

DISSERTATION

Development of a flue gas scrubber system for sewage sludge incineration with biological scrubbing liquid regeneration

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unter der Leitung von

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1 Abstract

Mono-combustion of sewage sludge is conducted currently only in few big incineration plants with a capacities of more than 50,000 t/a each. The technology used for the combustion and cleaning of the flue gas is quite complex and an economic downscale is quite difficult. Therefore, small wastewater treatment plants do not burn the sludge on-site but send it to one of the big plants or to a waste incineration plant respectively a coal power plant for co-firing. But the transport of wet sludge, which contains approx. 75 wt% water, is not economic. Especially co-firing is critical from a sustainable point of view, as the fertilizer phosphorous, which is contained in the sludge, is lost in the ash. Therefore, simpler but just as clean technologies should be developed to increase the amount of on-site combusted sludge.

A key factor, in technical and economical respect, is the flue gas cleaning system, which represents more than half of the apparatus of a conventional plant and it is also responsible for the biggest amount of operational costs. In the current research project a flue gas scrubbing system for simultaneous removal of HCl, SO₂ and NO was developed. Effluent water from a wastewater treatment plant is used as scrubbing liquid in a combined quench/spray scrubber system. Afterwards the scrubbing liquid is sent back to the wastewater treatment plant for regeneration in bioreactors.

The scrubber system was installed and tested at the site of the sludge incineration plant Bad Vöslau. Experiments with varying liquid-to-gas ratios, several gas and liquid temperatures and different scrubbing liquid recirculation rates have

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been carried out to investigate the capabilities of the new scrubber system. As the flue gas composition was measured simultaneously before and after the scrubber system, the removal rates could be determined for each operational state.

In these experiments removal efficiencies of up to 100% for HCl, 95% for SO₂ and 55% for NO were achieved. It was found that an increased liquid-to-gas ratio and the scrubbing liquid recirculation has a positive effect. A major finding was that the NO removal efficiency depends on the nitrite content of the fresh scrubbing fluid. During the absorption process NO can react with nitrous acid or nitrite and form N₂O₃, which has a much higher solubility compared to NO. This behavior makes high NO removal rates possible without preliminary oxidation of NO to NO₂ in the gas phase.

Based on the experiences from the experimental series and the investigated absorption mechanism, it was found that the system is suitable for the intended purpose and can be used as a single flue gas treatment system after a dust removal system. With an eye on the effluent composition of the wastewater treatment plant, this system is appropriate also for other sludge incineration plants located on or near a wastewater treatment plant.

2 Kurzfassung

Monoverbrennung von Klärschlamm wird zur Zeit nur in einigen wenigen Verbrennungsanlagen mit Kapazitäten von jeweils zumindest 50,000 t/a durchgeführt. Die für die Verbrennung und nachfolgende Rauchgasreinigung verwendete Technologie ist relativ aufwendig und eine wirtschaftlich sinnvolle Übertragung auf kleinere Anlagen ist schwierig. Der Schlamm wird deshalb von kleinen und mittleren Kläranlagen nicht am Ort der Entstehung verbrannt, sondern zu einer der großen Anlagen oder zu einer Müllverbrennungsanlage bzw. einem Kohlekraftwerk zur Mitverbrennung gebracht. Der Transport von Nassschlamm mit einem Wassergehalt von ca. 75 wt% erscheint aber als wenig sinnvoll. Auch die Mitverbrennung ist in Hinsicht auf die Nachhaltigkeit kritisch zu betrachten, da der im Schlamm enthaltene Phosphor als Düngemittel verloren geht. Daher werden einfachere, aber genauso saubere Technologien entwickelt, die den Anteil des direkt auf der Kläranlage verbrannten Schlamms erhöhen sollen.

Eine Schlüsselrolle in technischer und ökonomischer Hinsicht nimmt dabei die Rauchgasreinigung ein, die mehr als die Hälfte des apparativen Aufwandes und der Betriebskosten einer konventionellen Anlage beansprucht. In diesem Forschungsprojekt wurde ein Rauchgasreinigungssystem für die gleichzeitige Abscheidung von HCl, SO₂ und NO entwickelt. Dafür wird Ablaufwasser einer Kläranlage als Waschflüssigkeit in einem kombinierten Quench/Sprühwäscher-System verwendet. Die verbrauchte Waschflüssigkeit wird zurück in die Kläranlage geleitet, wo sie in den Belebungsbecken wieder aufbereitet wird.

Eine Pilotanlage des Rauchgasreinigungssystems wurde bei der Klärschlamm-

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verbrennungsanlage Bad Vöslau errichtet. An dieser Anlage wurden Versuche mit verschiedenen Waschflüssigkeit/Gas-Verhältnissen, mehreren Gas- und Flüssigkeitstemperaturen und unterschiedlichen Rezirkulationsraten durchgeführt, um die Eigenschaften des neuen Systems zu untersuchen. Die Rauchgaszusammensetzung wurde gleichzeitig am Ein- und Austritt des Wäschers gemessen. Somit konnte zu jedem Zeitpunkt die Abscheideleistung in jedem Betriebspunkt bestimmt werden.

Während dieser Experimente wurden Abscheideleistungen von bis zu 100% bei HCl, 95% bei SO₂ und 55% bei NO gemessen. Es stellte sich heraus, dass sich sowohl steigende Waschflüssigkeit/Gas-Verhältnisse als auch die Waschflüssigkeitsrezirkulation positiv auf die Abscheideleistung auswirken. Ein wichtiger Einflussfaktor auf die NO Abscheidung ist der Nitritgehalt der Waschflüssigkeit. NO kann während des Absorptionsvorganges mit salpeteriger Säure bzw. Nitrit zu N₂O₃ reagieren, das eine viel höhere Löslichkeit in Wasser im Vergleich zu NO hat. Dieser Vorgang macht hohe NO-Abscheideraten ohne vorhergehende Oxidation des NO zu NO₂ in der Gasphase möglich.

Aufgrund der Erfahrungen aus dem Versuchsbetrieb und mit Hilfe des untersuchten Absorptionsmechanismus kann festgestellt werden, dass das System für die geplante Anwendung und als alleiniges Rauchgasreinigungssystem, nach einer vorhergehenden Staubabscheidung, geeignet ist. Unter Beachtung der Zusammensetzung des jeweiligen Ablaufwassers kann dieses System auch an anderen Standorten auf oder nahe an Kläranlagen verwendet werden.

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The increasing energy demand and higher environmental requirements force the development of new processes for sewage sludge utilization. Sewage sludge is the residue from the cleaning process of municipal and industrial waste water. As it contains lots of different compounds, depending on the origin of the waste water, there is no typical sewage sludge composition. Nevertheless, the compositions of sludge from municipal wastewater treatment plants are similar. Hence, also the utilization of municipal sewage sludge is quite similar.

In Europe, lots of sludge is utilized as a fertilizer in agriculture. In other continents sludge is stored in landfills or, in coastal areas, dumped into the sea. Both is not allowed in most European countries anymore and also usage in agriculture is declining due to stricter regulations. Many European countries already banned sewage sludge entirely from agricultural use, the sludge has to be treated otherwise. On the other hand pure industrial sludge contains lots of valuables and, therefore, it is mostly utilized in industry to recover those resources.

As an alternative to agricultural usage different combustion processes have evolved. Typically, the sludge is co-fired in coal power plants or waste incineration plants, but also processes for mono-combustions or gasification have been developed. These processes cannot make use of the fertilizing qualities of the sludge, but they can make use of the energy content of the sludge. The mono-combustion or gasification processes have the advantage, that phosphorus, which is the fertilizing component of the sludge, can be found in reasonable concentrations in the combustion ash.

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The bubbling fluidized bed is the most common technology for sewage sludge mono-combustion. Other technologies are rotary kilns or multiple hearth furnaces. Independent from the used combustion technology, flue gas treatment is a main part of a sludge incineration plant. The amount of equipment needed for the flue gas treatment often exceeds those of the combustion reactor and the heat recovery system itself. This causes significant costs during construction and operation of such plants and especially small plants have troubles to operate economically. As a consequence plants for mono-combustion have been built and successfully operated only in big cities with more than one million inhabitants.

Smaller towns have to transport their sludge to plants big enough to operate economically. Governmental programs were started to develop technologies for a sustainable sewage sludge treatment in decentralized plants to avoid this disappointing situation. These programs focus not only on a reduction of the transport effort, but also on the possibility of phosphor recycling. Countries like Germany and Switzerland have legal rules, which demand phosphor recycling from sewage sludge within the next years. On the other hand France is going to prohibit transport of sewage sludge out of the local district.

All these facts forced the further development of small to medium size sludge treatment systems and also this work shall contribute to this topic. During the current research project a scrubber system was developed, which not only removes acidic gases like sulfur dioxide or hydrochloric acid but also reduces the NO_x emissions significantly. The scrubber uses effluent water from a wastewater treatment plant as scrubbing liquid, which is regenerated in the main bioreactor of the wastewater treatment plant.

Beside the erection of a full size scrubber pilot plant at the site of the 1MW_{th} sewage sludge incineration plant Bad Vöslau, the main focus of this work is on the performance of the scrubber at different operating conditions. Another point of special interest is a better understanding of the chemical mechanisms occurring

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during the absorption process. This is essential for further development of the system and for application of the technology at other sites.

In the first part of the thesis the current state of the art of flue gas cleaning in actual sewage sludge incineration plants as well as other research contributing to this topic are reviewed. The second part describes the design of the scrubber pilot plant, also a brief introduction to the sewage sludge incineration plant Bad Vöslau is given. Finally, the results from the conducted experimental series are outlined and the gained informations about the absorption mechanisms, especially those of NO_x , are presented.

4 State of the art of flue gas cleaning after sewage sludge incineration plants

4.1 Basic operations for emission control [1, 2, 3, 4]

Each combustion process produces flue gases with different compositions. Beside the harmless parts oxygen, nitrogen and water, flue gases contain carbon dioxide, carbon monoxide, sulfur oxides, nitrogen oxides, halogens, unburned organic carbons, dioxins, heavy metals etc. and ashes. Most of them are toxic to humans, animals or plants, some are greenhouse gases which contribute to global warming.

Flue gas treatment is applied to remove harmful substances from these combustion flue gases. In contrast to primary measures, which avoid the formation of these substances, flue gas treatment reduces the emissions from combustion by separation of the pollutants.

Flue gas cleaning systems use different processes for separation of the pollutants from the gas stream. Two or more of these processes are often combined, as each method has its own pros and cons and the intended separation efficiency is reached only in combination of different methods. The unit operations used in flue gas cleaning are shortly characterized, industrially realized processes are described in the further sections of this chapter.

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Mechanical separation can be used for particles and liquids as their physical properties (size, inertia, electrostatic charge,...) differ completely from gases and are used to separate them from the gas stream.

Absorption is the separation of gaseous pollutants from the flue gas into a liquid solution. Depending on the solvent and the solved component physical or chemical absorption can occur.

Adsorption is the deposition of gaseous or liquid components on the surface of a solid by physical adherence or chemical bonding.

Condensation is the deposition of liquid drops, due to subcooling of the flue gas below the dew-point. During condensation other non-condensable components can be absorbed to the condensate.

Catalytic flue gas treatment enables or accelerates reactions of pollutants with other flue gas components or added reactants to harmless substances.

Biological flue gas treatment is the biological degradation of pollutants to harmless substances, mainly through bacterias, but also algae or others.

Membrane processes allow the selective separation of components, as some of them pass the membrane while others are retained. The separation effect is not only based on the size of the substance, but also on other effects like the diffusion velocity within the membrane.

Post-combustion can be used to reduce pollutants like carbon monoxide or organic carbons through oxidation to carbon dioxide and water.

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According to these unit operations different technologies have been developed and established for industrial applications. These technologies can be differentiated through their ability to separate particulate or liquid matter or gaseous components from the gas stream, as well as they could be divided into dry and wet flue gas cleaning technologies (Fig. 4.1).

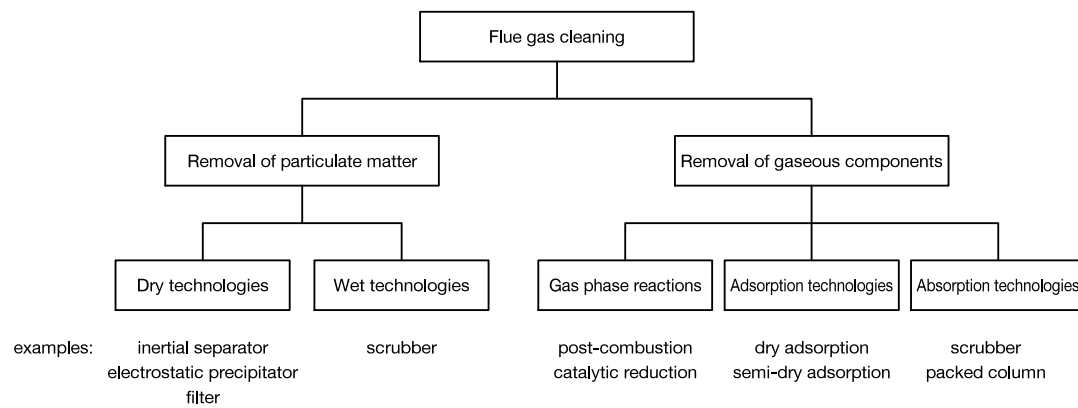


Figure 4.1: Classification of flue gas cleaning technologies

A typical flue gas cleaning system has equipment for dust collection, desulfurization, NO_x -reduction and for adsorption of heavy metals and PCDD/Fs. Some of them may be combined in one apparatus, but most of them need a special piece of equipment for each particular function. Conventional and biological flue gas treatment systems are equipped similarly, but as the microorganisms need a habitat where they can live, gas phase reactions are not possible.

4.2 Conventional gas cleaning technologies

4.2.1 Removal of particulate matter [1, 3, 5]

Dry technologies

Inertial separators use the different behavior of gas and dust particle, related to the density difference, when slowed down or accelerated.

A typical gravitational separator is as a *settling chamber* where the speed of gas stream is reduced and the particles have enough time to settle to the bottom of the chamber. The separation efficiency is directly depending on the length of the settling chamber, smaller particles have longer settling times and could therefore pass the chamber (Fig. 4.2).

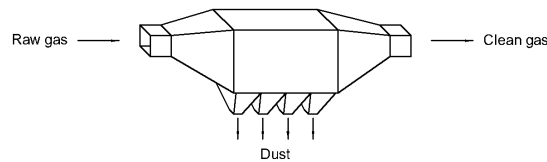


Figure 4.2: Settling chamber

In a *baffle chamber* the inertial force of the particles is used, as the direction of the gas stream makes a sudden change. The bigger the particles are, the higher is the percentage of particles which are not able to follow the stream and therefore settle down (Fig. 4.3).

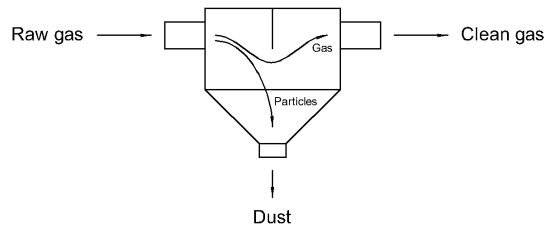


Figure 4.3: Baffle chamber

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In a *cyclone* the centrifugal force, created by the circular flow of the gas stream, transports the particles to the wall of the cyclone where they impact and further drop down to the dust container. A special type of cyclone is the multiclone where many small diameter cyclones operate in parallel and provide a high efficiency at an acceptable pressure drop (Fig. 4.4).

