactant was stopped for 40 min between each feed rate step to allow monitoring the acid gas concentrations in the untreated gas and to avoid any residual effect of the injected DS in the previous step of feed rate on the following step. Table 2 provides the list of all the tests completed.

Data analysis was carried out considering the trend of acid gas concentration at the boiler outlet during the tests. Figure 2 shows the supply of different DS feed rates and the effect on HCl and SO$_2$ concentrations at the boiler outlet during one of the test runs. The conversion

of HCl or SO_2 related to each step of DS feed rate was calculated from the difference in acid gas concentration before and after the start of DS injection:

$$\overline{\hspace{3cm}} \tag{1}$$

where $C_{acid\ gas,w/o\ DS}$ is the concentration measured in the minute before DS injection and $C_{acid\ gas,w\ DS}$ is the pseudo-stationary concentration measured during DS injection after the transient of concentration decrease was concluded. Similarly, the conversion of acid gases related to each step of DS feed rate was also calculated when DS injection was stopped. In these cases, $C_{acid\ gas,w\ DS}$ was taken as the concentration measured in the minute before the interruption of DS injection, while $C_{acid\ gas,w/o\ DS}$ was the pseudo-stationary concentration measured after DS injection was interrupted and the concentration increase transient was concluded.
A screening protocol was established to accept the experimental data, given the uncertainty due to fluctuations in acid gas concentration formed in combustion flue gas. The exclusion of outliers (i.e. data for which an abrupt variation of HCl and SO_2 generated by waste combustion either covers or magnifies the effect of DS injection) was performed during model fitting by least trimmed squares regression [35], as described in section 3.1.

Table 2. *Test runs performed and steps of reactant feed rate.*

DS feed rate (kg/h)	Plant A Test id.								Plant B Test id.
	1A	**2A**	**3A**	**4A**	**5A**	**6A**	**7A**	**8A**	**1B**
40	✓	✓	✓	✓	✓	✓	✓	✓	✓
60 [a]	✓	✓	✓	✓	✓	✓	✓	✓	
90						✓	✓	✓	
100	✓	✓	✓	✓	✓				
120									✓
200	✓	✓	✓	✓	✓				✓

[a] the feed rate at 60 kg/h was imposed before and after each test run, as it was the selected constant rate for the 3-month experimental campaign described in section 2.3.2

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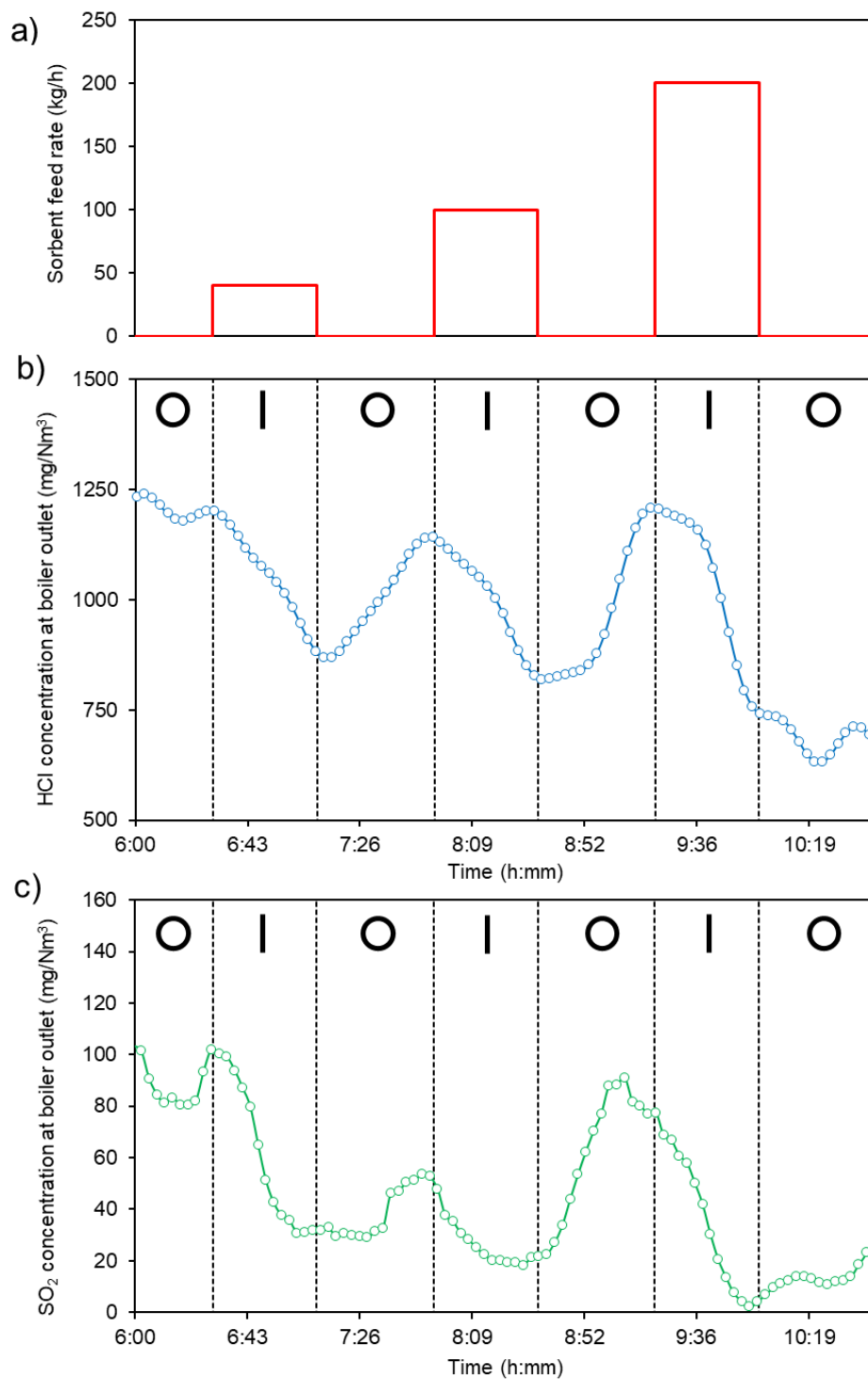
Figure 2. *(a) Feed rate of dolomitic sorbent, and corresponding (b) HCl and (c) SO$_2$ outlet concentration during test 3A. The symbols I and O indicate respectively the time intervals with and without DS injection. Dots are 5-min moving average values of measured concentration.*

181
2.3.2 Test run in normal operating conditions

In plant A, the injection of DS was also tested for 3 consecutive months during normal operation of the facility. DS was fed to the flue gas treatment line at a constant feed rate of 60 kg/h (0.55 kg reactant/10^3 Nm^3 of flue gas, considering the average flow rate reported in Table 1). The values of HCl and SO_2 concentration at the boiler outlet, continuously monitored by the FTIR device during the test period, were collected and compared to the same data measured in the previous 3 months of operation without DS injection. The comparison was made considering for the two periods the distribution of the hourly averages of concentration values. The overall 3-month average concentration in the period $\bar{C}_{period}$ and the related standard deviation σ_{period} were calculated starting from the hourly average concentrations, as follows:

$$\overline{\hspace{3cm}} \tag{2}$$

$$\overline{\hspace{4cm}}$$

$$\overline{\hspace{4cm}} \tag{3}$$

The total cumulative mass flow of HCl and SO_2 at the boiler outlet over the 3 months was also compared between the two periods.

The main objective of this analysis was to assess the global effect of the dolomitic sorbent in reducing the average concentration of acid gases exiting the combustion chamber. This comparison is based on the underlying assumption that the average concentration of acid pollutants in the raw flue gas upstream of DS injection during the 3 months of DS injection is approximately the same as that measured at the boiler outlet in the previous 3 months without DS injection. This assumption is valid only if there is no significant seasonal effect in waste composition that varies the raw HCl and SO_2 concentration between a 3-month period and the following. The validity of this assumption was checked by comparing the concentration data at boiler outlet for the same two 3-month periods in the previous year (referred to in the following as 'control year'), when no DS testing was carried out. As discussed in section 4.1, no appreciable variation of acid gas concentration at boiler outlet was found between the two 3-month periods of the control year.