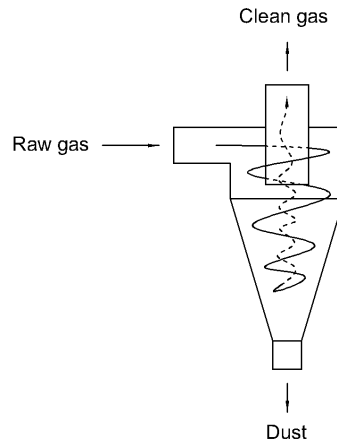


Figure 4.4: Cyclone

Filtrating separators In a filtrating separator the dust is collected during the passage of the gas stream through a porous media. This could be fabric, ceramics, metal sieves, sand,... depending on the requirements (e.g. particle size, temperature,...). In flue gas cleaning systems fabric filters (baghouse filters) and ceramic filter candles are common. Filtrating separators collect even smaller particles than according to their mesh size. Because of different diffusion and adhesive effects, particles are forced to adhere on the filtration media. These collected particles build up a filter cake which increases the filter efficiency, but also the pressure drop. Therefore, the filter cake has to be removed periodically by shaking the filter or counterflow of gas. Figure 4.5 shows a baghouse filter with reverse pulse jet cleaning technology.

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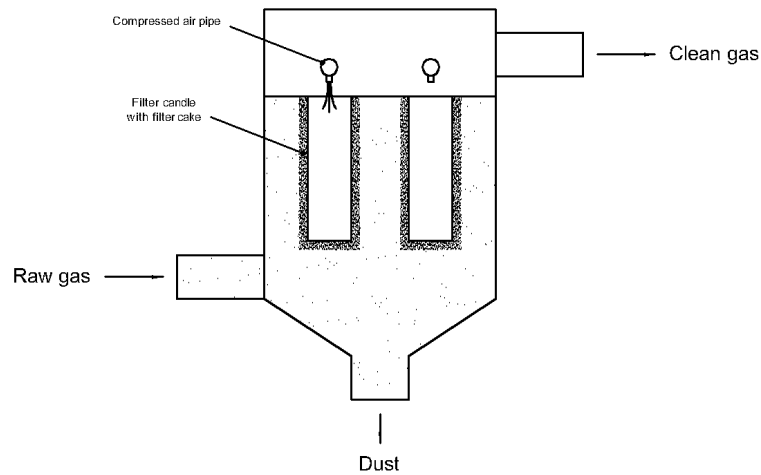


Figure 4.5: Baghouse filter

Electrostatic precipitators use electrostatic forces to remove the particles from the gas. The gas stream flows through an array of electrodes where a voltage of several kilovolts is applied in between. There the particles get loaded and then directed due to electrostatic forces to collection plates where they get deposited. Periodically, the collection plates are shaken and the particles settle down to the bottom of the precipitator.

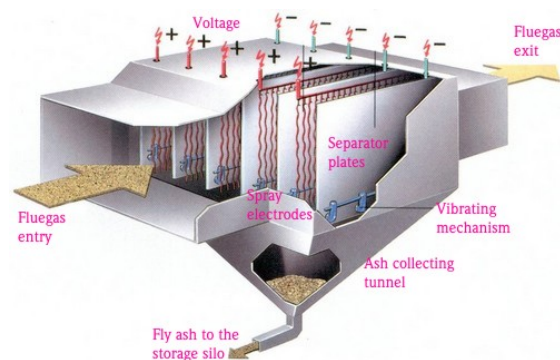


Figure 4.6: Electrostatic precipitator [6]

Wet technologies

Scrubbers In a wet scrubber the particles are suspended in the scrubbing solution and then the loaded solution is separated from the gas stream by gravitational or inertial forces. Typical types of wet scrubbers are spray towers, packed bed columns or venturi scrubbers (Fig. 4.7). They allow an intensive contact of gas and scrubbing solution, which is necessary for high removal efficiencies.

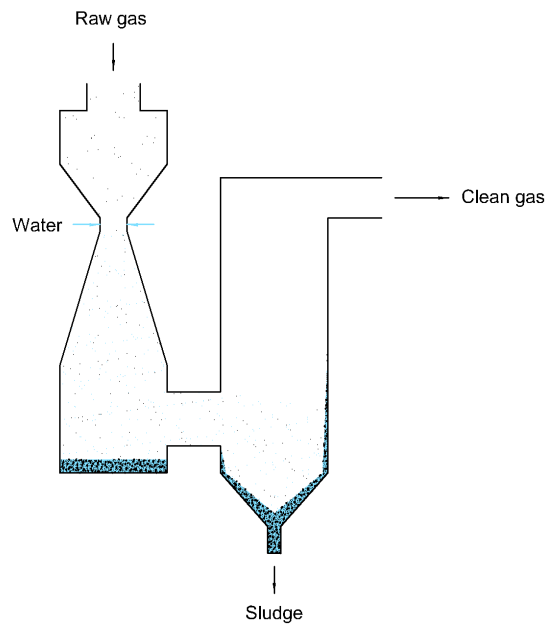


Figure 4.7: Venturi scrubber

4.2.2 Removal of gaseous components [1, 3, 7, 8, 9]

Beside the particulate matter also some of the gaseous components have to be removed from flue gas, the main components are sulfur oxides, nitrogen oxides, halogens and unburned organic hydrocarbons.

Sulfur oxides from combustion appear mainly as sulfur dioxide and only in minor concentrations as sulfur trioxide. The processes for SO₂ removal provide also sufficient separation efficiency for SO₃, therefore, specific cleaning methods are not necessary.

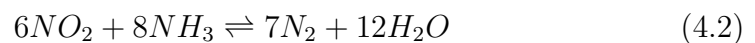
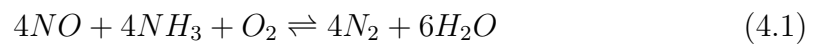
Also *hydrogen halides* are absorbed by the reactants for SO₂ and normally they will be removed simultaneously.

NO_x emissions can be lowered by primary methods like air staging or flue gas recirculation which avoid the formation. Alternatively nitrogen oxides can be removed by reaction with ammonia or urea (secondary measures). These methods have significant higher installation and operating costs and, therefore, they are only applied if primary measures are not adequate to reduce the NO_x emissions below the limits. In combustion gases nitrogen oxides appear mainly as NO, therefore, processes for NO elimination are the most important.

Hydrocarbons could be removed via catalytic oxidation if they could not be avoided through primary measures.

Gas phase reactions

Reduction of NO At appropriate conditions ammonia (or urea) reacts at high temperature or presence of a catalyst with nitrogen oxides to elemental nitrogen and water (Eq. 4.1 and 4.2). Both are natural components of the atmosphere and no further action is necessary.



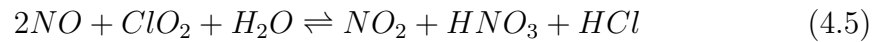
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Selective non-catalytic reduction (SNCR) takes place at temperatures between 850 and 950 °C. At higher temperatures the NH_3 reacts with the remaining oxygen to H_2O and NO , while at lower temperatures a slip of unreacted NH_3 can occur. The ammonia has to be inserted at the correct temperature level into the combustion chamber to meet these requirements, also a high stoichiometric ratio NH_3/NO is necessary to gain appropriate reduction rates. As secondary emissions of ammonia (slip stream) have to be kept low, only elimination rates of approx. 50-60% can be realized.

Selective catalytic reduction (SCR) is used at temperatures below 700 °C, typically between 300 and 500 °C, depending on the catalyst. Compared to the SCNR the usage of a catalyst allows higher utilization of the NH_3 and a lower slip, a stoichiometric ratio NH_3/NO of 0.8 to 1.1 is sufficient. Vanadium, molybdenum or wolfram are used as catalyst material, which are applied on a porous ceramic matrix, for example titanium oxide. At optimized conditions the NO_x elimination rate can reach up to 90%.

Oxidation of NO Instead of reduction also oxidation of NO to NO_2 can be used. Normally the reaction of NO with O_2 to NO_2 is quite slow and takes place in the atmosphere. O_3 or ClO_2 are used to accelerate this reaction (Eq. 4.3 - 4.5), alternatively also electronic beams promote the reaction.

In the atmosphere NO_2 reacts with OH -radicals to HNO_3 , which causes acid rain and, therefore, it has to be removed from the off gas.



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NO₂ has compared to NO a high solubility in aqueous solutions and can be removed by means of flue gas scrubbing. Alkaline chemicals like sodium hydroxide, ammonia or calcium hydroxide can be added to increase the absorption rate and to lower the water consumption. In this combination absorption rates of up to 90% can be achieved.

Alternatively NH₃ can be injected, which reacts with NO₂ to crystalline ammonium nitrate (NH₄NO₃), which is separated from the gas stream by means of a filter or other equipment for particle removal. This method is also suitable for simultaneous elimination of SO₂ which would react with ammonia to crystalline ammonium sulfate ((NH₄)₂SO₄). Both can be used as fertilizers. In combination with the electronic beam method for the oxidation step, elimination rates of more than 70% for NO_x as well as SO₂ can be reached.

Removal of organic hydrocarbons In combustion processes emissions of polychlorinated dibenzo-p-dioxine/-furans (PCDD/Fs) are of particular importance due to their extreme toxicity. They are formed in presence of oxygen, chlorine and aromatic compounds at temperatures between 200 and 700 °C. This means that they are formed after the combustion process in or after the flue gas heat exchanger.

The same catalysts like for NO_x reduction can also be used for catalytic decomposition of PCDD/Fs. If SCR is installed, simultaneous removal of both pollutants will be possible.

However most of the PCDD/Fs are adsorbed on dust particles and will therefore be removed with the particulate matter. The remaining part can be removed by adsorption or catalytic decomposition. Activated carbon is used for adsorption, which is applied in an entrained bed reactor, a fluidized bed reactor or a fixed bed adsorber. In dry and semi dry SO₂ removal processes the activated carbon can be added to the hydrated lime or limestone for simultaneous removal of both (for description of these processes see the next paragraph).

Adsorption technologies

Dry desulfurization For dry desulfurization alkaline substances like calcium carbonate (CaCO_3 , limestone), calcium oxide (CaO , burnt lime), calcium hydroxide ($\text{Ca}(\text{OH})_2$, slaked lime, hydrated lime) or dolomite ($\text{CaCO}_3\text{-MgCO}_3$) are used. They are induced into the combustion chamber or into the flue gas stream where they dissociate to CaO respectively MgO (Eq. 4.7 - 4.8). For calcium hydroxide the dissociation starts at temperatures above 400°C , while in the case of carbonate or dolomite minimum temperatures of $700\text{-}800^\circ\text{C}$ are required. Therefore, carbonate and dolomite are used only directly in the combustion chamber, while calcium hydroxide and calcium oxide can also be injected into the cooled flue gas stream.



The oxides, formed during the dissociation process, react with the sulfur oxides to CaSO_4 (anhydrite) and MgSO_4 (Eq. 4.9 and 4.10).



Figure 4.8 shows the temperature dependency of the equilibrium partial pressure of SO_2 . It is clearly visible that at higher temperatures only calcium compounds can be used while magnesium compounds are not suitable to bind SO_2 at temperatures above 800°C .

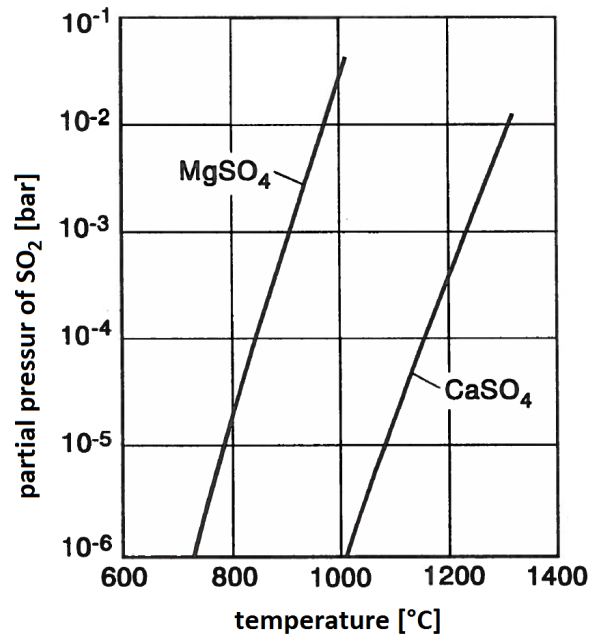


Figure 4.8: Equilibrium partial pressure of SO₂ [7]

Desulfurization in the combustion chamber allows efficiencies of up to 60% with carbonate and 70% with calcium hydroxide. Here a stoichiometric ratio of 2 is needed, because of the relatively short residence time of the calcium particles. Efficiencies of up to 95% can be reached with residence times of several minutes as they occur in fluidized bed adsorbers. Entrained particles are separated from the flue gas stream together with the fly ash by means of a filter or an electrostatic precipitator.

If hot desulfurization is not suitable, hydrated lime will also be used at lower temperatures, around 150 °C. At this low temperature the dehydration process does not occur and the Ca(OH)₂ reacts directly with SO₂ to CaSO₄ · 2 H₂O (gypsum). The efficiency of this reaction is not as high as it is at high temperatures. Furthermore, also water is needed for the reaction as shown in reaction eq. 4.11

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and 4.12 . If the flue gas does not contain enough water, it has to be conditioned by injection of a water spray. Methods to extend the residence time of the lime have been developed to reach higher efficiencies. This can be obtained by means of recirculation of the ash from the filter back into the flue gas stream.



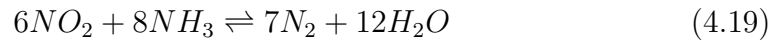
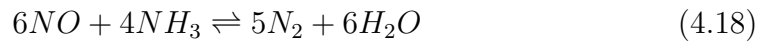
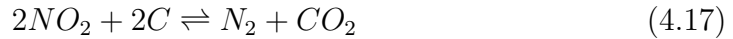
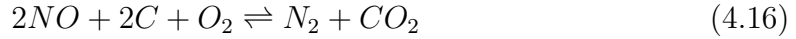
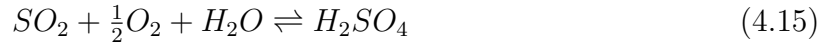
Semi-dry desulfurization For the semi-dry process the hydrated lime or limestone is introduced into the flue gas stream as a suspension with water. Then the SO_2 is absorbed by the water droplets and reacts with the lime (Eq. 4.13 and 4.14) while the water is evaporated. This process provides higher efficiencies compared to the dry process at low temperatures. After the residues (gypsum) are dried, they can be removed from the gas stream in the same way like in dry processes.



Adsorption to activated carbon In the so called "Bergbauforschung/Uhde"-process flue gas is led into an adsorber, where at temperatures of 100-150 °C SO_2 and NO_x are bond to the activated carbon. There SO_2 is oxidized to sulfuric acid while NO_x is reduced to elemental nitrogen and carbon dioxide or through addition of NH_3 to elemental nitrogen and water (Eq. 4.15 to 4.19). As a side

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product, also ammonium sulfate $((\text{NH}_4)_2\text{SO}_4)$ is formed (Eq. 4.20). The activated carbon can be regenerated by means of thermal desorption where a SO_2 enriched gas is produced.