3. Models

3.1 Modelling of the performance of dolomitic sorbent

When the sorbent is injected in the combustion chamber of a WtE plant, at temperatures higher than 800 °C, the hydroxides decompose releasing water vapor and forming meso and macropores in the material [30].

$$\tag{4}$$

The development of the porous structure promotes the reactivity of the sorbent towards acid gases. The possible reactions with HCl and SO_2 are the following:

$$\tag{5}$$

7

| 222 | (6) |

| 223 | (7) |

| 224 | (8) |

225 However, it is generally assumed that MgO is chemically inert to acid gases at the
226 temperatures of interest (800-1000 °C), on the basis of experimental findings at laboratory
227 [21] and pilot scale [15]. A wide literature on the high-temperature CO_2 sorption by dolomitic
228 compounds provides similar evidence [23,36]. In addition, a thermodynamic analysis of
229 reactions 5-8, shown in section S1 of the Supplementary Material, confirms that, at
230 temperatures higher than 900°C, only the reactions between acid gases and the Ca-based
231 phase of the sorbent present a negative Gibbs free energy of reaction. Therefore, in the
232 activated dolomitic sorbent after reaction 4, it is considered that CaO is the reactive
233 compound towards acid gases, while MgO acts almost exclusively as structural agent,
234 preventing the collapse of the porous structure developed in reaction 4, as suggested by
235 Biganzoli et al. [30].
236 A phenomenological model for the description of the acid gas removal process by solid
237 sorbents was used to generalise the information obtained from the test runs. The modelling
238 approach is described comprehensively in a previous study [37]. The model correlates the
239 removal efficiency of an acid pollutant i (X_i) to the stoichiometric ratio SR of sorbent j at a
240 generic time:

| 241 | —— | (9) |

242 where SR can be assessed as the ratio of the actual feed rate of the sorbent to the feed rate
243 theoretically required to convert all the acid pollutant i in the inlet gas stream over a time
244 interval of interest. The exponent n is an empirical parameter that lumps together the effect
245 of all the variables influencing the gas-solid reaction (reactivity of the sorbent, operating
246 conditions such as temperature and humidity, fluid-dynamics of the system). As such, the
247 value of n needs to be calculated from the test run data and is specific of the facility.
248 In the present study, least trimmed squares fitting was used to calculate the values of
249 interest of $n_{i,j}$ using the experimental data from the test runs. The model in Eq. 9 was log-
250 linearised as follows:

| 251 | —— | (10) |

252 Linear regression was then used to fit the model to the test run data and the least trimmed
253 squares procedure [35] for outlier exclusion was applied until a satisfactorily high value of R^2
254 was obtained for the fitted model ($R^2 > 0.99$ for HCl, $R^2 > 0.95$ for SO_2).
255 In order to apply the model to the dolomitic sorbent, it was assumed that its sorption capacity
256 towards acid gases is entirely ascribed to the Ca-based fraction (*i.e.* $Ca(OH)_2$ that
257 decomposes to CaO upon furnace injection), while the Mg-based compounds (MgO,
258 $Mg(OH)_2$) have only the role to avoid the high-temperature sintering of the material and do
259 not participate to the gas-solid reactions, as discussed above. Therefore, the SR was
260 defined on the basis of the Ca content in the sorbent. According to the typical composition of
261 the sorbent reported in section 2.1, the weight fraction of $Ca(OH)_2$ in the DS is 0.58. A
262 weight fraction of 0.42 is obtained when considering the amount of CaO formed from
263 hydroxide dehydration at high temperature after DS injection in the furnace. It was thus

264 possible to express the data from test runs in terms of SR_{CaO} and the related acid gas
265 conversions $X_{HCl,CaO}$ and $X_{SO2,CaO}$.
266 In particular, the mass feed rates of sorbent injected in the test ==, were converted to==
267 runs, SR_{CaO} as follows:

268 $$\rule{4cm}{0.4pt} \qquad \rule{7cm}{0.4pt} \tag{11}$$

269 where $\overline{\rule{1.5cm}{0pt}}$ is the volumetric flue gas rate, and $\overline{\rule{1.5cm}{0pt}}$ and $\overline{\rule{1.5cm}{0pt}}$ are the
270 concentrations of the acid gases in the untreated flue gas, measured in the periods without
271 DS injection as described in section 2.3.1.
272 In order to model the overall performance of the multi-stage acid gas removal process
273 implemented by both plant A and plant B, Eq. 9 was applied as well to the other stages of
274 acid gas removal. Details on the models and on the procedure for model calibration are
275 reported in section S4 of the Supplementary Material.
276

3.2 Modelling approach for the optimisation of acid gas removal
278 The cost-effectiveness of the introduction of DS injection in the acid gas removal process
279 can be evaluated estimating the overall operating costs of the process given the initial and
280 final acid gas concentrations. In a multi-stage treatment system, the required overall removal
281 efficiency can be obtained by different distributions of reactants in the removal stages.
282 Therefore, the objective of the optimisation is to assess the share of acid gas removal by DS
283 injection that minimises the overall operating costs of the process.
284 The performance model of section 3.1 was used to simulate the removal efficiencies of each
285 step in the acid gas abatement line. Simulations were performed for both plant A and plant
286 B. To take into account the variability of acid gas load, six different cases of raw flue gas
287 composition, listed in Table 3, were considered. The six cases cover the typical range of HCl
288 and SO_2 concentrations in the untreated flue gas for the WtE plants that hosted the
289 experimental campaign (see Table 1). The average flue gas flow rate of each plant (see
290 Table 1) and a concentration at stack of 2 mg/Nm^3 of HCl were considered in all the cases.
291

292 **Table 3.** *Cases of raw flue gas composition considered in the optimisation study.*

	C_{HCl} (mg/Nm^3)	C_{SO2} (mg/Nm^3)
Low Cl, low S	600	10
Medium Cl, low S	1000	10
High Cl, low S	1400	10
Low Cl, high S	600	200
Medium Cl, high S	1000	200
High Cl, high S	1400	200

293

294 Different configurations of the multi-stage treatment system were explored varying the HCl
295 conversion in the different stages, while keeping equal the overall efficiency of HCl removal
296 in the system:

297 $$\tag{12}$$

298 where N is the total number of abatement stages (2 for plant B, 3 for plant A), and $X_{HCl,j}$ is
299 HCl conversion in stage j. The value of $X_{SO2,overall}$ could not be kept constant as well, as

9

different repartitions of abatement between the stages were explored in the simulations, because each sorbent has a different relative reactivity towards HCl and SO_2. Nonetheless, given the significantly lower amount of SO_2 compared to HCl in the raw flue gases considered in Table 3, the concentration of SO_2 at stack resulted similar and lower than 1 mg/Nm^3 in all the simulations.

The consumption of reactants associated to the acid gas conversion in each abatement stage can then be calculated considering the reaction stoichiometry (Eqs. 4-6 for the dolomitic sorbent, Eqs. S5-S9 in section S4 of the Supplementary Material for the other reactants) and the model correlation provided by Eq. 9. Given the acid gas conversion and the required consumption of reactants, the mass rate and composition of the solid process residues generated in each stage can be quantified by means of a mass balance.

The overall operating cost of acid gas treatment was thus estimated as:

$$ \tag{13} $$

where $C_{reactants}$ and $C_{residues}$ are the costs respectively associated to the purchase of reactant and the management of process residues for each treatment stage. The values adopted for the unit costs are listed in Table 4. These numbers are average values for the Italian context. For a detailed overview on the cost data of dry acid gas treatment operations the reader is referred to Dal Pozzo et al. [37]. Here, it is worth recalling that the management of process residues is associated to two different fates and costs. Streams of residues that are mixed with fly ash coming from the combustion chamber and potentially carrying hazardous materials in traces (heavy metals, organochlorinated compounds) are to be sent to dedicated landfill sites [38,39]. Hence, recalling the schemes in Figure 1, this is the case for the residues of dolomitic sorbent, calcium hydroxide, and sodium bicarbonate, if no upstream dedusting equipment is installed (see Fig. 1b). Conversely, the process residues generated by the reaction between sodium bicarbonate and acid pollutants can be sent to dedicated recycling facilities [40] if a fabric filter is installed upstream of $NaHCO_3$ injection to collect separately the fly ash (see Fig. 1a).