Absorption technologies

Wet desulfurization In contrast to the previously mentioned processes in a wet scrubber the sulfur dioxide is absorbed from the the scrubbing fluid and remains dissolved until the scrubbing fluid is regenerated.

If the process is designed as a *physisorption* process, the SO_2 will be dissolved according to the phase equilibrium. The scrubbing solution can be regenerated by simply heating the fluid or by stripping with steam or air. Typical scrubbing fluids for physisorption are tributylphosphate (TIBP), tetraethyleneglykoldimethylether (TEG-DME) or triethyleneglykoldimethylether (TIEG-DME). Figure 4.9 shows the phase equilibrium of SO_2 in different solvents, while Figure 4.10 shows the temperatur dependency of the phase equilibrium of SO_2 and TEG-DME.

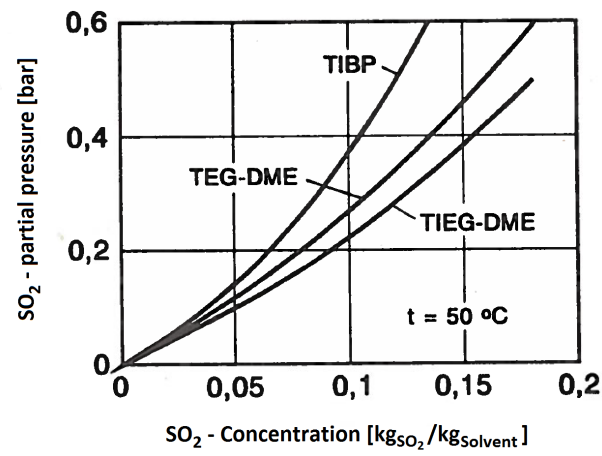


Figure 4.9: Phase equilibrium of the system SO_2 and different organic solvents (TIBP, TEG-DME, TIEG-DME) [7]

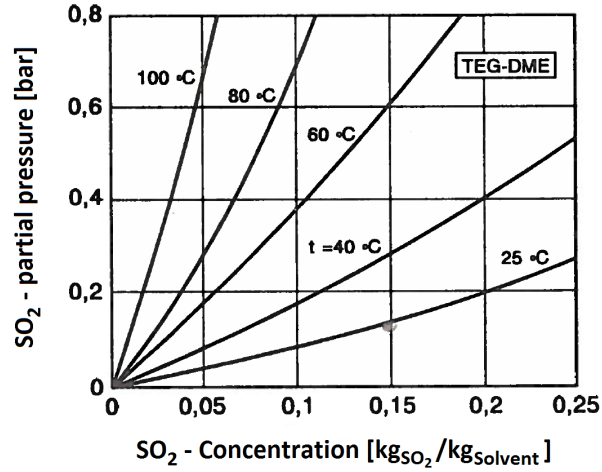
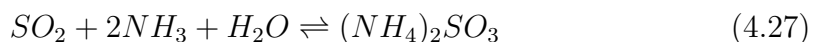
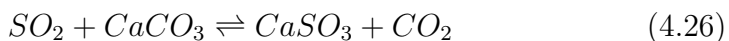
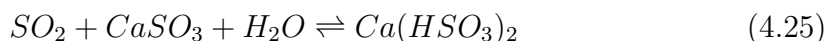
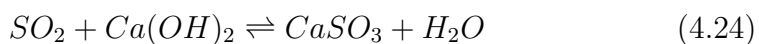
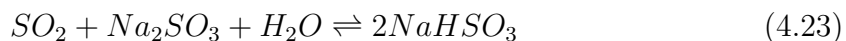
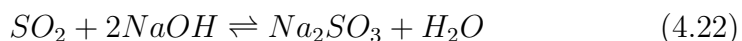


Figure 4.10: Phase equilibrium of the system SO_2 and TEG-DME at different temperatures [7]

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In a *chemisorption* process the dissolved SO_2 reacts with the scrubbing fluid. This allows much higher concentrations in the scrubbing fluid compared to the physisorption process, but on the other hand regeneration of the scrubbing fluid is much more difficult if not even impossible. Typically, aqueous solutions with sodium hydroxide, ammonia, hydrated lime or limestone are used as solvent. The absorbed SO_2 dissociates in water to sulfurous acid (Eq. 4.21) and further reacts with the alkaline components to different sulfites (Eq. 4.22 - 4.28)



If sodium hydroxide is used, the scrubbing solution can be regenerated via evaporation of SO_2 or addition of hydrated lime which will react to gypsum. The regeneration will not work if the sulfite or hydrogen sulfite has reacted with oxygen to sulfate.

In a limestone or hydrated lime scrubber the sulfites are further oxidized to sulfate. They could be separated from the solution by sedimentation as the solubility of calcium sulfate is very low. The poor solution can be reused as it contains still unreacted lime, also fresh lime can be added.

If ammonia is used for the scrubbing process, the solution will contain ammonia sulfite and ammonia hydrogen sulfite which are oxidized to ammonia sulfate.

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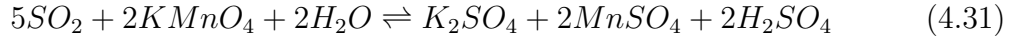
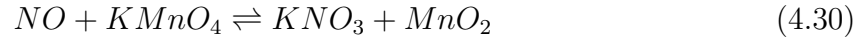
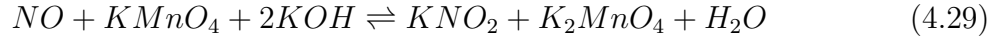
The solution is led to a spray dryer where unreacted ammonia and sulfur dioxide evaporates, while pure ammonia sulfate remains, which can be used as a fertilizer.

NO_x scrubbing Compared to the previously mentioned technologies for NO_x reduction, which convert the harmful substances to either to nitrogen and oxygen or to NO₂, scrubbers absorb NO and NO₂ in an aqueous solution. According to the low solubility of NO in pure water, the sorption can be implemented as chemisorption or combined with an oxidation process, to convert NO to NO₂ (see also 4.2.2, Oxidation of NO) to gain sufficient elimination rates. All these wet NO_x treatment technologies have not reached broad usage in industry until now but they are continuously improved and also the current scientific work shall contribute to this topic.

The key factor to high absorption rates is to increase the mass transfer coefficient between gas and solution. In the case of NO the diffusivity in the liquid phase is the limiting factor and must, therefore, be increased. This can happen by addition of an oxidizing reagent, for example potassium permanganate (KMnO₄) or hydrogen peroxide. Alternatively coordination compounds like iron-ethylenediaminetetraacetic acid (Fe²⁺-EDTA) also increase the absorptivity.

Wet oxidation with alkaline KMnO₄ produces potassium nitrate and potassium nitrite, presence of sulfur dioxide leads also to production of potassium sulfate and, therefore, increasing the usage of the solvent (Eq. 4.29 - 4.31). The NO absorption rate is between 25 and 50% which is quite low. Also acidic KMnO₄ or acidic NaCl can be used, where absorption rates of up to 99% can be reached, but in this case there are secondary emissions of NO₂ and, therefore, a second scrubber stage would be necessary.

4 State of the art of flue gas cleaning after sewage sludge incineration plants



Application of an aqueous Fe^{2+} -EDTA solution allows simultaneous elimination of SO_2 and NO from the flue gas, the reaction paths are shown in Figure 4.11. Beside the intended NO and SO_2 elimination, many other secondary reactions appear which increase the EDTA consumption. That leads to formation of inactive Fe^{3+} -EDTA, N_2O and different sulfur compounds. Therefore, a high scrubbing solution stream is necessary for NO_x absorption rates of above 70%, which also leads to nearly total elimination of SO_2 .

Cooled crystallization towers are used for regeneration of the solution, where the sulfur compounds are separated. The inactivated Fe^{3+} -EDTA is reactivated to Fe^{2+} -EDTA by heating to 90-100 °C.

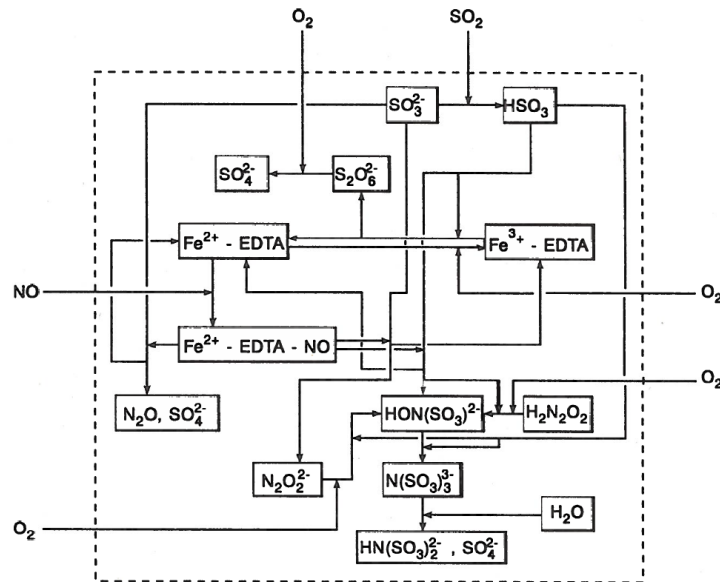


Figure 4.11: Reactions during the absorption of NO and SO_2 in an aqueous Fe^{2+} -EDTA solution [7]

Removal of halogen compounds Elemental halogens have a quite low solubility in water and, therefore, scrubbing with pure water is not suitable. If alkalines like NaOH or $\text{Ca}(\text{OH})_2$ are added, the solubility increases significantly and proper absorption rates can be reached. Figure 4.12 shows the phase equilibrium of chlorine and water at different temperatures.

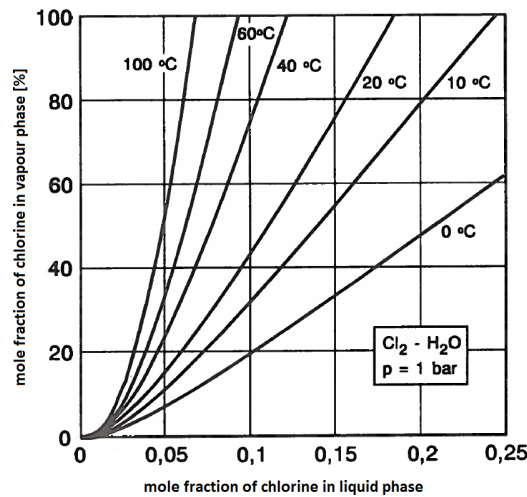


Figure 4.12: Phase equilibrium of the system Cl_2/water [7]

Hydrogen halides have a very high solubility in pure water, therefore, absorption to plain water is the method of choice. Figure 4.13 shows the phase equilibrium of hydrogen chloride and water at different temperatures. It is clearly visible that the solubility of the halides are some orders of magnitude higher than the solubility of the elemental components.

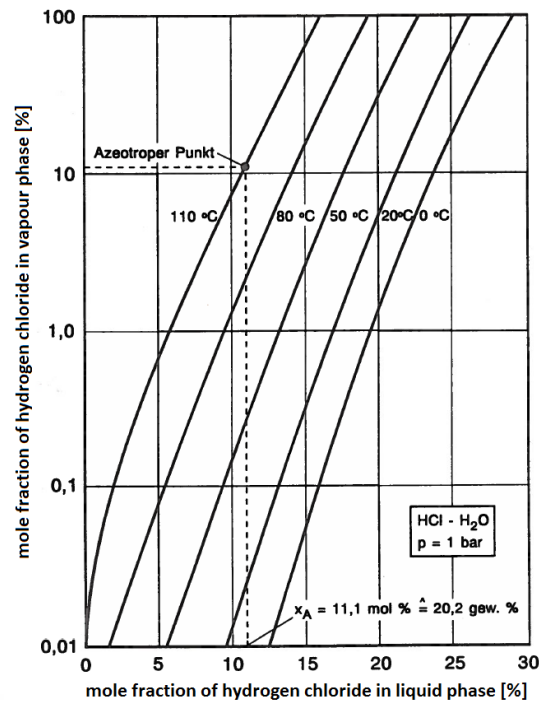


Figure 4.13: Phase equilibrium of the system HCl/water [7]

4.3 Biological flue gas cleaning technologies

[5, 7, 10, 11]

Like conventional flue gas cleaning systems, biological flue gas cleaning systems need equipment for removal of particulate and gaseous pollutants. The particulate matter is removed in the same way like in conventional systems and therefore the methods are not described again. The gaseous pollutants are adsorbed to a fixed bed or absorbed to a scrubbing liquid. Then microorganisms are used for the degradation of the pollutants. These microorganisms grow in a humid habitat, which can be either a water film on a solid carrier or an aqueous solution. Therefore, only adsorption or absorption processes can be realized, but no gas phase processes (compare Fig. 4.1). In the first case the flue gas is led over a packed

bed and the pollutants are adsorbed to the bed material and further converted by the microorganisms. This system is commonly known as biofilter. In contrast to this system is the bioscrubber system, where the pollutants are absorbed into the scrubbing solution like in a conventional scrubber and the scrubbing solution is regenerated by microorganisms living on the packed bed of the scrubbing column or in a separate bioreactor.

4.3.1 Biofilters

A typical biofilter is built as a fixed bed reactor. The flue gas is inserted from the bottom, passes the bed and exits on the top. As bed material and carrier for the microorganisms normally wood or bark chips, foliage, compost, peat or similar biomass is used (Fig. 4.14).

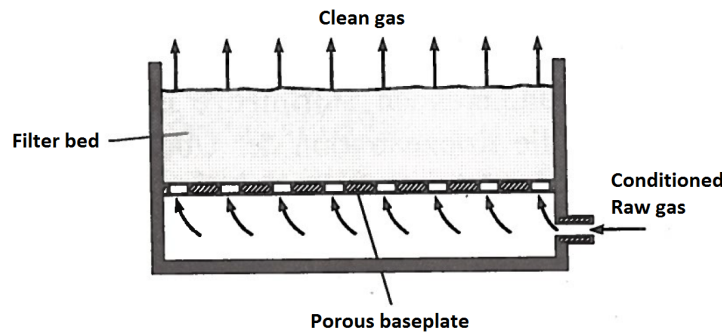


Figure 4.14: Typical setup of a biofilter [7]

The pollutants are absorbed into the humid film on the surface of the biomass, where the degradation takes place. The microorganisms degrade not only the pollutants but also the biomass, therefore, the bed has to be replenished or renewed periodically. To avoid drying out of the biomass, the flue gas stream has to be conditioned with water as well as water spray onto the top of the package is necessary. Biofilters of this type are commonly used for treatment of organic

pollutants, especially for odorants. The reaction conditions depend on the type of microorganisms used for degradation of the specific pollutant and their preferred environmental conditions.

4.3.2 Bioscrubbers

Bioscrubbers use the solubility of pollutants in water for removal from the gas stream. Similar to conventional scrubbing systems the scrubbing solution has to be regenerated before it can be used again. This regeneration is carried out by the microorganisms. Depending on the biodegradability of the pollutants two process types can be realized.

For fast degradable substances a trickling filter is used, which consists of a packed bed where the microbial biofilm grows. Like in a packed bed scrubber the pollutants are absorbed in the downwards flowing water stream and converted by the microorganisms. At the bottom of the column the water is collected and returned to the top, where also nutrient solution for the microorganisms can be added (Fig. 4.15).

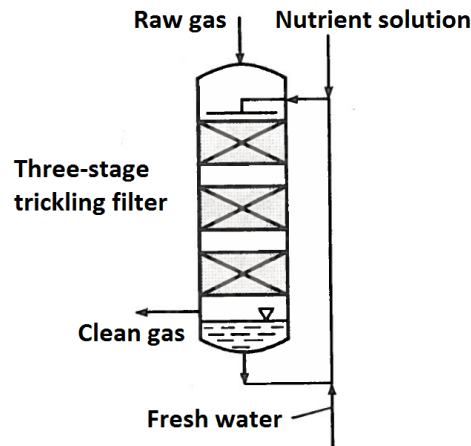


Figure 4.15: Typical setup of a bioscrubber/trickling filter [7]

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If the conversion of the pollutants needs longer time, the absorption of the pollutants and the regeneration of the scrubbing fluid takes place in different reactors. For absorption a simple packed bed column or a spray tower is used, while the bioreactor for the scrubbing fluid regeneration normally uses the activated sludge process (Fig. 4.16). The separation of both processes allows different residence times of the scrubbing liquid in the column and in the bioreactor, which enables the time consuming degradation of the pollutants by the microorganisms.

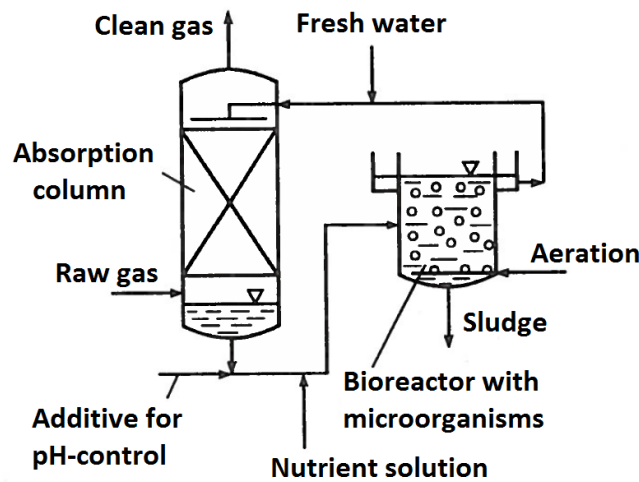


Figure 4.16: Typical setup of a bioscrubber with external regeneration [7]

Bioscrubbers with external regeneration can be used for all types of water soluble substances. The absorption column can be designed like conventional scrubbers for water soluble substances, while the bioreactor has to be designed following the design rules for biological sewage treatment plants. Also existing bioreactors can be used, e.g. if the scrubber is erected on the site of a sewage treatment plant. This technology is used in the current research project, therefore a closer look on the mechanisms of NO_x absorption and literature on different types of bioscrubbers is taken in the next chapter.

5 Review of absorption technologies using water as a scrubbing liquid

5.1 Mechanisms of NO_x, SO₂ and HCl absorption into water

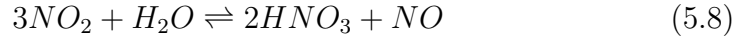
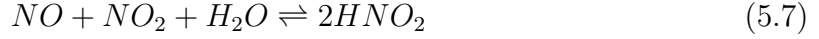
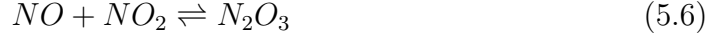
Sulfur dioxide and hydrogen chloride have a high solubility in water, the usage of aqueous scrubbing for removal of these components are well known and widely used as mentioned in chapter 4.2.2. In water, HCl dissociates to H⁺ and Cl⁻ (Eq. 5.1). SO₂ reacts to sulfurous acid (Eq. 4.21), which also dissociates to H⁺ and SO₃²⁻ (Eq. 5.2 and 5.3).



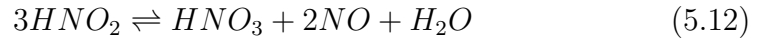
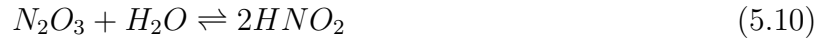
The absorption process of nitrogen oxides (NO and NO₂) is based on several reactions, which proceed in both, the gas and the liquid phase. One key reaction is that nitrogen monoxide is oxidized to nitrogen dioxide (Eq. 5.1). This can be forced by cooling of the gas stream to a temperature below 50°C combined with a sufficient O₂ concentration, like it is performed during the Ostwald process for nitric acid production [12]. The solubility of NO₂ in water is one order of magnitude higher than the solubility of NO and, therefore, also the absorption efficiency increases substantially.

5 Review of absorption technologies using water as a scrubbing liquid

Considerable reactions for absorption of NO_x to water and aqueous solutions were found in [13, 14, 15, 16, 17, 18], the gas phase reactions are:



In the liquid phase the following reactions are mentioned:



As already mentioned the oxidation of NO (Eq. 5.1) is one of the most important reactions. For temperatures below 350°C it can be assumed that the equilibrium is on the left side, but it is kinetically controlled and, therefore, the residence time is a key parameter for high conversion rates. The reactions (5.2) to (5.5) are known as prompt reactions and, therefore, not critical.

Not only NO_2 but also the other products of the reactions (5.2) to (5.5) have a higher solubility in water which reduces the needed amount of scrubbing solution to remove a certain amount of NO.

The liquid phase reactions describe the formation of nitrous and nitric acid. Like the oxidation of NO they are kinetically controlled.

5.2 Experimental experience with aqueous NO_x scrubbing

Compared to the usage of water for sulfur dioxide and hydrogen chloride elimination, removal of nitrogen oxides from flue gases using water as a scrubbing solution is not common and rarely used in industrial applications. Therefore, literature about research and experimental activities related to this topic was reviewed in detail. A summary of this literature is given in the following sections.

5.2.1 Experiments with packed scrubbers

Suchak et. al. [19] studied the absorption of NO_x from off gas of an oxalic acid plant into a dilute nitric acid and sulfuric acid mixture. The experiments were carried out in a pilot scale plant, consisting of a series of packed columns. Figure 5.1 shows the setup of the columns and the directly connected reactor vessels of the oxalic acid plant. The first two of the columns had a diameter of 0.25 m and were 6 m high, the third column had a diameter of 0.8 m and was 3 m high. The total NO_x content of the gas stream ranged between 20 vol% and 23 vol% at the inlet of the first column, detailed informations about the concentrations of NO and NO_2 were not given in this publication. The HNO_3 concentration of the scrubbing fluid reached 10.5 wt% in the first column, respectively 5.6 wt% in the second and 3 wt% in the third column. During the experiment NO_x removal rates of 59.5% in the first column and 76.7% for the whole setup were observed.

5 Review of absorption technologies using water as a scrubbing liquid

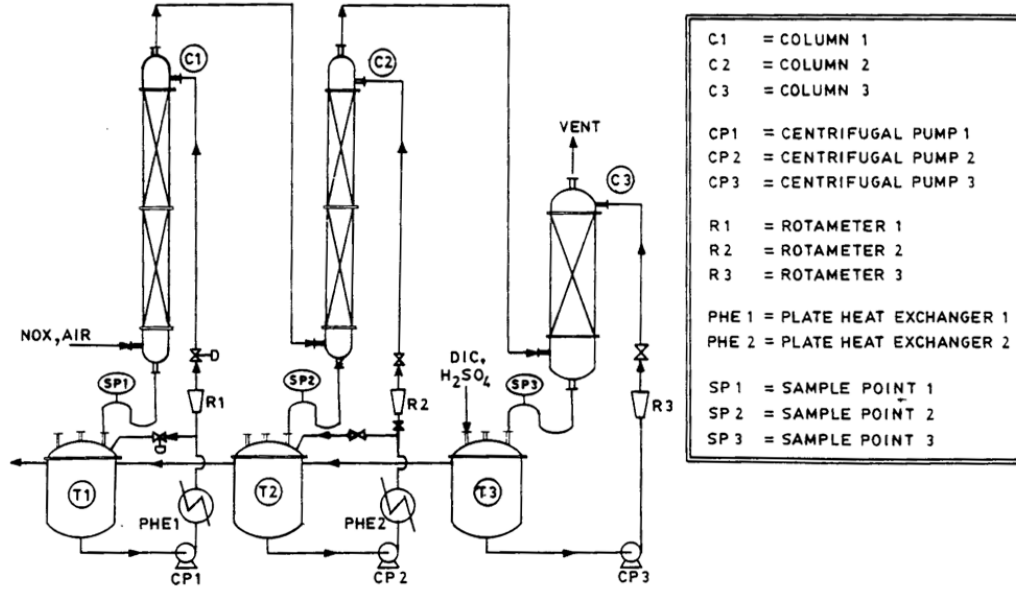


Figure 5.1: Experimental setup of scrubbing system for NO_x removal from the off gas of an oxalic acid plant consisting of three packed columns [19]

Suchak et. al. [19] conducted also an other similar experiment using two packed columns, each with a diameter of 0.8 m and 6 m high (Fig. 5.2). This time the NO_x concentration at the inlet was 14 vol%. In the first column nitric acid concentrations of 6 wt% and 2.5 wt% in the second column were reached. The removal rate was 56.3% in the first column and 77.0% totally.

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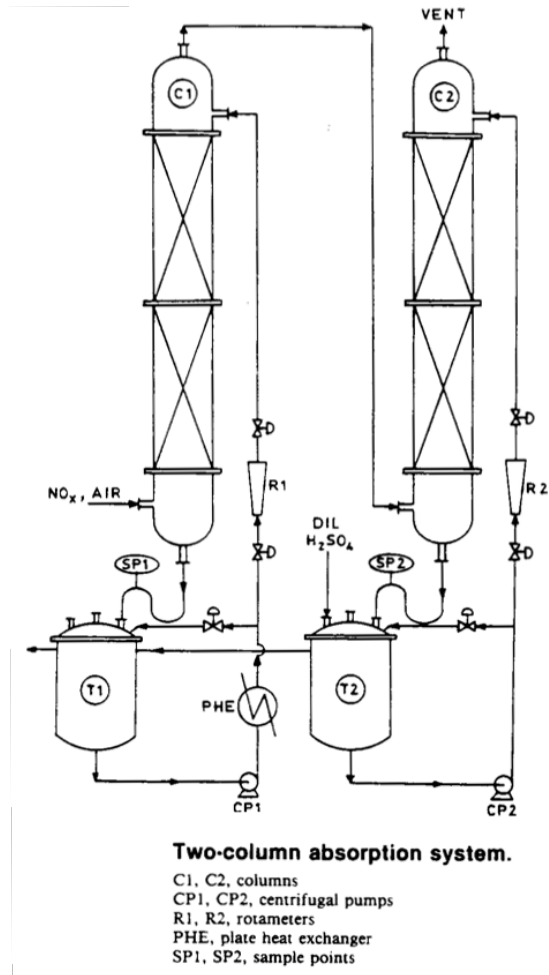


Figure 5.2: Experimental setup of scrubbing system for NO_x removal from the off gas of an oxalic acid plant consisting of two packed columns [19]

5.2.2 Experiments with biotrickling filters

Another way for NO_x removal was tested by Flanagan et. al. [20], who used a biotrickling filter instead of a conventional scrubber column (Fig. 5.3). Different organic (wood, compost) and inorganic materials (perlite, lava rock) were used as packing material, which were cultivated with thermophilic denitrifying bacteria. This system was aerated with a humidified gas stream with a NO concentration of 500 ppm_v. Depending on the residence time of the gas (up to 80s) and the used packing material removal efficiencies of up to 90% were reached.

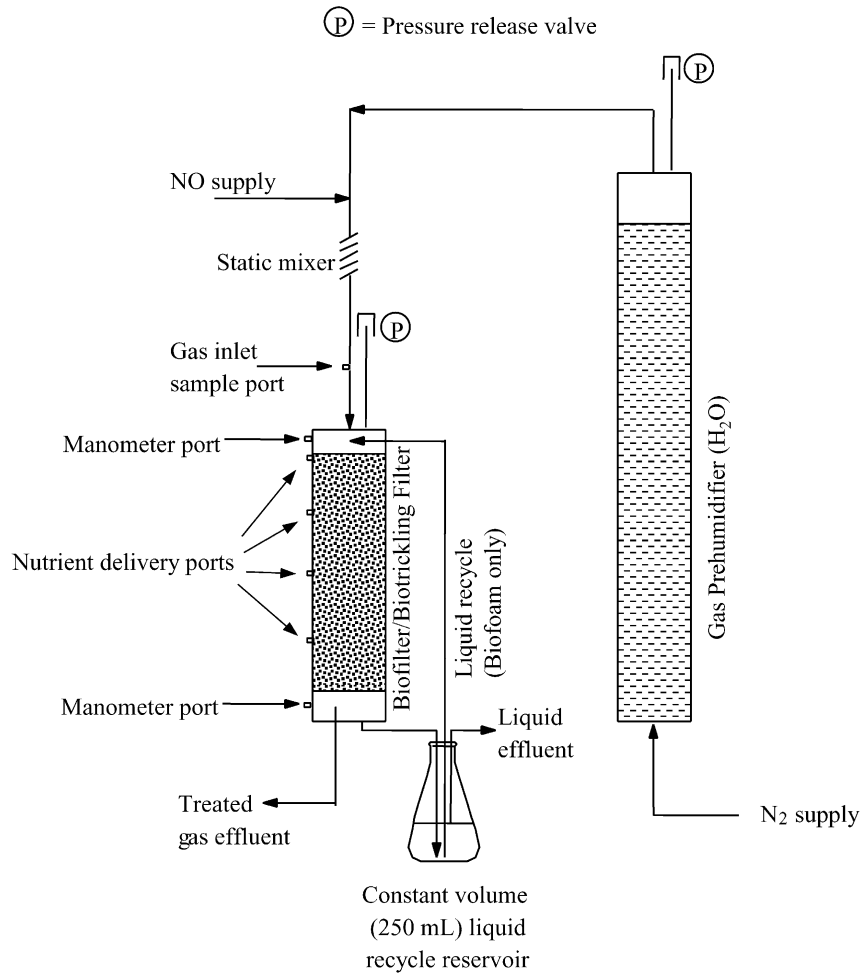


Figure 5.3: Scheme of the biofilter [20]

5 Review of absorption technologies using water as a scrubbing liquid

Huang et.al. [21] and Jiang et. al. [22, 23] also investigated the removal of nitrogen oxides in a biotrickling filter. A packed column with an internal diameter of 8 cm and 40 cm height filled with ceramic balls (4-6 mm diameter) was used. The packing was inoculated with different denitrifying and nitrifying bacteria and the column was heated to maintain a constant temperature level. Flue gas was simulated with a synthetic mixture of nitrogen, oxygen (up to 20 vol%) and nitrogen monoxide (up to 400 ppm_v) during the experiment (Fig. 5.4). With this setup NO removal rates of up to 50% with the blank column and up to 90% with the cultivated column were achieved. It was also found that a high oxygen concentration in the flue gas is a key factor for high removal efficiencies when the blank column was used, while this factor is of minor importance when the cultivated column was used.