Table 4. *Values of the unit costs adopted in the study for the considered cost entries.*

Cost entry	Unit cost (€/t)
Reactants	
Dolomitic sorbent	100
Calcium hydroxide	50
Sodium bicarbonate	240
Management of process residues	
Disposal (residues mixed with fly ash)	200
Recycling (Na residues separated from fly ash)	170

4. Results and discussion

4.1 Test run in normal operating conditions: effect of DS injection on the concentration of HCl and SO_2 leaving the furnace

The test run in normal operating conditions conducted at plant A had the aim of assessing the effect of a fixed feed rate of DS injection on the flue gas composition leaving the

10

337　furnace/boiler. The outcome of the test run on plant A in normal operating conditions was
338　evaluated comparing the hourly data of HCl and SO_2 concentrations in the gas at the boiler
339　outlet (first sampling point in Figure 1a) for the two 3-month periods with and without DS
340　injection, as described in the sampling protocol of section 2.3.2.
341　The hourly concentration data measured at the boiler outlet in each 3-month period were put
342　in ascending order and multiplied with their respective hourly flue gas flowrate data, in order
343　to obtain a curve of cumulative mass flow of the acid pollutants exiting the boiler.
344　Fig. 3a shows the cumulative mass flow of HCl at the boiler outlet for the two reference
345　periods. The total mass flow of HCl leaving the combustion chamber measured in the 3
346　months of DS injection amounted to 164.7t, compared to 218.2t for the 3 months without DS
347　injection. A 25% reduction of the HCl mass flow released in the reference period is thus
348　observed as a consequence of DS injection.
349　More specifically, the cumulated mass flow curves in Fig. 3a trace how the hourly data of
350　HCl concentration measured at the boiler outlet contributed to the total mass flow in the two
351　3-month periods. The concentration of HCl measured at the sampling point fluctuates over
352　time as a natural consequence of the variable Cl content of the waste burnt in the furnace: in
353　this plant, Fig. 3a shows that during the 3-month period without DS injection the HCl
354　concentration varied mostly from 400 to 1500 mg/Nm^3 with peaks up to 3000 mg/Nm^3. In the
355　3 months of DS injection this range of variation moved towards lower values (the cumulated
356　mass flow curve with DS injection is shifted leftward compared to that for the period without
357　DS injection). This is further clarified by the distributions of the hourly averages of HCl
358　concentration at the boiler outlet for the two reference periods shown in Fig. 3b. The hourly
359　values of HCl concentration recorded at the boiler outlet during the two 3-month periods
360　were subdivided in the 5 classes of the figure. For each class, the frequency of occurrence
361　of concentration values within that class range is reported as percent of the total hourly
362　concentration data in the 3-month period. The furnace injection of DS clearly resulted in a
363　shift of the distribution of HCl concentration towards lower values. With DS injection, the
364　frequency of concentrations higher than 800 mg/Nm^3 is more than halved, while the
365　occurrence of concentrations in the interval 400-600 mg/Nm^3 is approximately tripled.
366　The overall 3-month average of the hourly HCl concentration data at the boiler outlet with DS
367　injection resulted of 702 mg/Nm^3, compared to a value of 897 mg/Nm^3 for the 3-month
368　period without DS injection. This corresponds to an average 22% reduction of the HCl load
369　leaving the combustion chamber thanks to the DS effect. Furthermore, also the standard
370　deviation of the hourly HCl concentration data exhibited a noticeable decrease: a value of
371　287 mg/Nm^3 was obtained during DS injection, with respect to a value of 323 mg/Nm^3 in the
372　period without DS injection. A reduction of the standard deviation means that the hourly HCl
373　concentration values at the sampling point were less dispersed during DS injection and,
374　thus, less peaks at high HCl concentration were present compared to the period without DS
375　injection.
376　To confirm that all the aforementioned effects were owing to the DS injection and not to a
377　change of other operating conditions between the two 3-month periods, an identical analysis
378　was carried out for the same 3-month periods in the previous year of operation of the plant,
379　when no DS testing was performed. The results are shown in section S2 of the
380　Supplementary Material. No appreciable difference in the mass flow of HCl and in the
381　distribution of HCl concentration was observed between the first and the second period of
382　the control year. A difference lower than 2% is present among the 3-month average HCl
383　concentrations for the two periods. Therefore, it is reasonable to assume that the reduction
384　of the average HCl concentration in the flue gas observed in Fig. 3a and 3b is due to the

11

385 addition of DS and not to a variation in the overall average composition of the waste
386 feed between the periods.
387 The results observed with reference to SO_2 are similar to those obtained for HCl. Fig.
388 3c shows that the cumulative mass flow of SO_2 in the 3 months of DS injection was
389 equal to 5.3t, a 37% reduction compared to the 8.4 t calculated for the 3-month period
390 without DS injection.
391 It is evident that during DS injection, as observed for HCl, lower SO_2 concentrations at
392 the sampling point appeared more frequently than in the reference period without DS
393 injection. As shown in Fig. 3d, more than 60% of the hourly SO_2 concentration data
394 measured in the period with DS injection show values lower 15 mg/Nm^3, while the
395 occurrences in all the frequency classes higher than 15 mg/Nm^3 diminish systematically
396 compared to the period without DS injection.
397 The overall 3-month average of the hourly SO_2 concentration data at the boiler outlet in the
398 3-month period with DS injection resulted of 22.7 mg/Nm^3, compared to a value of 31.4
399 mg/Nm^3 without DS injection (a 28% reduction). Again, also the standard deviation of the
400 hourly SO_2 concentration data was noticeably lower during DS injection: 40.6 mg/Nm^3 vs.
401 52.6 mg/Nm^3.
402 For the control year (see section S2 in the Supplementary Material), no significant difference
403 in the global 3-month average concentration was recorded: the average SO_2 concentration
404 at the boiler outlet was 32.1 mg/Nm^3 in the first period and 31.0 mg/Nm^3 in the
405 second period. The total mass flow of SO_2 measured in the 3 months varied less than 2%
406 between the two periods. Again, it is confirmed that in normal operation the overall
407 conditions at the boiler outlet remains generally constant between contiguous 3-month
408 periods, thus the observed reduction of HCl and SO_2 mass flow for the 3 months of
409 DS testing can be reasonably ascribed to the effect of DS injection.
410
411
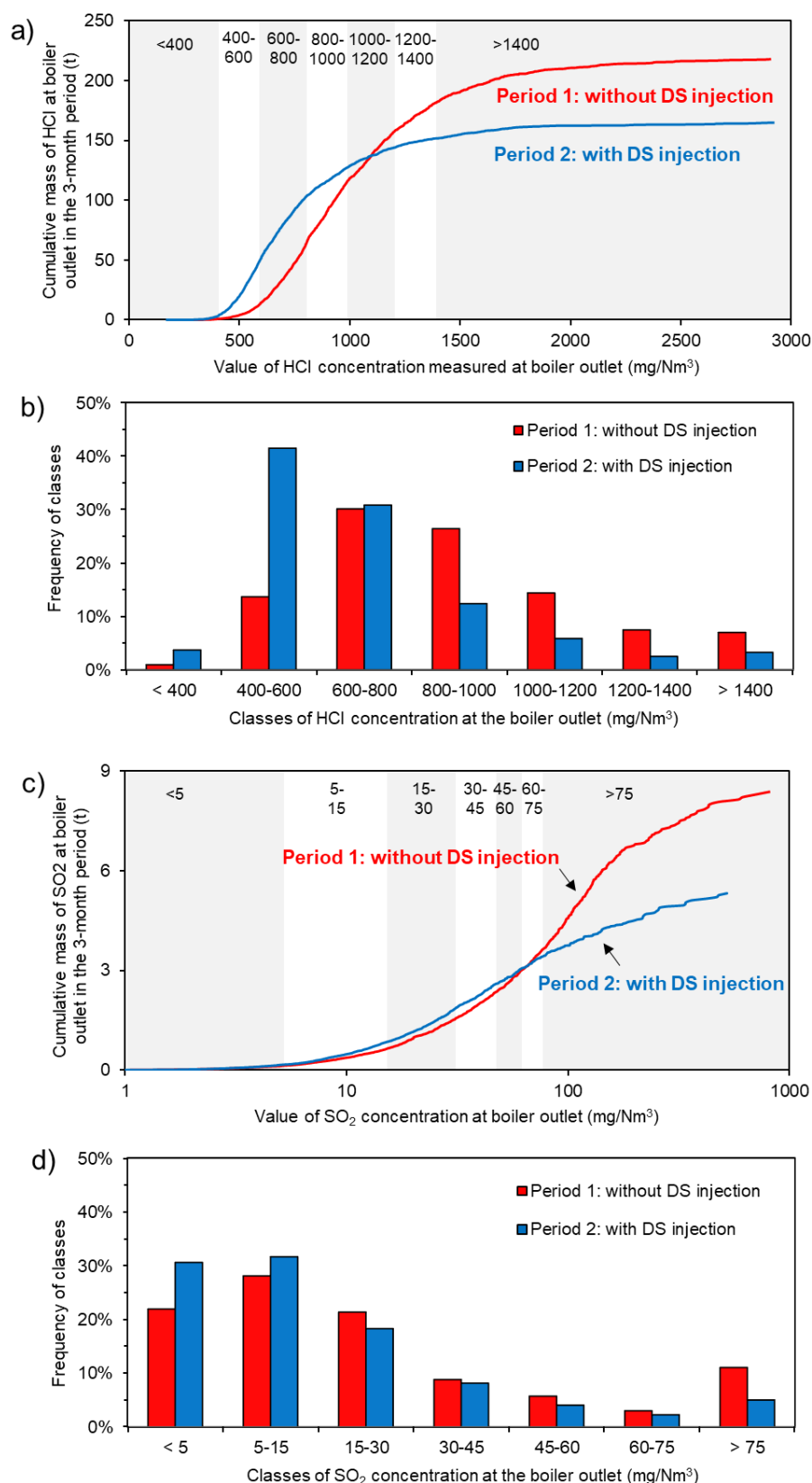
Figure 3. *Analysis of the effect of DS injection: a) cumulative mass flow of HCl at the boiler outlet in period 1 (without DS injection) and in period 2 (with DS injection); b) distribution of the hourly averages of HCl concentration at the boiler outlet in the two periods; c) cumulative mass flow of SO_2 at the boiler outlet in the two periods; d) distribution of the hourly averages of SO_2 concentration at the boiler outlet in the two periods.*