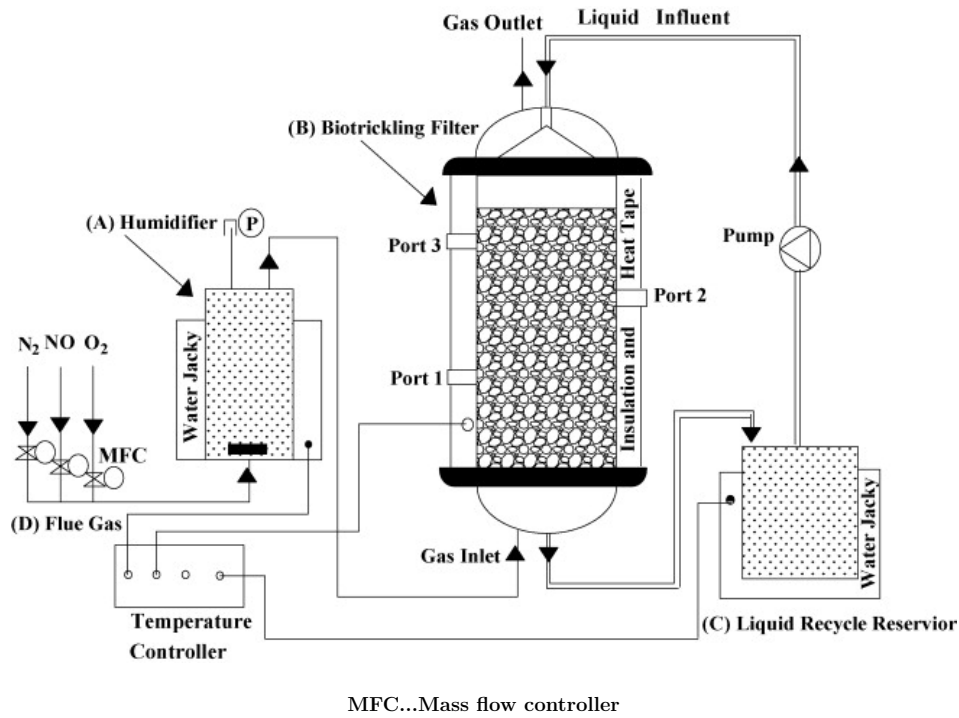


Figure 5.4: Scheme of the biotrickling filter system [23]

5 Review of absorption technologies using water as a scrubbing liquid

Y. Yang et. al. [24] used a similar design to investigate the microbial NO removal in a biotrickling filter. The tested column was 50 cm high, had an internal diameter of 8 cm and was filled with a mixture of 2-3 mm ceramic balls and 50 mm polypropylene spheres (Fig. 5.5). The column was temperature controlled by means of a heat tape. A synthetic mixture of nitrogen, oxygen (up to 20 vol%) and nitrogen monoxide (up to 600 ppm_v) was used as flue gas, a removal efficiency of 70% to 90% was reached. The oxygen content of the gas stream seems to have no influence on the removal efficiency.

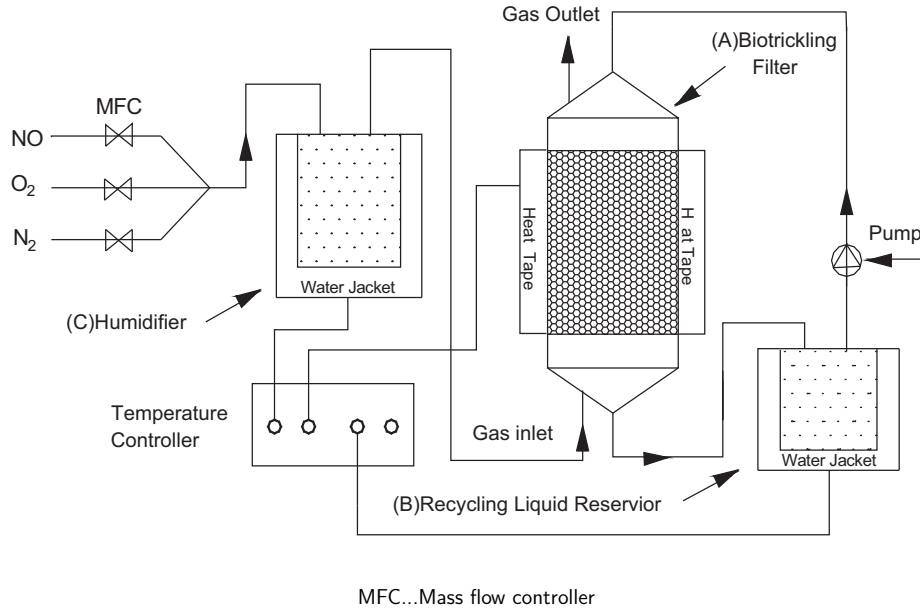


Figure 5.5: Design of the biotrickling filter [24]

W. Yang et. al. [25] used a multistage biofilter, consisting of six stages, each stage 35 cm high with a diameter of 8 cm (Fig. 5.6). The packing consisted of wood chip and compost and was cultivated with denitrifying bacteria. This system was aerated with a synthetic flue gas mixture with up to 200 ppm_v of NO. Contrary to the observations of Huang, Jiang and Y. Yang, it was found that changing the

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oxygen content of the synthetic flue gas from 0 vol% to 6 vol% lowers the removal efficiency from 99% to 55%. As reason for this behavior was stated, that the increasing oxygen content inhibits the NO utilization of the denitrifying bacteria.

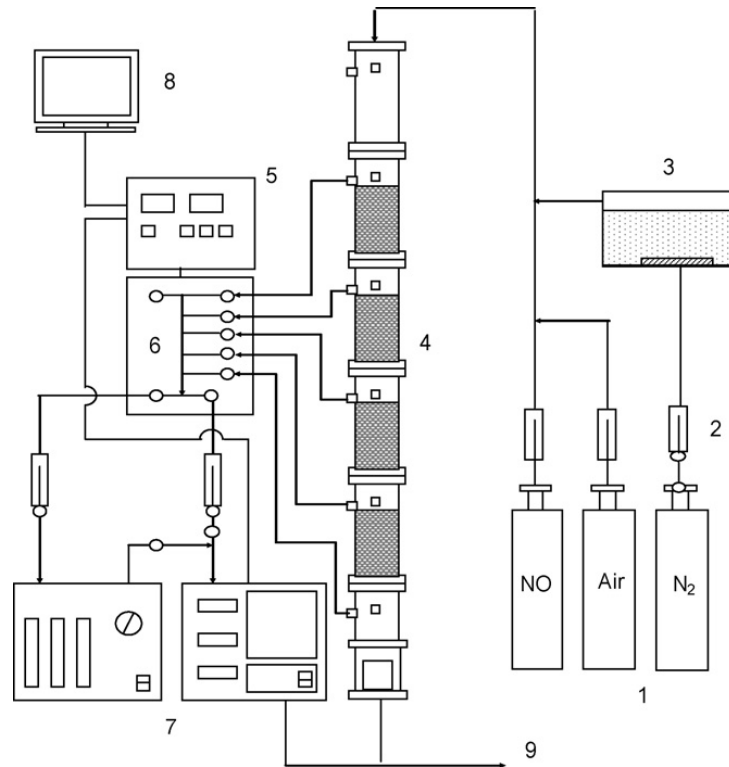


Fig. 1. Biofilter design. Component: 1, gas supply; 2, flow control; 3, humidifier; 4, biofilter column; 5, control pannel; 6, multi-switch sampling ports; 7, nitrous oxide analyzer; 8, data acquisition device; 9, ventilation.

Figure 5.6: Design of the multistage biofilter [25]

5.2.3 Experiments with a bubble column

Nagase et. al. [26, 27, 28] used a bioreactor aerated from the bottom to investigate the nitric oxide removal from flue gas. The reactor was 2.5 m high and had a diameter of 50 mm. The simulated flue gas, consisting of nitrogen (83 vol%), oxygen

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(2 vol%), carbon dioxide (15 vol%) and nitrogen monoxide (300 ppm_v), was introduced through a glass ball filter with a grain size of 40 - 50 µm. The column was cultivated with the green alga *Dunaliella tertiolecta*. During the experiments removal rates of up to 65% were reached with the cultivated system. For comparison, with deionized water a removal rate of approx. 20% was reached. Furthermore, it was observed, that an increasing oxygen content of the flue gas has a positive effect on the removal efficiency.

With an improved design of the bioreactor (Fig. 5.7), which used a mixture of air and carbon dioxide for circulation of the scrubbing liquid, the removal efficiency could be increased to 96%.

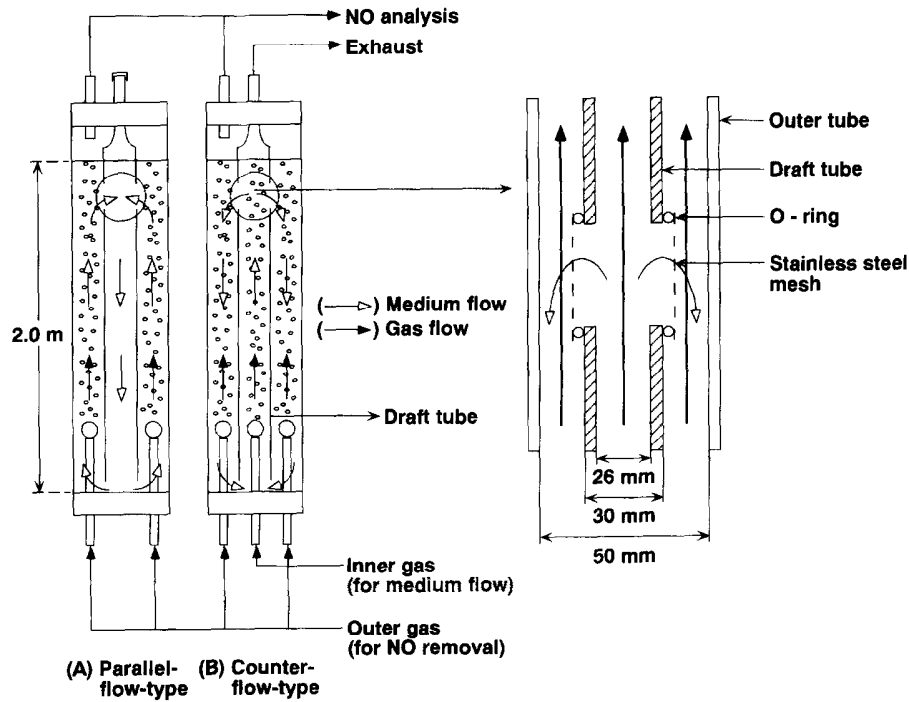


Figure 5.7: Improved design of the bubble column [27]

5.2.4 Critical comments and conclusion about the reviewed technologies

The literature review indicates that NO_x removal with an aqueous scrubbing system is possible, but not very well understood. Beside the scrubber systems used in the chemical industry for nitric acid production, which were investigated by Suchak et. al. [19], the explained systems were only lab scale models. Nevertheless, the gained removal rates in those systems are quite impressive, but the given values are difficult to compare, because no data about the used amount of scrubbing liquid or the contact surface was provided. Table 5.1 shows a summary of the removal rates cited in the previous section.

Type	Researcher	Gained removal efficiency
Packed Scrubbers	Suchak et. al. [19]	55 - 60%
Biotrickling filter	Flanagan et. al. [20]	90%
	Huang et. al. [21]	75 - 85%
	Jiang et. al. [22, 23]	80 - 90%
	Y. Yang et. al. [24]	70 - 90%
Multistage biofilter	W. Yang et. al. [25]	55 - 99%
Bubble column	Nagase et. al. [26, 27, 28]	65 - 96%

Table 5.1: Removal rates of the cultivated systems

Biotrickling filter systems and bubble columns reach the highest removal efficiencies. One disadvantage is the high pressure drop of the system and another that the bacteria or algae, used for the treatment of the dissolved pollutants, are inside the scrubbing device. This demands a constant operation of the system to avoid inactivation of the biological system. The effect of the oxygen concentration in the flue gas on the removal efficiency is not clear. Huang [21], Jiang [22, 23], Y.

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Yang et. al. [24] and Nagase et. al. [26, 27, 28] observed a positive effect of an increasing oxygen concentration, while W. Yang [25] observed a negative effect.

The bubble column system could be realized in an existing wastewater treatment plant with compressed air aeration. It would be imaginable to use the aeration system to lead flue gas directly into the bioreactor. Then the bioreactor would act like a bubble column and the problem of the reduced flexibility would be solved. On the other hand the dimensions of this "bubble column" are given and not especially designed for the best possible mass transfer. Also the legal regulations might be a problem, because the flue gas composition has to be measured, which is not possible after it is introduced into the bioreactor.

Packed scrubbers reach removal efficiencies in the same range like conventional technologies (e.g. SNCR). Compared to the biotrickling filters the scrubbing liquid is regenerated in a separate reactor attached to the scrubber system.

Instead of a special reactor for regeneration of the scrubbing liquid, also existing bioreactors could be used as already mentioned in chapter 4.3.2. This will be only possible if the microorganisms are suitable and there is free capacity to treat the pollutants. In this combination the bioreactor is always used and supplied with nutrients which is a big advantage compared to the trickling filters.

6 Description of the Wastewater Treatment and Sludge Incineration Plant Bad Vöslau

6.1 Wastewater treatment

The wastewater treatment plant is a biological treatment plant using the activated sludge process (Fig. 6.1). The main line of the plant has a design capacity of 105,000 PE (Population equivalent), while the second line, especially designed for wastewater from a dairy production, has a capacity of 50,000 PE.



1 main bioreactor, 2 settling tanks, 3 bioreactor for treatment of dairy wastewater, 4 filter press building, 5 digester, 6 greenhouses for sludge drying, 7 sludge incineration plant

Figure 6.1: Overview of the wastewater treatment plant Bad Vöslau

Main line After passing the bar screen the raw water is led to the bioreactors. The plant possess two tanks with circulating flow which can be used in serial or parallel flow. Here the biological content of the sewage is degraded and phosphates are removed. The carbon content of the raw water is used as nutrition for the bacteria which convert ammonia to nitrite, nitrate and finally nitrogen. The treated water is led to the settling tanks where the biological flocs (= sludge) are separated from clear water. The sludge is returned to the bioreactors (= recycle sludge), only a small stream (= excess sludge) is cut off to keep the total amount of the sludge in the bioreactors constant (Fig. 6.2).

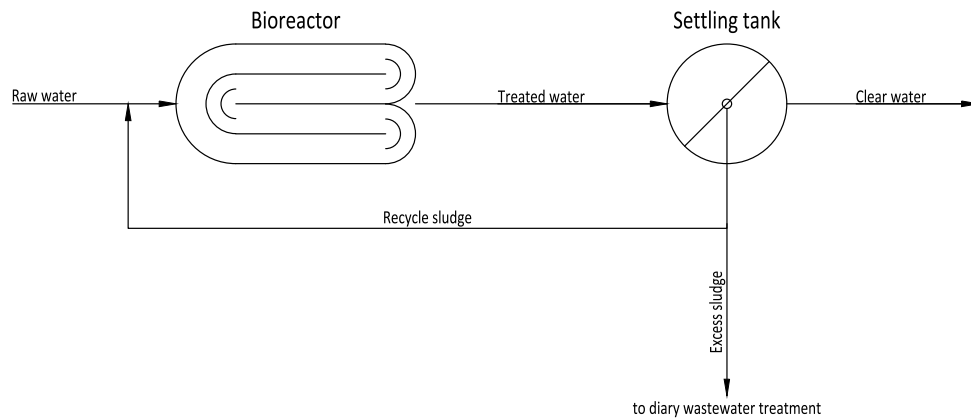


Figure 6.2: Activated sludge process

Treatment of dairy wastewater The excess sludge from the main line is used to treat the wastewater from the dairy production. In a smaller bioreactor the sludge and the wastewater are mixed and the nutrients are adsorbed to the sludge (Fig. 6.3). The residence time has to be short to avoid bulking sludge, a typical problem in plants which treat wastewater from food industry.

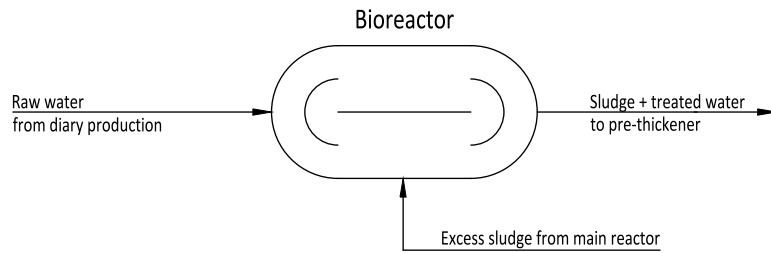


Figure 6.3: Treatment of dairy wastewater

Sludge treatment The sludge-wastewater mixture from the dairy wastewater treatment reactor is led to the pre-thickener, a belt filter. The filtrate returns to the main bioreactor, while the thick sludge is directed to the sludge treatment system.

The first step of the sludge treatment is the anaerobic digestion process. The process is run as a mesophilic digestion, where the sludge is fermented at a temperature of approx. 35 °C to produce biogas. The gas is utilized in two combined heat and power engines for production of electricity for the own demand and for heating of the digesters. After the digestion process the sludge is dewatered by means of a chamber filter press. The dewatered sludge is transported to the sludge incineration plant while the filtrate is returned to the main bioreactor. The filtrate is produced batchwise in this process which could disturb the biological process in the reactor, therefore, a retention tank was installed to homogenize the flow to the bioreactor (Fig. 6.4).

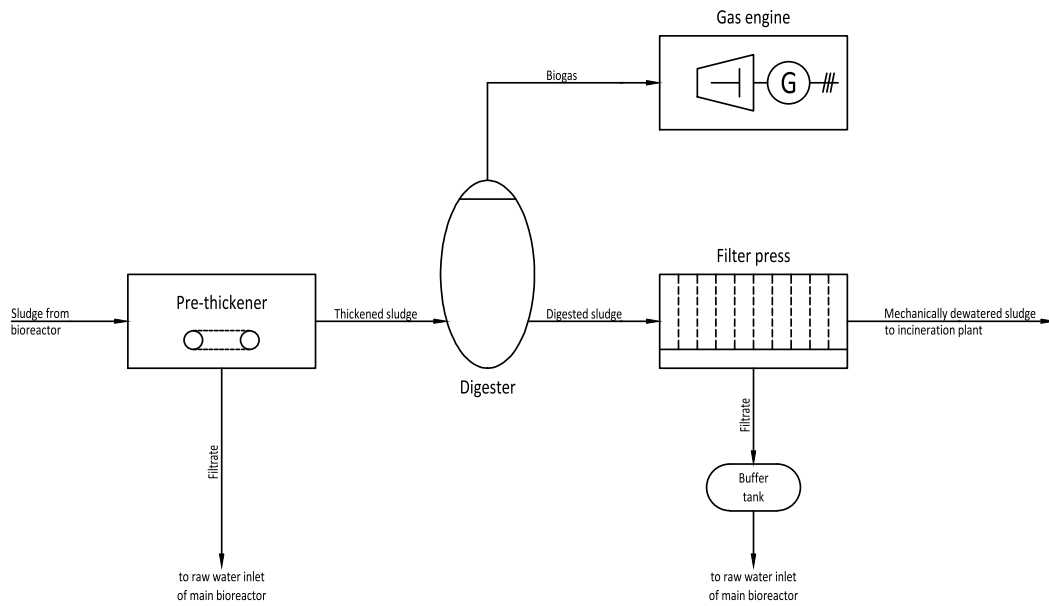


Figure 6.4: Digestion and dewatering of the sludge

6.2 Sludge incineration

The sludge incineration plant was built for combustion of the mechanically dewatered sludge from the wastewater treatment in 2004. The main parts of the plant are the solar drying system and the thermal utilization, consisting of a fluidized bed combustion reactor, a heat recovery and a flue gas cleaning system (Fig. 6.5). A more detailed flowsheet, including also the new scrubber system, is attached in Appendix C.

The solar drying system As the water content of the mechanically dewatered sludge is still above 70 wt% (dry substance below 30 wt%), it would not burn autothermally and has to be dried. The solar drying system consists of six greenhouses, each about 50 m long and 10 m wide. In these greenhouses the sludge is spread out and then dried, using solar thermal energy which heats the houses (Fig. 6.6, left). The air in the greenhouses is nearly saturated with moisture and, therefore, the air has to be changed when the moisture exceeds a certain limit.

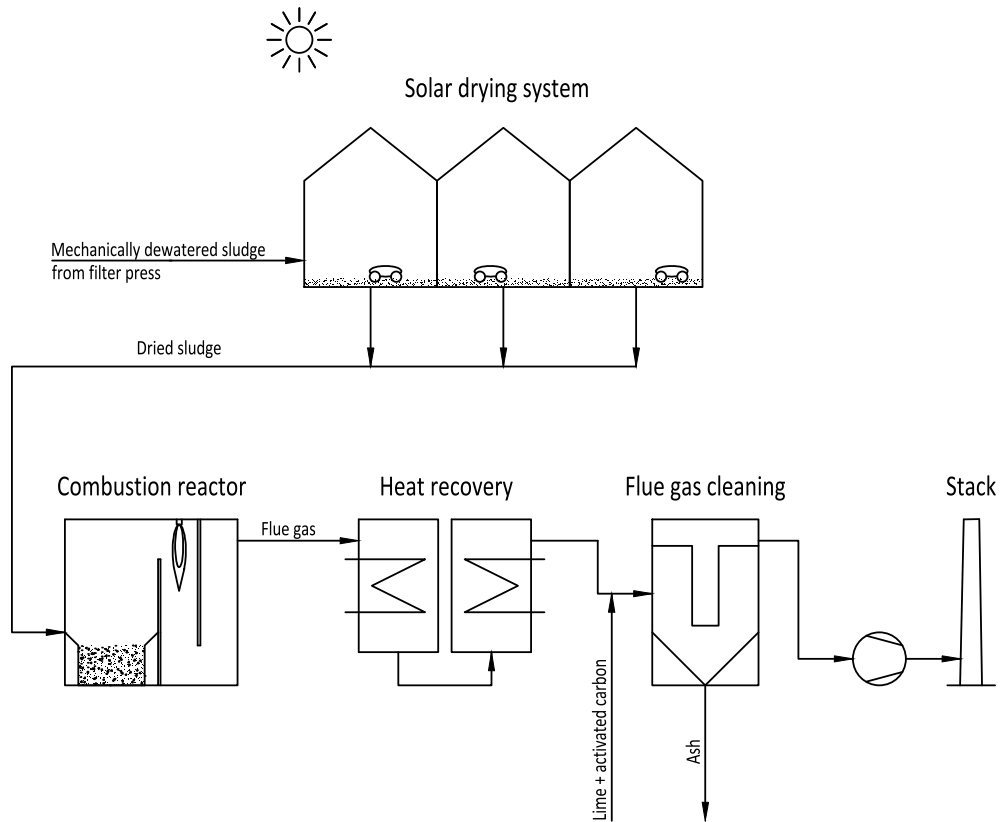


Figure 6.5: Solar drying and incineration of the sludge

The sludge on the surface of the bed dries much faster than the sludge below. It is regularly plowed from a machine called "Wendeschwein" (German for plowing pig, regarding to wild pigs, which plow the soil searching for beechnuts, chestnuts, buckeyes, acorn,...). In this way the dried sludge is quite uniform in moisture and particle size. As soon as the dry substance exceeds 75 wt% the sludge is dumped to a bunker, where it is stored till combustion (Fig. 6.6, right).



Figure 6.6: Spreading of the sludge in a greenhouse (left), "Wendeschwein" plowing the sludge (right)

The water free substance of the sludge has a lower heating value of approximately 10-12 MJ/kg, mostly depending on the ash content. Based on a minimum combustion temperature of 850 °C and adiabatic combustion conditions, a maximum water content of 50 wt% would be allowed for autothermal combustion in a fluidized bed reactor. This state is in the middle of the sticky phase of sewage sludge, where mechanical conveying is nearly impossible (Fig. 6.7) [29, 30]. Therefore, the sludge is dried to a lower water content to avoid this phase. Table 6.1 shows the main technical data of the solar drying system.

Solar drying system		
Number of drying lines	6	-
Total drying surface	3000	m ²
Average drying time	9-14	days
water content of the wet sludge	70-80	wt%
water content of the dried sludge	20-25	wt%

Table 6.1: Technical data of the solar drying system

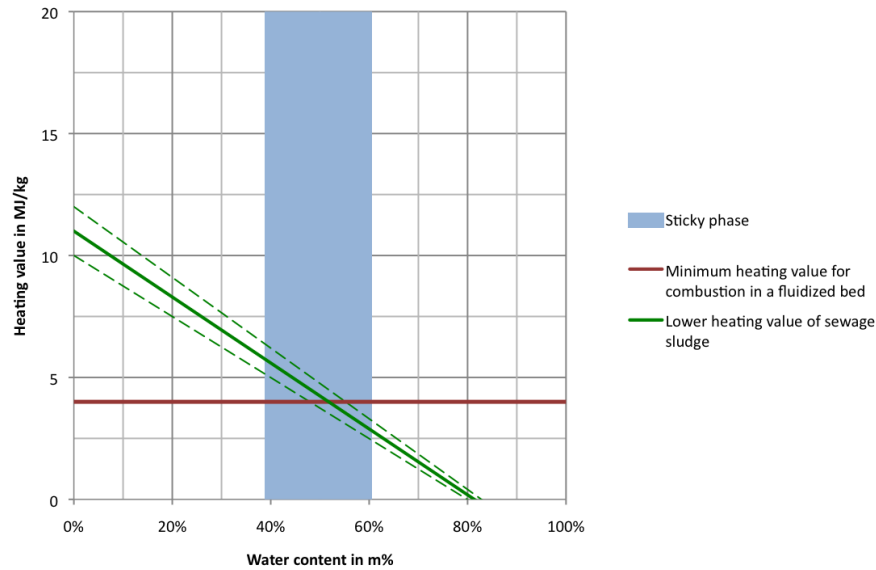


Figure 6.7: Dependency of the lower heating value of sewage sludge on the water content (based on a lower heating value of the dry sludge of 10-12 MJ/kg) [29]

The combustion reactor The heart of the plant is a fluidized bed incinerator, where the sludge is burned in a staged combustion process. After the sludge is gasified in a fluidized bed reactor the produced gas is combusted in the freeboard and the afterburner chamber (Fig. 6.9). The incinerator is designed for a feed of 800 kg/h, equivalent to a fuel power of approximately 1.2 MW. The main technical data of the combustion reactor are shown in Table 6.2, more detailed data about the dried sewage sludge are shown in Table 8.4.

The gasification reactor is built as a bubbling fluidized bed reactor with a rectangular cross-section, silica sand with a particle size between 700 and 1000 μm is used as bed material. The reactor is refractory lined and designed adiabatically without a heat exchanger in the bed or the freeboard area. The sludge is transported from the hopper by means of screw feeders and dumped into the fluidized bed with a spreader stocker. Screenings from the sewage plants inlet bar screen

Fluidized bed combustion reactor		
Cross section bed	1200 x 1500	mm
Cross section freeboard	1700 x 2800	mm
Silica sand (bed material)	1000	kg
Bed temperature	750	°C
Freeboard temperature	880	°C
Fuel feed rate	600 - 800	kg/h
Fuel power	1200	kW
Utilizable power (heat recovery system)	800	kW

Table 6.2: Technical data of the combustion reactor

can be inserted as additional fuel into the reactor from a separate hopper via a screw feeding system (Fig. 6.8).

The primary air is used as gasification and fluidization agent. The air-fuel equivalence ratio is around 0.6 to 0.7, this allows an autothermal process for this low caloric fuel at a gasification temperature of approximately 750 °C. In this way the fuels are decomposed in the reactor to CO, CO₂, H₂, H₂O, light hydrocarbons, tars and ashes. As air is used as gasification agent, the produced gas contains also a considerable fraction of N₂ and, therefore, the gas has a quite low caloric value.

The fluidization air is inserted to the reactor via air tubes with downwards targeted nozzles. The space between the nozzle tubes is filled with ceramic balls to protect the tubes and the nozzles from erosion through the hot sand. Compared to plants for wood combustion no flue gas recirculation is necessary to ensure proper fluidization or for regulation of the temperature in the gasification zone. The low calorific value of the sludge allows to use air as fluidization agent without extra cooling of the bed or flue gas recirculation in the design case. For sludges with a slightly higher caloric value a water spray system is installed to control the bed temperature.

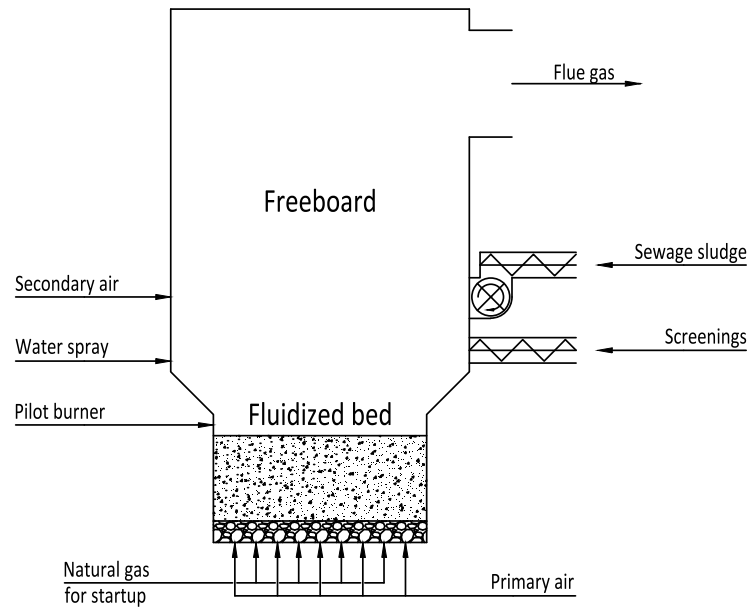


Figure 6.8: Scheme of the fluidized bed reactor

Natural gas is inserted through separate gas nozzles for the start-up procedure. This gas/air mixture coming out of the fluidized bed is ignited by two pilot burners positioned slightly above the upper bed level of the expanded bed. As soon as the selected bed temperature is reached, the sludge feed is started and the gas flow can be continuously reduced and finally stopped.