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4.2 Test runs in specific operating conditions: effect of DS feed rate

As introduced in section 2.3.1, the test runs at specific operating conditions were carried out varying the feed rate of the DS injected in the furnace. The objective of this specific set of test runs was to collect data with a finer time resolution than the macroscopic analysis of section 4.1 and to provide a quantitative evaluation of the relation between feed rate of reactant and removal efficiency of the acid pollutants. The test runs allowed collecting data on the difference in HCl and SO_2 concentration measured at the sampling point before and after DS injection at different feed rates, hence obtaining a dataset of HCl and SO_2 conversion at different SR of dolomitic sorbent injection. Tables S2 and S3 in section S3 of the Supplementary Material report the complete datasets obtained, which were used to calibrate the phenomenological model developed for acid gas conversion. The detailed results of all the test runs are also provided in section S3 of the Supplementary Material.
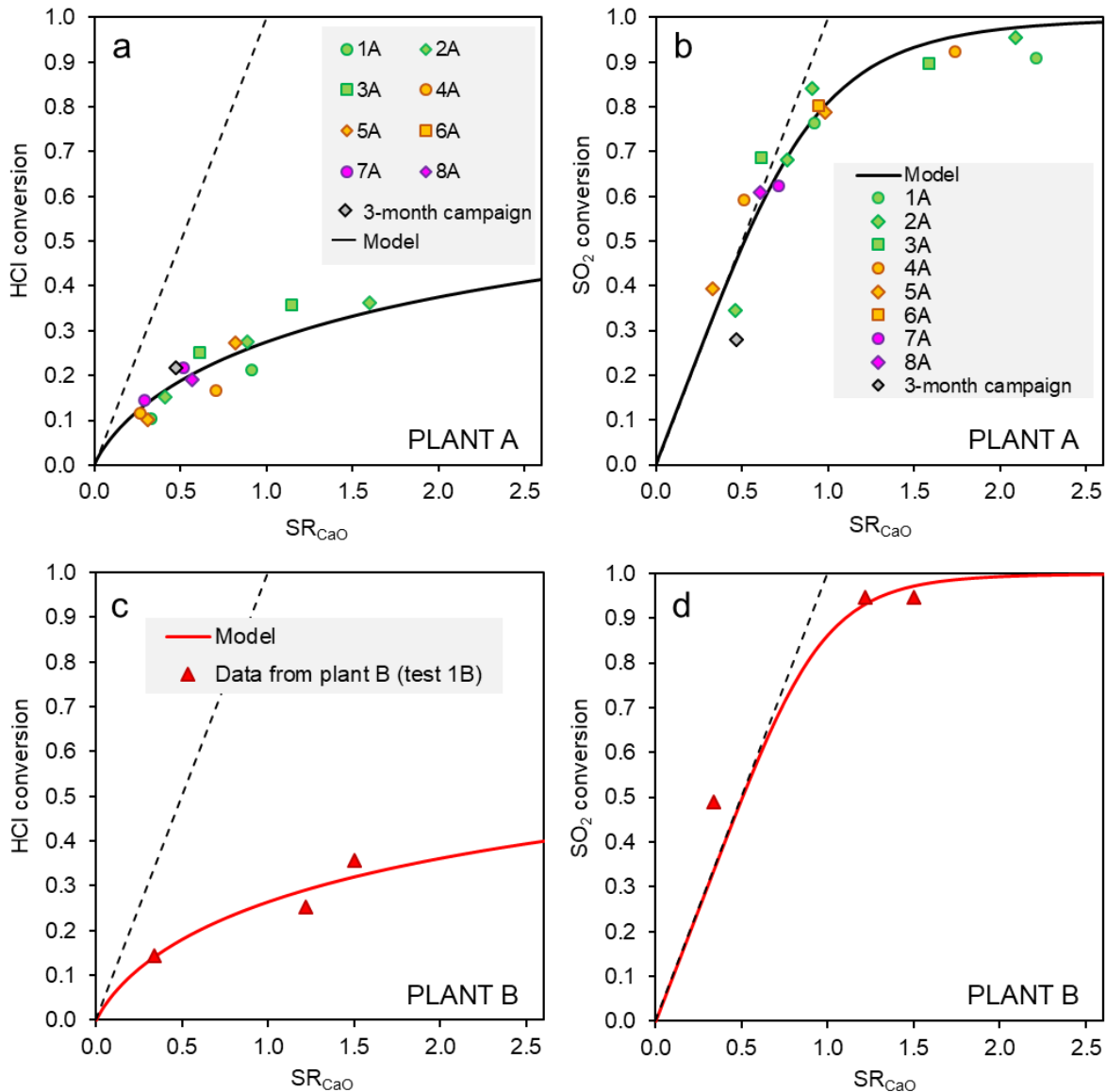
4.3 Model results

The datasets related to plant A and plant B obtained from the test runs were used to calibrate the phenomenological model presented in section 3.1 and to determine the values of parameter n for the two acid pollutants. As mentioned in section 3.1, the least trimmed squares method was adopted for concurrent outlier exclusion and model fitting. The following values of n were obtained for plant A: 1.38 for HCl and 5.26 for SO_2. The corresponding values of n obtained for plant B are: 1.36 for HCl and 7.29 for SO_2.

The fitting of calibration data points for plant A is shown in Fig. 4a and 4b, for HCl and SO_2 sorption respectively. Only the sample points after outlier removal are reported, while the complete dataset is presented in section S3 of the Supplementary Material. An additional data point representing the average performance of the dolomitic sorbent during the 3-month campaign analysed in section 4.1 is also reported in the figure, showing a quite good alignment with the data from the test run campaign. Fig. 4c and 4d reports the fitting of data points for plant B.