Small ash particles are blown out of the reactor with the hot gas stream. Bigger particles remain in the bed. Some of the ash particles have a quite stable structure and, therefore, the erosion of these particles is slow. An increasing concentration of ash in the bed and an immoderate growth of the bed have to be avoided and, therefore, these particles should be removed. The ash particles have a lower density compared to the bed material and can be found in higher concentrations in the upper regions of the bed. This behaviour is utilized to remove the ash particles from the bed. An overflow pipe is installed where the ash/bed material mixture

from the top of the bed flows in, then it is transported out of the reactor with a screw conveyer to a roller crusher. There the particles are milled to a size smaller than 1 mm, the milled material is returned to the bed via pneumatic transport.

The gaseous products from the gasification process are burned in the freeboard, where secondary air is added. Here the temperature reaches approximately 870 to 880°C. In order to increase the residence time for complete combustion the gas is further led to an afterburner chamber. There a gas burner is installed to ensure an exhaust temperature above 850 °C during two seconds, which is required by law for waste incineration processes. The power of this gas burner is controlled by the exhaust temperature and it stops when no extra power is required to reach the temperature setpoint. Like the reactor also the afterburner chamber is refractory lined for minimum heat loss.



Figure 6.9: The fluidized bed reactor with the attached afterburner chamber

The heat recovery system From the afterburner chamber the hot flue gas is led to the heat recovery boiler. Here the gas is cooled to approximately 200 °C in two thermal oil heat exchangers (Fig. 6.10). Both are coil type heat exchangers with a guide pipe in the middle surrounded by the heat exchanger bundle. The hot gas flows downwards in the center pipe and streams back up along the bundle on the outer side, while the thermal oil flows in counter current direction from the top downwards.



Figure 6.10: Thermal oil heat exchanger

The first bundle which is operated at higher temperature is made from carbon steel. The second which is operated at lower temperatures and therefore facing a higher risk of condensation is made from stainless steel. After the second heat

exchanger the cooled flue gas is further led to the flue gas cleaning system. When a shutdown in the thermal oil system happens, the hot flue gas is bypassed to protect the system from over temperature. In such a case the gas is cooled by maximum water injection in the quench of the flue gas cleaning system.

The thermal oil system is coupled to the plants hot water system via a plate type heat exchanger. As there is no air preheating installed, the complete thermal power is distributed to the hot water heating system of the wastewater treatment plant. There it is used for heating of the digesters, the buildings and to support the solar drying system. The surplus is directed to the local district heating system, especially in summer when no support of the solar drying is necessary.

The flue gas cleaning system After the heat recovery system pollutants like HCl, SO₂, heavy metals and dust are removed from the flue gas in a conditioned dry sorption process. Table 6.3 shows a typical composition of the flue gas.

Component		Concentration	
Carbon dioxide	CO ₂	8 - 12	vol%
Nitrogen	N ₂	65 - 70	vol%
Oxygen	O ₂	6 - 10	vol%
Water	H ₂ O	12 - 18	vol%
Hydrochloric acid	HCl	0 - 100	mg/Nm ³
Carbon monoxide	CO	0 - 10	mg/Nm ³
Nitrogen monoxide	NO	50 - 200	mg/Nm ³
Nitrogen dioxide	NO ₂	0	mg/Nm ³
Sulfur dioxide	SO ₂	100 - 500	mg/Nm ³

Table 6.3: Typical composition of the wet flue gas entering the gas cleaning system

First the flue gas is quenched by means of water injection to a temperature of 160 °C. Afterwards hydrated lime and activated carbon are added to the flue gas stream in a plug flow reactor, which ends in the particle filter. The hydrated lime reacts with acidic pollutants like SO₂, HCl or HF while the activated carbon adsorbs volatile heavy metals like mercury or cadmium and also PCDD/Fs. Finally, the adsorption products and the ashes are separated from the gas stream in a multiple tube filter (Fig. 6.11).



Figure 6.11: Multiple tube filter for separation of flue gas and particles

The filter contains porous ceramic tubes which are streamed from the outside to the center. On the outside, the particles build a filter cake. In the filter cake, there is still unreacted lime and carbon which can adsorb pollutants. Therefore, the filter cake remains on the candles as long as possible. Once the pressure drop exceeds 20 mbar, the cake is removed by periodical air pulses from the inner side. The ashes and the adsorption products are collected at the cone shaped bottom of the filter and transported via screw feeders to storage containers. Table 6.4 lists the main technical data of the flue gas cleaning system.

Flue gas cleaning system	
Flue gas conditioning	water spray
Adsorption agents	hydrated lime
	activated carbon
Filter material	porous ceramic
Number of filter candles	540
Total filter surface	150 m ²

Table 6.4: Technical data of the flue gas cleaning system

After filtration the gas passes the induced draft fan and is exhausted through the chimney to the environment. From the chimney the gas samples for the clean gas analysis are taken, also the gas velocity and the dust content of the flue gas are measured there (for details about the gas measuring system see 8.1.1).

7 Development of the scrubber system

7.1 Initial situation

Experiments with a small testing facility carried out by Hermann Schild et al. in 2009 had already shown that the emissions from sludge incineration could be significantly reduced by means of a biological flue gas cleaning system [31]. In these experiments the biological scrubbing technology showed not only a reduction of acidic gases as HCl and SO₂, but also an unexpected positive effect on the NO concentration.

These promising findings led to the start of a new research project to obtain knowledge about the possibilities of such a biological gas cleaning system, which was funded by "Kalogeo Anlagenbau" and the "Österreichische Forschungsförderungsgesellschaft". Within the scope of the current research was not only a development of a suitable flue gas cleaning system for sewage sludge incineration, but also a further investigation of the NO_x reduction capabilities of such a system.

The new scrubber was built at the sludge incineration plant Bad Vöslau which is attached to a wastewater treatment plant, both owned and operated by the "Gemeindeverband Abwasserbeseitigung Raum Bad Vöslau". It is an ideal place for testing such a biological flue gas cleaning process because the bioreactors of the plant can be used for regeneration of the scrubbing fluid. An overview to the incineration plant and the biological water treatment system is given in chapter 6, the design of the new scrubber system is explained in the next sections.

7.2 Design of the scrubber

The scrubber system is built as a spray scrubber with an upstream quench cooler and a downstream flue gas reheater. The spray scrubber and the quench cooler are mounted on top of the scrubber basin, which acts as a reservoir for the used scrubbing solution. The scrubber system was installed subsequently between the existing multiple tube filter and the induced draft fan. The existing flue gas channel was dismantled and rebuilt as a bypass, in case of the scrubber system has to be shut down. The scrubbing solution is taken from the sewage plants process water system and after usage it is sent back to the plant for regeneration. Figure 7.1 shows the main components of the system, a detailed flowsheet is attached in Appendix D. The installation of the scrubber system at the site of the sludge incineration plant is shown in Figure 7.2. The previously explained multiple tube filter is visible in the background.

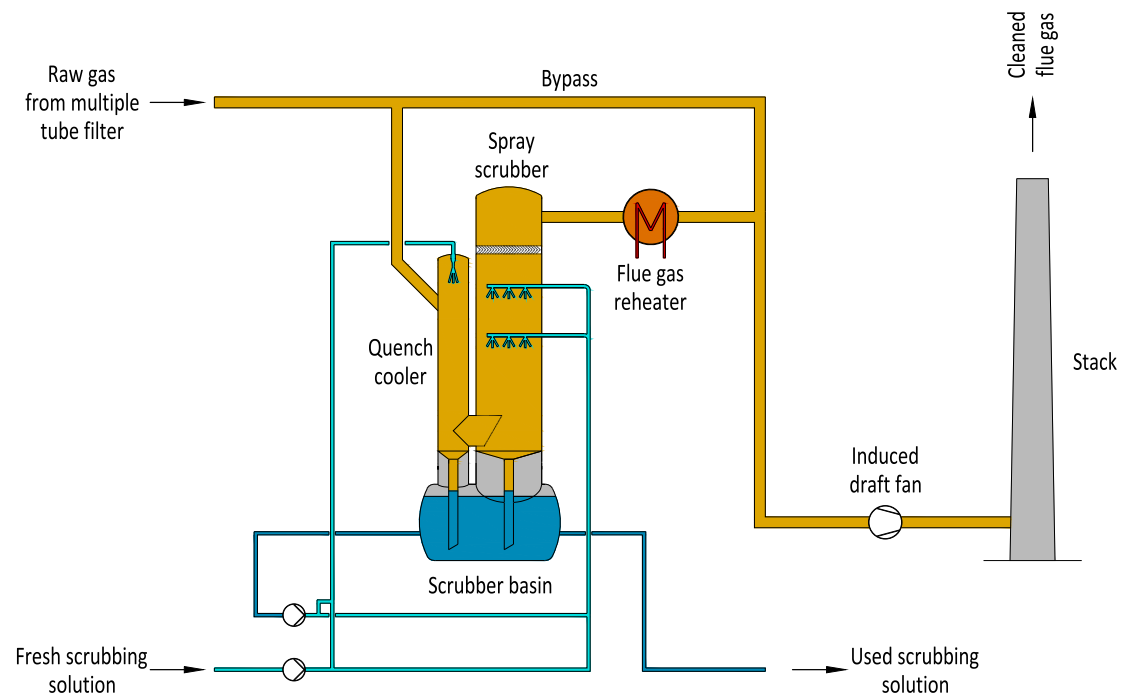


Figure 7.1: Scheme of the scrubber system

7 Development of the scrubber system



Figure 7.2: Scrubber system installation at the sludge incineration plant

7.2.1 Quench cooler

The quench cooler is built as a separate column, in which the cooling fluid is in direct contact with the flue gas. Therefore, the quench cooler is not only used to cool the flue gas but also to saturate the gas stream with water vapor. Figure 7.3 shows the design of the quench column.

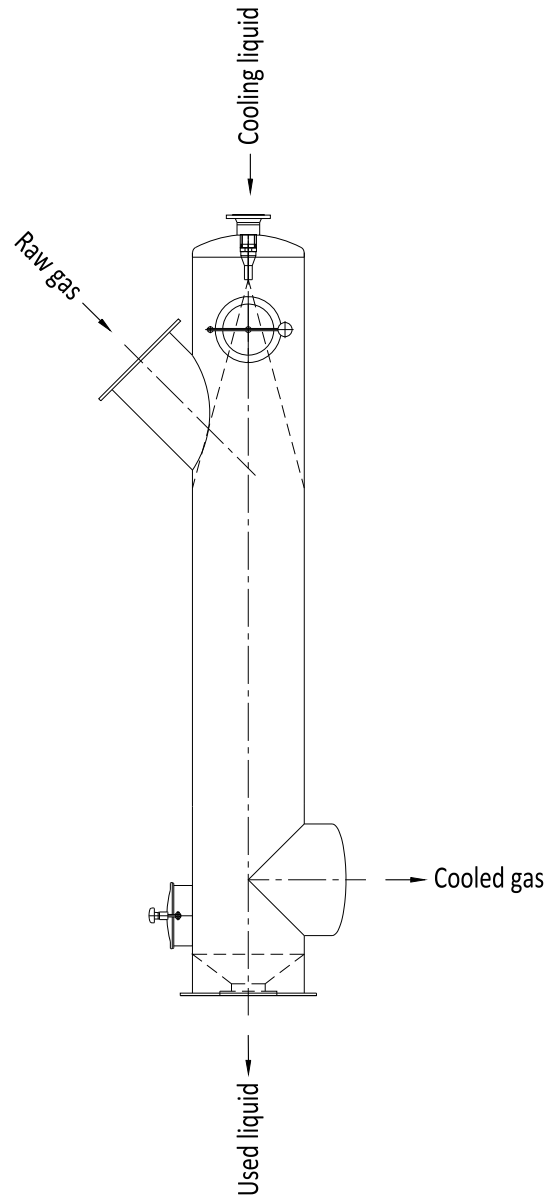


Figure 7.3: Design of the quench column

7 Development of the scrubber system

Construction The column has a diameter of 560 mm and a total height of 3915 mm and it is equipped with two inspection lids, one at the top near the spray nozzle and the other one at the base. The shell is made of 1.4571 stainless steel. The flue gas enters the column near the top at an angle of 45° downwards and leaves the column at the base towards the scrubber column. The cooling liquid is injected at the top of the column and flows in cocurrent stream with the flue gas to the bottom of the column. A full cone nozzle with a spray angle of 30° is used for vaporization of the cooling liquid. This maintains a good cooling effect as the full gas stream is in contact with the fluid. The nozzle has a nominal flow of 23 m³/h at a pressure of 1 bar. At the bottom of the column the used cooling liquid is led to the common scrubber basin through a dip pipe, which acts as a siphon to cover the pressure difference between the column and the environment.

The gas enters the column with a temperature of approx. 140°C. According to the nominal gas flow of 8500 Nm³/h (wet), the gas velocity is about 12 m/s. The contact time between gas and cooling liquid is around 0.2 s. Pure water from the process water network or used scrubbing solution can be used as cooling liquid.

Mass and energy balance Depending on the chosen outlet temperature of the gas stream, the saturation concentration of water vapor is above or below the inlet concentration. This means that cooling liquid evaporates and water vapor condenses in the quench column, both need to be considered in the mass and energy balance.