In relative terms, the dolomitic sorbent shows a markedly higher sorption capacity towards SO_2 than towards HCl. Hence, the DS behaves differently than the reactants injected at low temperatures (calcium hydroxide and sodium bicarbonate), which exhibit a higher reactivity towards HCl (see data in section S4 of the Supplementary Material). In absolute terms, the reactivity of the DS via furnace injection towards SO_2 is significantly higher than that of calcium hydroxide via low-temperature injection in the flue gas ductwork. Conversely, the DS appears less reactive than calcium hydroxide with respect to HCl.

Figure 4. *Phenomenological model results (continuous line) compared to calibration data for plant A: a) HCl removal, b) SO$_2$ removal; and plant B: c) HCl removal, d) SO$_2$ removal. The dotted line corresponds to stoichiometric efficiency.*

The model results for plant A and plant B were compared to each other in Fig. 5. The tests on plant B had the main intent to confirm the general trends identified on plant A. Indeed, the model curve obtained fitting the data points available for plant B appears quite similar to the model curve obtained for plant A. Clearly enough, the semi-empirical nature of the model implies that its calibration parameter, *n*, is site-specific, thus valid only for the description of the performance of the plant where the model was calibrated via test runs. However, the results in Fig. 5 confirm that the model has the potential to fit (after a specific tuning) experimental results across different WtE facilities, with different capacity, layout and average acid gas loads.

Fig. 5 also compares the model results of this study with data of DS performance collected in four WtE plants by Biganzoli et al. [30], the only other report on DS furnace injection available in the literature, to the best of our knowledge. The data of ref. [30] were obtained
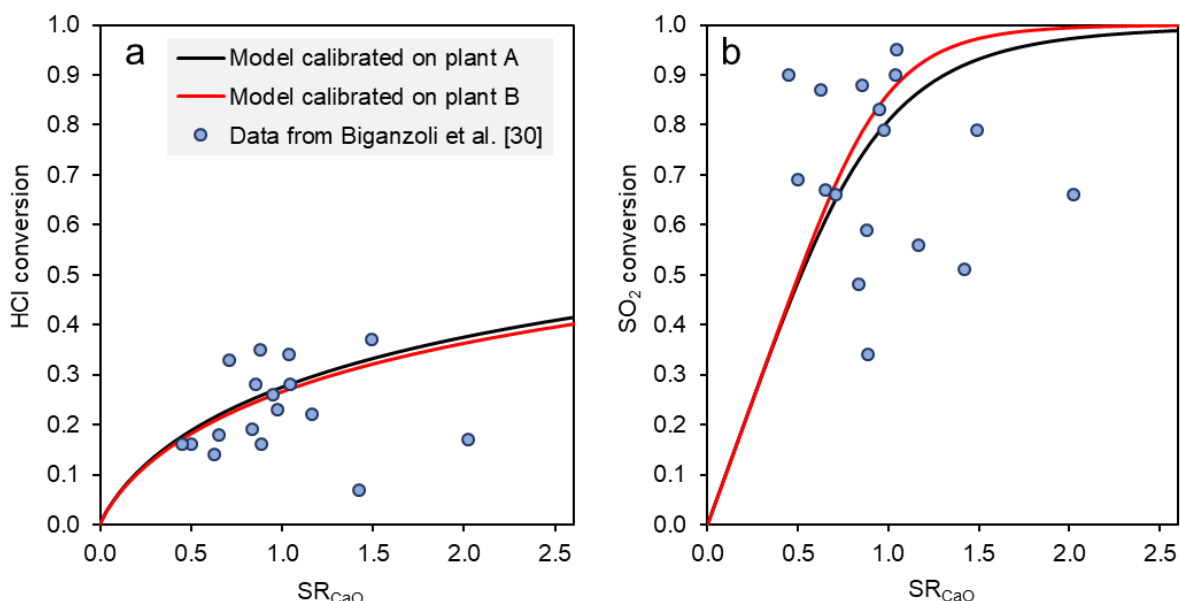
470   by calculating the SR of DS injection from the information on reactant feed rate and acid gas

471   concentrations given by the authors, under the assumptions of section 3.1.

472   At least for HCl, there is a fairly good agreement between the model results and the

473   experimental data set by Biganzoli et al. [30], even if no specific tuning of the model was

474   made. This further supports the comparability of HCl removal efficiency data by DS furnace

475   injection across different WtE facilities.

476   With respect to SO_2, the data by Biganzoli et al. [30] exhibit a much wider scatter than those

477   presented in Fig. 4b, arguably owing to the uncertainty associated to the lower and more

478   variable loads of this pollutant formed in waste combustion across different WtE plants. In

479   that regard, a merit of the test runs of the present study is to offer a robust and ample set of

480   homogeneous data on SO_2 removal by DS injection obtained in a single WtE plant (Fig. 4b).

481   In general, the analysis of the results in Fig. 5 support the robustness of the experimental

482   and modelling approach adopted in the present study.

483   In addition, the data of Fig. 4 are of clear interest for WtE operators, because only full-scale

484   testing can assess the performance of sorbent in presence of specific process conditions

485   (e.g. fluid-dynamics of the process in the furnace/boiler environment, interference between

486   gas species, morphological evolution of sorbent particles injected in the furnace) that are not

487   easily reproducible at laboratory scale. For instance, laboratory data available for the

488   dolomitic sorbent [29], collected in a fixed bed reactor at 900 °C, showed a conversion of

489   SO_2 in the range 78-82% (tests in SO_2/N_2 atmosphere, SO_2 inlet concentration of 1070

490   mg/Nm^3, SR_{CaO} = 2) and a conversion of HCl in the range 67-72% (tests in HCl/N_2

491   atmosphere, HCl inlet concentration of 1430 mg/Nm^3, SR_{CaO} = 2). The present full-scale

492   results demonstrated that the conversion of HCl that should be expected in a real furnace is

493   lower than the laboratory results (here, a SR_{CaO} = 2 corresponds to a HCl conversion slightly

494   lower than 40%; see Fig. 4). Conversely, the conversion of SO_2 registered at full-scale is

495   higher than that observed at laboratory scale (here, a SR_{CaO} = 2 corresponds to a SO_2

496   conversion higher than 90%; see Fig. 4), presumably owing to the significantly lower

497   concentration of SO_2 typically measured at plant (< 200 mg/Nm^3) compared to the value

498   adopted in laboratory, and a possible synergic effect on SO_2 sorption given by the presence

499   of HCl or its product of reaction $CaCl_2$ [41,42]. It is thus clear that the quantification of

500   removal efficiencies at full scale is necessary to obtain reliable data for the economic

501   assessment.

502

Figure 5. *Results of phenomenological models for plant A and plant B compared to literature data for: a) HCl, b) SO$_2$. The models were used without calibration to the specific operating conditions of plants where literature data were obtained.*

4.4 Optimisation of multi-stage acid gas removal processes

The developed phenomenological model allows addressing the minimisation of the operating costs of multi-stage acid gas removal processes. In the following, the cost-effectiveness of introducing a FSI of dolomitic sorbent was preliminarily assessed for both plants A and B, given the process constraints of section 3.2 and using the specific calibration data obtained for the model. In the economic assessment, only the costs for the procurement of sorbents and the management of residues were considered, being the most important operating cost entries in determining process economics [37,43].

4.4.1 Two-stage optimisation: dolomitic sorbent + NaHCO$_3$ (plant B)

The search for the optimal repartition of acid gas abatement between DS and sodium bicarbonate feed for a plant equipped with a furnace injection of dolomitic sorbent and a subsequent in-duct injection of NaHCO$_3$ (see the scheme of plant B, in Fig. 1b) was performed varying the conversion of HCl at the DS stage, while keeping constant the HCl concentration at stack. The results are shown in Fig. 6 in terms of overall operating cost (€ per Nm^3 of treated flue gas) as a function of HCl conversion by the dolomitic sorbent, for the different cases of flue gas composition introduced in Table 3. The minimum of each curve represents the optimal operating point for the corresponding inlet flue gas composition. Clearly enough, the cost associated to the operation of the acid gas removal system without DS injection is that corresponding to $X_{HCl,dolomitic\ sorb} = 0\%$.
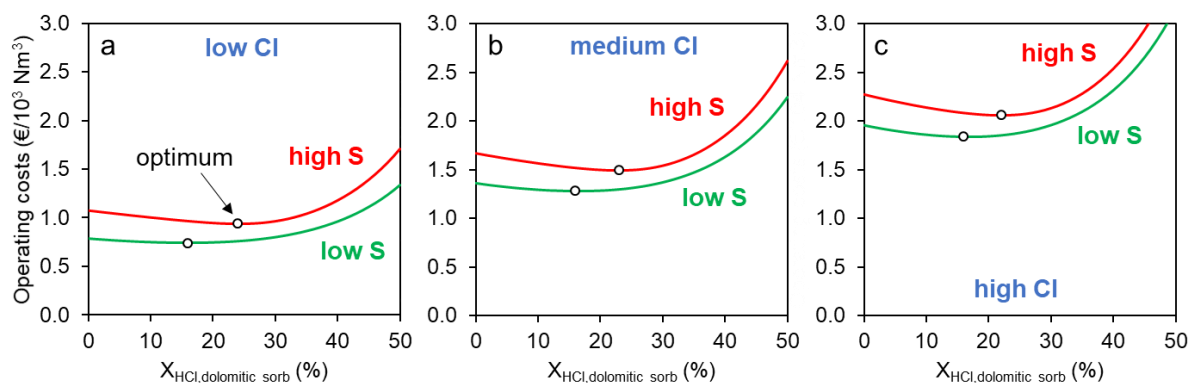
Figure 6. *Operating costs of the dolomitic sorbent + NaHCO$_3$ two-stage system as a function of HCl conversion in the DS stage for the six cases of flue gas composition in Table 3: a) low content of Cl; b) medium content of Cl; c) high content of Cl. Open circles: minimum cost.*

The curves for the six scenarios considered for flue gas composition exhibit a common behaviour. There is a region where variation of costs is limited ($X_{HCl,dolomitic\ sorb} < 30\%$), while costs start increasing steeply at higher $X_{HCl,dolomitic\ sorb}$ values. As shown by the open circles, an optimum is always present and it corresponds to a $X_{HCl,dolomitic\ sorb}$ value of 16 and 23%, respectively for the low S and high S scenarios. The cost savings of the optimal operating point of the DS + NaHCO$_3$ configuration with respect to the single stage NaHCO$_3$ injection are of about 6% for the low S cases. For the high S cases, instead, the savings range from 10% to 13%. The cost-effectiveness of the DS injection at higher SO$_2$ concentration and higher S/Cl ratio in the inlet flue gas stems from the higher relative reactivity of the dolomitic sorbent towards SO$_2$ compared to sodium bicarbonate.

Figure 7 details the contribution of the different entries to the overall operating cost for the scenarios of medium Cl and low or high S in the flue gas. The single stage NaHCO$_3$ system and the DS + NaHCO$_3$ configuration at the optimal operating point are compared. The addition of DS reduces the total cost related to reactant procurement, as the expense for DS and NaHCO$_3$ in the two-stage system is 11 and 20% lower than the expense for NaHCO$_3$ alone in the single stage configuration, respectively for the low and high S cases. Conversely, when DS injection is present, the cost associated to the management of process residues increases of 5 and 10% for the low and high S cases. This is due to the combined effect of a higher unit cost for the management route of DS residues (see Table 4) and the weight gain associated with acid gas capture: considering the sorption of HCl, a mole of dolomitic sorbent, upon dehydration and chloridisation (Eqs. 1 and 2), exhibits a 22% mass increase, while a mole of sodium bicarbonate, after thermal activation and chloridisation, undergoes a 69.5% net weight loss. On the whole, the net effect of DS injection at the optimal feed rate is the slight reduction of operating costs shown in Fig. 6, resulting from lower costs for reactants and higher costs for residues handling.
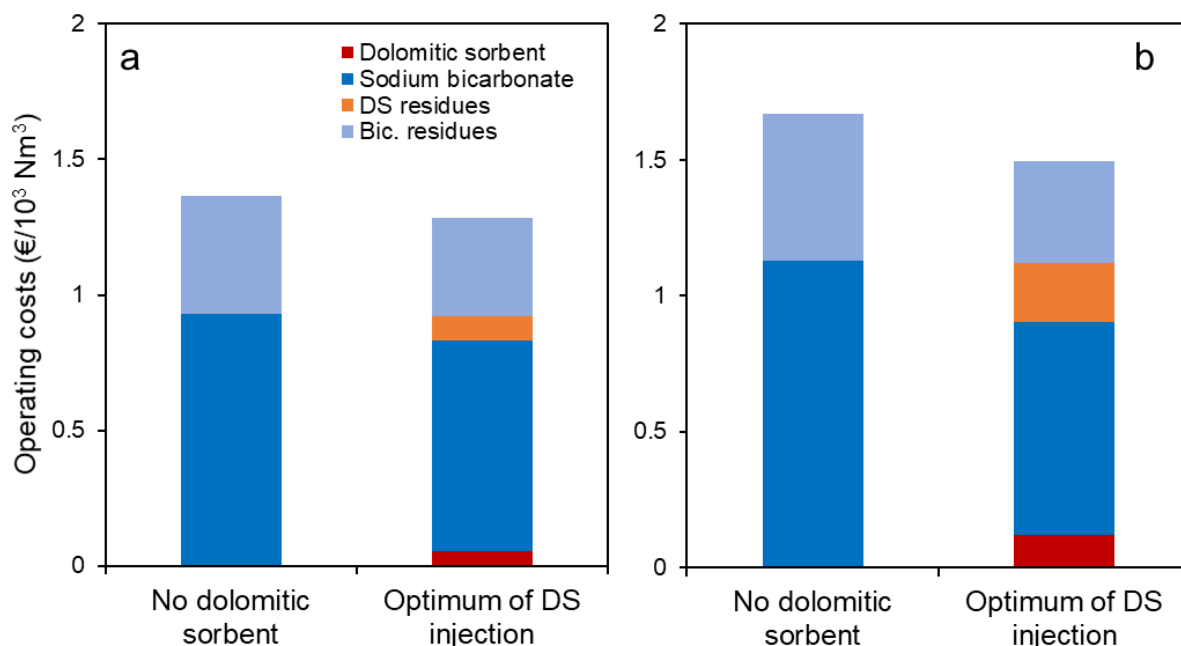
Figure 7. *Contribution of the different cost entries to the overall operating costs of acid gas removal for the single stage bicarbonate system (no furnace injection) and for the two-stage system with DS injection at the optimal operating point (optimum). Medium Cl and: a) low S, b) high S compositions were considered for flue gas.*

4.4.2 Three-stage optimisation: dolomitic sorbent + $Ca(OH)_2$ + $NaHCO_3$ (plant A)

For an acid gas treatment line as that of plant A, already presenting two stages of sorbent injection (calcium hydroxide and sodium bicarbonate), the introduction of DS injection forms a three-stage system. As such, the system exhibits two degrees of freedom in the selection of acid gas removal repartition between the stages, at a fixed overall HCl removal efficiency. Therefore, the evolution of the overall operating cost (€ per Nm^3 of treated flue gas) was explored as a function of HCl conversion in both the DS injection stage and the $Ca(OH)_2$ injection stage, as shown in the contour plots of Fig. 8.

As for the two-stage system, the optimal operating point of the three-stage system, pinpointed by a circle dot in the panels of Fig. 8, includes an injection of dolomitic sorbent, corresponding to 9-10% $X_{HCl,dolomitic\ sorb}$ for the low S cases and to 16-20% $X_{HCl,dolomitic\ sorb}$ for the high S cases. However, the cost savings associated to the three-stage optimum are in all scenarios lower than 10%, compared to the two-stage optimum of the $Ca(OH)_2/NaHCO_3$ system, which corresponds to values in the range 47-50% $X_{HCl,Ca(OH)2}$ when $X_{HCl,dolomitic\ sorb}$ is set to zero. Actually, a properly operated $Ca(OH)_2/NaHCO_3$ two-stage system is already an optimised system compared to the single stage injection of sodium bicarbonate [37], thanks to the use of cheaper calcium hydroxide (see again table 4) for the reduction of the acid gas concentration entering the bicarbonate stage. Thus, the margin for further cost reduction is lower.

Nonetheless, in operational terms, maintaining the optimal operating point of the $Ca(OH)_2/NaHCO_3$ two-stage system is not an easy task for the control system of a flue gas treatment line, as the concentration of inlet acid pollutants fluctuates widely and frequently. Therefore, DS injection can help because it enlarges the operational region where costs are within +10% of the optimum. As an example, focusing on panel 8b, the window of operating
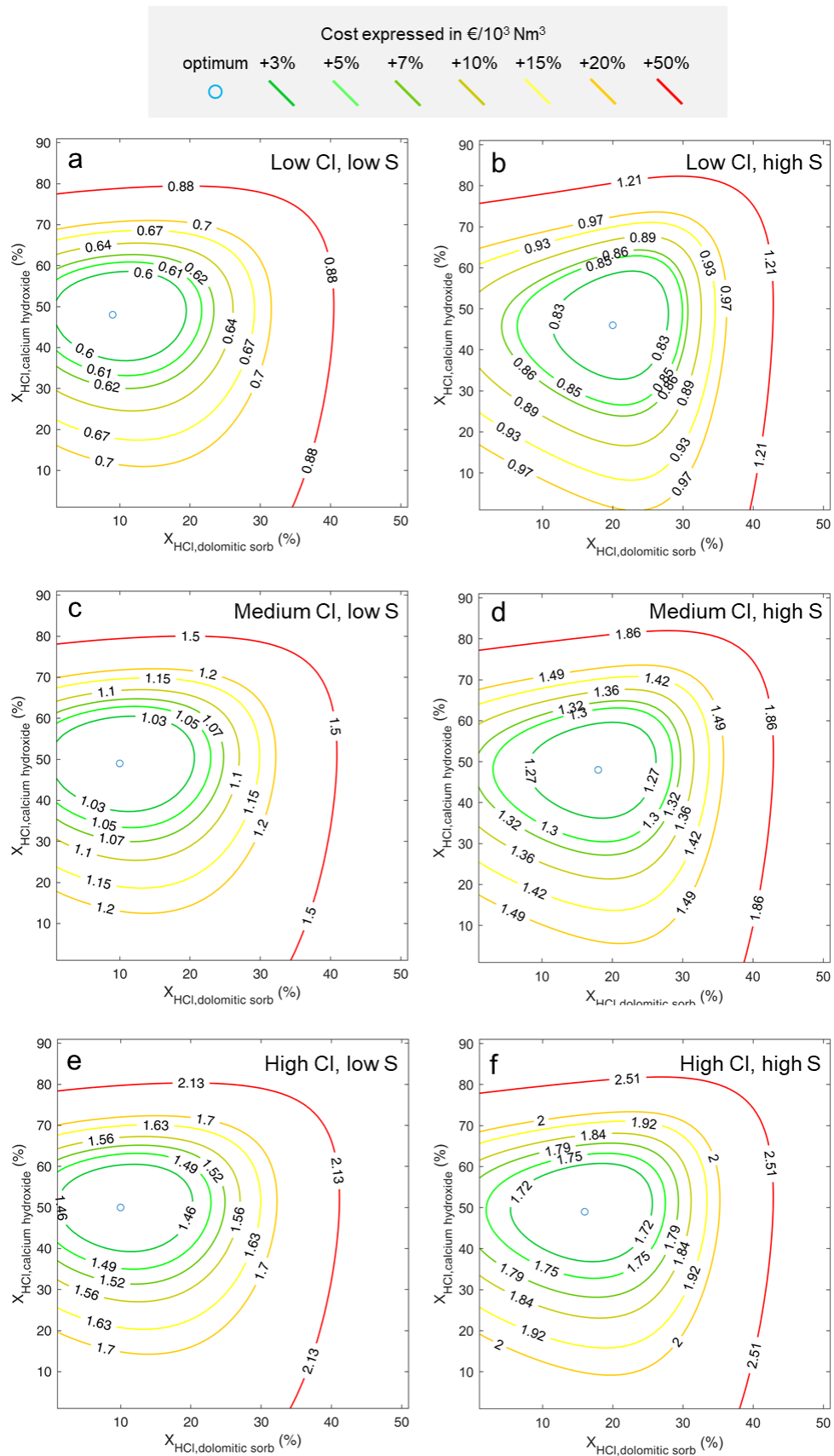
593 costs within +10% of the optimum, without DS injection, is comprised between $X_{HCl,Ca(OH)2}$
594 ranging from 35 to 55%, while for a DS injection corresponding to $X_{HCl,dolomitic\ sorb}$ = 20%, it is
595 possible to remain within +10% of the minimum operating costs even for $X_{HCl,Ca(OH)2}$ ranging
596 from 20 to 65%. Therefore, the addition of a dolomitic sorbent stage flattens the cost
597 variation related to the repartition of acid gas removal between the treatment stages,
598 possibly limiting the deviation from optimal operation.
599 Table 5 summarises the findings of sections 4.4.1 and 4.4.2, listing the specific feed rates
600 of dolomitic sorbent that correspond to the optimal $X_{HCl,dolomitic\ sorb}$ identified for
601 the configurations of plant A and plant B in the six cases of flue gas composition. It is
602 recalled that the conversion of HCl is linked to the SR of sorbent by the model of Fig. 4,
603 calibrated on the plant-specific test runs, and the SR of sorbent is linked to the mass feed
604 rate of sorbent by Eq. 11.

605

606 **Table 5.** *Specific feed rate of dolomitic sorbent (kg/10^3 Nm^3 of flue gas) corresponding to the*
607 *optimal $X_{HCl,dolomitic\ sorb}$ shown in Fig. 6 and 8.*

Cases of flue gas composition		System	
		DS + $Ca(OH)_2$	DS + $Ca(OH)_2$ + $NaHCO_3$
		Optimal DS feed rate (kg/10^3 Nm^3)	Optimal DS feed rate (kg/10^3 Nm^3)
Low Cl	Low S	0.39	0.20
	High S	1.04	0.53
Medium Cl	Low S	0.65	0.33
	High S	1.43	0.79
High Cl	Low S	0.90	0.46
	High S	1.75	1.05

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610
611

Figure 8. *Operating costs of the dolomitic sorbent + Ca(OH)₂ + NaHCO₃ three-stage system as a function of HCl conversion in the DS and Ca(OH)₂ stages, for the six cases of flue gas composition in Table 3.*

4.4.3 Effect of unit cost variability

The economic analysis in sections 4.4.1 and 4.4.2 is clearly influenced by the adopted unit cost values for the considered cost entries. To assess the effect of different unit costs than those listed in Table 4 on the economic optimisation, two scenarios were explored:

- an unfavourable scenario for the DS, named '*costly DS, cheap NaHCO₃*', in which the costs for the procurement of DS and the disposal of process residues are 20% higher than the values in Table 4, while the costs for the procurement of sodium bicarbonate and the costs for the recycling of Na-based process residues are 20% lower than the values in Table 4;

- a favourable scenario for the DS, named '*cheap DS, costly NaHCO₃*', in which – vice versa – the costs for the procurement of DS and the disposal of process residues are 20% lower than the values in Table 4, while the costs for the procurement of sodium bicarbonate and the costs for the recycling of Na-based process residues are 20% higher than the values in Table 4.

The ±20% range assumed is a conservative interval of variation, considering the relatively low differences in unit costs reported by several technical and scientific sources (see the overview of cost data by Dal Pozzo et al. [37]). The overall operating costs estimated for the systems without DS injection and for the systems with DS injection at the optimal operating point were compared for the different cost scenarios, as shown in Fig. 9. The comparison was carried out for both the two-stage and the three-stage system, with reference to medium Cl and high or low S flue gas composition   cases.
The first evidence is that the 'costly DS, cheap NaHCO₃' scenario always generates lower overall operating costs, while the upper end of the operating costs range corresponds to the the 'cheap DS, costly NaHCO₃' scenario. This finding is obvious for the cases with no DS injection. For the cases with optimal DS injection, it results from the much higher NaHCO₃ feed with respect to the DS feed rate in the optimal repartition of feed rate between stages, that results in a higher fraction of the total costs (as shown in Fig. 7). In the 'costly DS, cheap NaHCO₃' scenario, the slight cost savings ensured by DS injection, discussed in sections 4.4.1 e 4.4.2, are further reduced, and the operating costs of the systems with and without DS injection are almost equivalent. The exception is the case of the DS + NaHCO₃ system in presence of a flue gas composition with high S, for which the injection of DS still results in a 7% cost reduction. Vice versa, in the 'cheap DS, costly NaHCO₃' scenario the gap between the operating costs with and without DS injection is enlarged. The addition of DS results in savings, ranging from 9 to 16% for the NaHCO₃ system and from 5 to 12% for the $Ca(OH)_2$/NaHCO₃ system (lower and upper end of the range are for the flue gas compositions with low   and   high   S,   respectively).
Overall, these results clarify that, even in presence of uncertainty in the unit costs, the introduction of a DS injection stage reduces the variation of the total operating costs, as evident from the reduction of the cost brackets in Fig. 9. In the cost scenario that is unfavourable for DS, the overall costs of the system with or without DS injection are the same, while in the cost scenario that is favourable for DS, its injection generates appreciable savings. Therefore, from the viewpoint of WtE operators, adding a furnace injection of DS can increase the robustness of the entire acid gas treatment line.
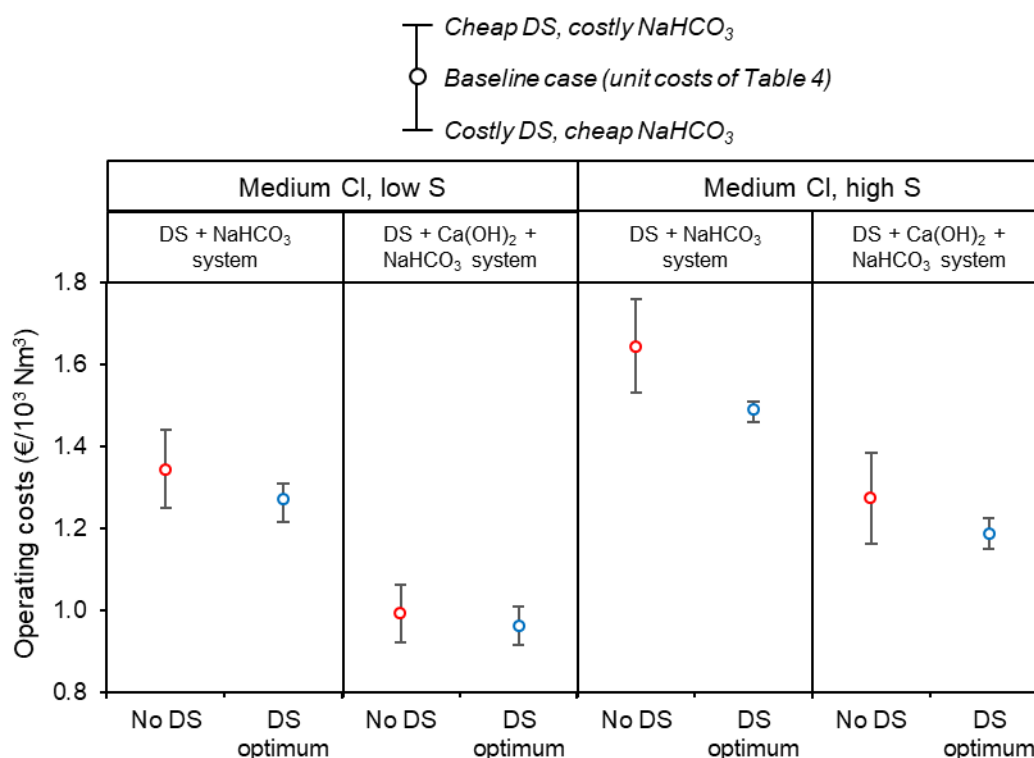
Figure 9. *Operating costs of acid gas removal in the different cost scenarios indicated in the legend. Comparison between no DS injection and DS injection at the optimum, for the DS + NaHCO$_3$ two-stage system and for the DS + Ca(OH)$_2$ + NaHCO$_3$ three-stage system with DS injection at the optimal operating point (optimum). Medium Cl and: a) low S, b) high S compositions were considered for flue gas.*

5. Conclusions

A quantitative analysis of the HCl and SO$_2$ removal performance of a dolomitic sorbent injected in the combustion chamber of WtE plants was carried out. Data obtained via dedicated test runs at full plant scale and their interpretation by means of a phenomenological model allowed deriving a non-linear relationship between the stoichiometric ratio of sorbent feed rate and the resulting conversion of acid pollutants. At the typical concentrations of HCl and SO$_2$ expected in WtE flue gas, a SR of sorbent equal to 1 was found to achieve a mean 27% HCl conversion and 83% SO$_2$ conversion, while a SR of 2 resulted in conversions of HCl and SO$_2$ respectively equal to 37% and 98%.

Since the introduction of a furnace injection of DS can be considered as an additional deacidification stage to the existing acid gas treatment line of a WtE facility, the developed model was used to identify the optimal feed rate of DS, *i.e.* the feed rate that minimises the overall operating costs of the line, for a WtE plant already equipped with a single or a two-stage acid gas treatment system. An optimum is generally present because the DS injected at high temperature, as opposed to calcium hydroxide or sodium bicarbonate injected at low temperature, has a higher relative reactivity towards SO$_2$ than HCl, hence it is complementary to the low-temperature sorbents, and it is cheaper than sodium bicarbonate, hence it is suitable as a pre-treatment step to limit the consumption of the more expensive reactant.

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In particular, for WtE plants operating with furnace DS injection plus a single downstream acid gas removal stage, the optimal operating point consists in demanding a conversion of HCl in the range 16-23% from DS injection. Such an optimised feed rate of DS can achieve a reduction of operating costs, which was estimated as high as 13% when the concentration of SO_2 in the raw flue gas is relevant.

For WtE plants operating with furnace DS injection plus a downstream two-stage acid gas treatment system, the optimal feed rate of DS corresponds to a conversion of HCl in the range 9-20%, depending on the SO_2 concentration in the flue gas. The savings that may be obtained thanks to the use of DS were estimated to be limited (less than 10% reduction of operating costs even for flue gas with high SO_2 concentration). However, the DS injection can still be a valid addition in order to stabilise the operation of the multi-stage acid gas removal process, smoothing the variability of acid gas load coming from the combustion chamber.

The advantages of DS injection are confirmed even in the presence of uncertainties in the unit costs considered in the economic analysis. Overall, the dolomitic sorbent injection emerges as a useful technology to control acid gas emissions, particularly as a cost-effective retrofitting option to single-stage removal processes.

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Supplementary Material

*Declaration of Interest Statement

Declaration of interests

☒ The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

☐The authors declare the following financial interests/personal relationships which may be considered as potential competing interests:

CREDIT AUTHOR STATEMENT

Techno-economic performance of HCl and SO$_2$ removal in waste-to-energy plants by furnace direct sorbent injection

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STATEMENT OF NOVELTY

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Direct furnace injection of a dolomitic sorbent may be a viable cost-effective technique to improve removal of acid gases in waste-to-energy plants, required by the oncoming new release of Best Available Techniques conclusions. However, apart from laboratory-scale results, no full-scale data is reported in the literature to assess the actual performance of such technology and optimal operational parameters. The present study, based on a full-scale campaign, carried out a first, systematic assessment of the effectiveness of adding furnace sorbent injection to different flue gas treatment processes, establishing a quantitative correlation between feed rate of reactant and resulting HCl/SO$_2$ conversion.