The dry and the wet gas flow before and after the quench column at normal conditions are needed for calculation of the water balance (Eq. 7.1 to 7.3).

$$\dot{V}_{g,dry,in,n} = \dot{V}_{g,wet,in,n} \cdot (1 - c_{water}) \quad (7.1)$$

$$\dot{V}_{g,dry,out,n} = \dot{V}_{g,dry,in,n} \quad (7.2)$$

$$\dot{V}_{g,wet,out,n} = \frac{\dot{V}_{g,dry,n}}{1 - c_{water,sat}} \quad (7.3)$$

7 Development of the scrubber system

The saturation concentration of water vapor is calculated with the saturation partial pressure of water vapor and the absolute pressure of the flue gas according to equation (7.4). Values for the saturation partial pressure of water can be found in Table 7.1.

$$c_{water,sat} = \frac{p_{sat}}{p_{abs}} \quad (7.4)$$

Temperature [°C]	Saturation partial pressure p_{sat} [bar]
35	0.0563
50	0.1235
65	0.2504
140	3.6150

Table 7.1: Saturation partial pressure of water [32]

With equation (7.5) the volume flows are converted to mass flows using the normal density of the flue gas respectively of water vapor. The density of the dry flue gas is calculated according to the flue gas composition with equation (7.6). The normal densities of nitrogen, oxygen, carbon dioxide and water vapor are listed in Table 7.2.

$$\dot{m} = \dot{V}_n \cdot \rho_n \quad (7.5)$$

$$\rho_{g,dry,n} = \rho_{N_2,n} \cdot c_{N_2,dry} + \rho_{O_2,n} \cdot c_{O_2,dry} + \rho_{CO_2,n} \cdot c_{CO_2,dry} \quad (7.6)$$

Component	Normal density ρ_n [kg/m³]
Nitrogen	1.234
Oxygen	1.410
Carbon dioxide	1.951

Table 7.2: Normal densities of the main flue gas components [32]

7 Development of the scrubber system

The condensed (or evaporated) amount of water is calculated with equation (7.7). If the result is positive, water vapor will condense, otherwise water will evaporate.

$$\dot{m}_{v,con} = \dot{m}_{g,wet,in} - \dot{m}_{g,wet,out} \quad (7.7)$$

The energy balance for the quench cooler is set up in equation (7.8), losses due to heat transfer through the shell of the column are neglected.

$$0 = \dot{m}_{g,dry,in} \cdot h_{g,in} + \dot{m}_{v,in} \cdot h_{v,in} + \dot{m}_{liq,in} \cdot h_{l,in} - \dot{m}_{g,dry,out} \cdot h_{g,out} - \dot{m}_{v,out} \cdot h_{v,out} - \dot{m}_{l,out} \cdot h_{l,out} \quad (7.8)$$

The enthalpy of the dry flue gas is calculated with equation (7.9). Enthalpies of nitrogen, oxygen, carbon dioxide and water vapor are listed in Table 7.3.

$$h_{g,dry} = h_{N_2} \cdot c_{N_2,dry} \cdot \frac{\rho_{N_2,n}}{\rho_{g,dry,n}} + h_{O_2} \cdot c_{O_2,dry} \cdot \frac{\rho_{O_2,n}}{\rho_{g,dry,n}} + h_{CO_2} \cdot c_{CO_2,dry} \cdot \frac{\rho_{CO_2,n}}{\rho_{g,dry,n}} \quad (7.9)$$

Temperature [°C]	Enthalpy h [kJ/kg]				
	Nitrogen	Oxygen	Carbon dioxide	Water (vapor)	Water (liquid)
35	319.7	8.96	514.3	146.6	2564.6
50	335.3	22.79	527.4	209.3	2591.3
65	350.9	36.67	540.7	272.1	2617.5
140	429.2	106.80	609.7	589.2	2733.4

Table 7.3: Enthalpies of the main flue gas components [32]

The mass flow of cooling liquid out of the column is calculated according to equation (7.8).

$$\dot{m}_{l,out} = \dot{m}_{l,in} + \dot{m}_{v,con} \quad (7.10)$$

Inserting equation (7.10) into the energy balance (7.8) allows to calculate the required flow of cooling liquid for a given outlet temperature.

7 Development of the scrubber system

Results and operating range The mass and energy balance was calculated for a nominal flue gas flow of 8500 Nm³/h (wet) and a typical water content of 16 vol%, the dry flue gas composition was estimated as shown in Table 7.4. More details about the estimated flue gas composition can be found in Appendix A.

Component		Concentration
Carbon dioxide	CO ₂	9.7 vol%
Nitrogen	N ₂	80.7 vol%
Oxygen	O ₂	9.5 vol%

Table 7.4: Composition of the dry flue gas as used for design calculations

Figure 7.4 shows the calculated outlet temperature of the quench column for different cooling liquid flows. If recirculated scrubbing liquid (assumed temperature 40°C, nominal flow 23 m³/h) is used for the quench column, a minimum temperature of the flue gas of 48°C and a water content of 11 vol% will be reached. Compared to this scenario, if fresh water is used (temperature 15°C, nominal flow 23 m³/h), a minimum temperature of the flue gas of 34°C and a water content of 5 vol% can be reached.

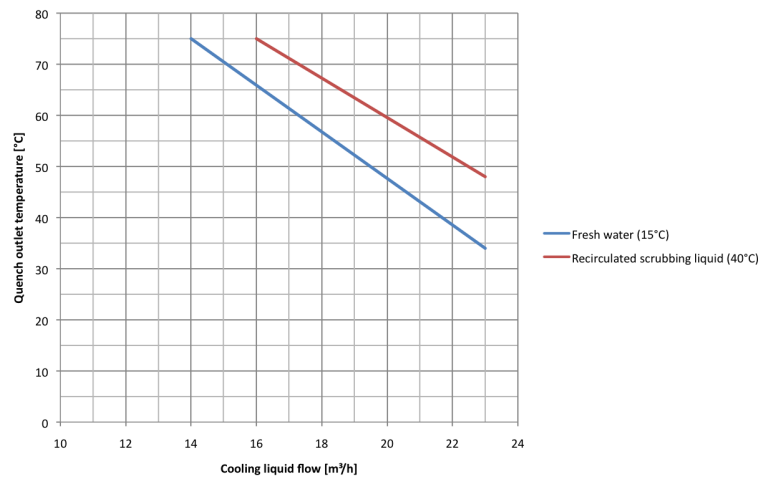


Figure 7.4: Quench column outlet temperature vs. cooling liquid flow

7.2.2 Scrubber column

The scrubber column is built as a spray scrubber with two spray levels. Here the flue gas is in intense contact with the scrubbing liquid and the pollutants are absorbed into the liquid phase. The basic design of the column is shown in Figure 7.5.

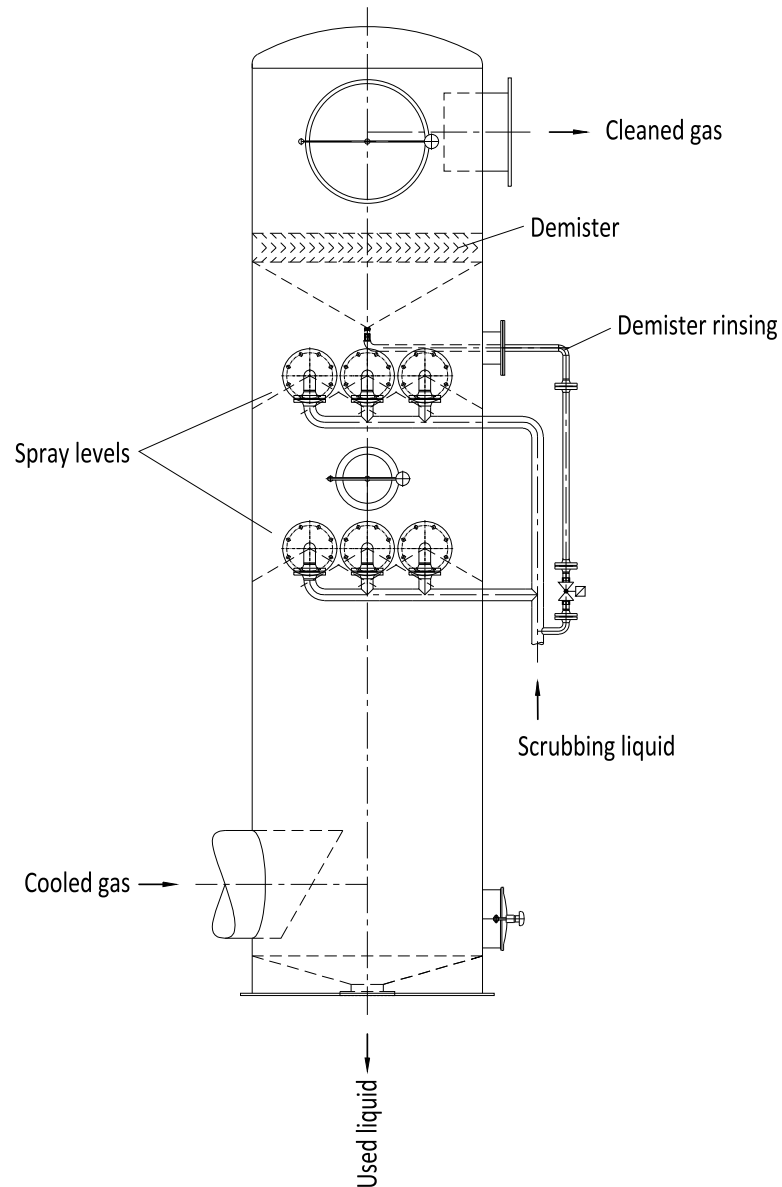


Figure 7.5: Design of the scrubber column

7 Development of the scrubber system

Construction The scrubber column has a diameter of 1200 mm and a total height of 5050 mm and is made of 1.4571 stainless steel. Like the quench column it is equipped with inspection lids, one at the bottom, one between the spray levels and one at the top of the column. The cooled flue gas from the quench column enters the scrubber at the bottom and leaves it at the top of the column. The scrubbing liquid is inserted to the scrubber at two spray levels, each consisting of seven spray nozzles. The lower level is installed at a height of 1750 mm above the gas entrance, while the upper level is situated at 2650 mm. The spray nozzles are full cone nozzles with a spray angle of 120° and a nominal flow of $1.4 \text{ m}^3/\text{h}$ each, at a pressure of 1 bar. This makes a total flow of $19.6 \text{ m}^3/\text{h}$, which can be increased up to $31.9 \text{ m}^3/\text{h}$ at 3 bar. According to the supplier, the Sauter mean diameter of the droplets is between $350 \text{ }\mu\text{m}$ and $900 \text{ }\mu\text{m}$, depending on the flow rate (Fig. 7.6). The used scrubbing solution leaves the column at the bottom and is led to the scrubber basin through a dip pipe like in the quench column.

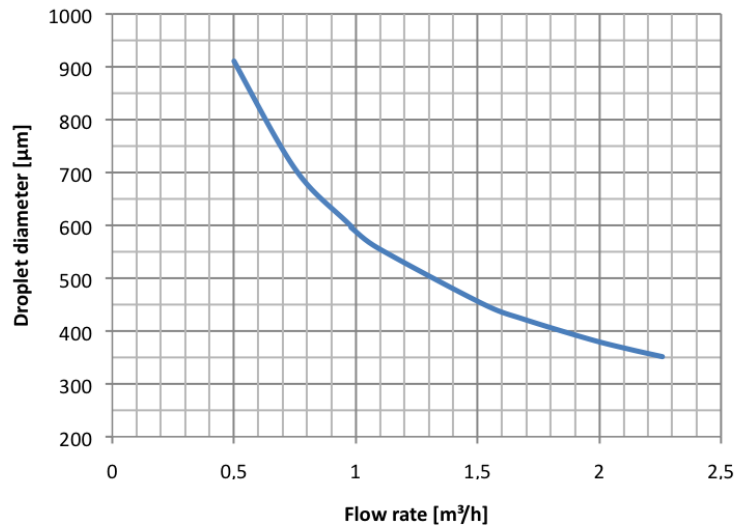


Figure 7.6: Sauter mean diameter of the droplets [33]

7 Development of the scrubber system

The gas velocity in the scrubber column is approximately 2.5 m/s, based on the nominal flow of 8500 Nm³/h (wet). This allows a contact time between flue gas and scrubbing liquid of about one second.

Above the spray levels a demister is installed to minimize the amount of entrained water drops. It is a vane type demister with vertical gas flow and made of polypropylene (Fig. 7.7). The removed scrubbing liquid drains back down to the scrubbing area. If necessary the demister can be rinsed with scrubbing liquid by a separate rinsing nozzle.



Figure 7.7: Vane type demister [34]

Mass and energy balance During the design phase of the scrubber, only the physical solubilities of the gases were considered because there was not enough data available for a more detailed calculation (especially for NO). Based on the Henry coefficients of the substances (Tab. 7.5), the equilibrium concentration of these substances is calculated according to Henry's law (eq. (7.11)).

$$c_{x,l} = k_{H,x} \cdot p_x \quad (7.11)$$

The amount of the transferred component is calculated with equation (7.12).

$$\dot{n}_{x,t} = (c_{x,out} - c_{x,in}) \cdot \dot{m}_{l,in} + c_{x,out} \cdot \dot{m}_{v,con} \quad (7.12)$$

7 Development of the scrubber system

Equation (7.13) shows the calculation of the removal efficiency for the individual components.

$$\eta_{rem,x} = \frac{\dot{m}_{x,t}}{c_{x,dry,raw} \cdot 10^{-6} \frac{kg}{mg} \cdot \dot{V}_{g,dry,n}} \quad (7.13)$$

Component		Solubility $k_H \left[\frac{mol}{kg \cdot bar} \right]$
Carbon monoxide	CO	9.5×10^{-4}
Carbon dioxide	CO ₂	3.4×10^{-2}
Hydrochloric acid	HCl	11
Nitrogen	N ₂	6.1×10^{-4}
Nitrogen monoxide	NO	1.9×10^{-3}
Nitrogen dioxide	NO ₂	1.2×10^{-2}
Oxygen	O ₂	1.3×10^{-3}
Sulfur dioxide	SO ₂	1.2

Table 7.5: Henry coefficients of typical substances in flue gases [35, 36]

The mass and energy balance for the scrubber column is calculated in a similar way like for the quench column. Equations (7.1) to (7.5) and (7.10) are used again. The heats of solution need to be included into equation (7.8) as illustrated in equation (7.14). The used heats of solution are shown in Table 7.6

$$\begin{aligned} 0 = & \dot{m}_{g,dry,in} \cdot h_{g,in} + \dot{m}_{v,in} \cdot h_{v,in} + \dot{m}_{l,in} \cdot h_{l,in} - \\ & - \dot{m}_{g,dry,out} \cdot h_{g,out} - \dot{m}_{v,out} \cdot h_{v,out} - \dot{m}_{l,out} \cdot h_{l,out} + \\ & + \sum_x \dot{m}_{x,t} \cdot h_{sol,x} \end{aligned} \quad (7.14)$$

In the scrubber column the flow of scrubbing liquid is specified, therefore the exit temperature of the flue gas is not known and the balance has to be calculated by iteration.

7 Development of the scrubber system

Component		Heat of solution h_{sol} [kJ/kg]
Carbon monoxide	CO	475.1
Carbon dioxide	CO ₂	453.5
Hydrochloric acid	HCl	136.7
Nitrogen	N ₂	386.0
Nitrogen monoxide	NO	471.2
Nitrogen dioxide	NO ₂	451.9
Oxygen	O ₂	441.7
Sulfur dioxide	SO ₂	402.1

Table 7.6: Heats of solution of typical substances in flue gases [36]

Operating range and expectations for the scrubber system Based on the given values for the physical solubility, a nominal scrubbing liquid flow of 20 m³/h and a flue gas stream of 8500 Nm³/h (wet) removal efficiencies were calculated. A raw gas composition was estimated for this calculation, based on the sludge composition and the operational experience from the incineration plant Bad Vöslau (Tab. 7.7 and Appendix A).

Component		Concentration	
Carbon dioxide	CO ₂	8.1	vol%
Nitrogen	N ₂	67.8	vol%
Oxygen	O ₂	8.0	vol%
Water	H ₂ O	16.0	vol%
Hydrochloric acid	HCl	100	mg/Nm ³
Carbon monoxide	CO	10	mg/Nm ³
Nitrogen monoxide	NO	200	mg/Nm ³
Nitrogen dioxide	NO ₂	0	mg/Nm ³
Sulfur dioxide	SO ₂	500	mg/Nm ³

Table 7.7: Composition of the wet flue gas as used for design calculations

7 Development of the scrubber system

It was found that almost 100% of the hydrochloric acid but only about 10% of the sulfur dioxide can be removed by physical solution into the scrubbing liquid. The dissolved SO_2 reacts immediately to sulfite and sulfate and, therefore, new SO_2 can be dissolved, which leads to total absorption rates of more than 90%.

The calculated removal efficiency for NO is another three orders of magnitude lower than from SO_2 , but as stated in chapter 5 much higher removal rates were observed by numerous researchers, which is also related to quite fast reactions in the scrubbing liquid.

Figure 7.8 shows these calculated values of the removal efficiency for different scrubbing liquid flows.

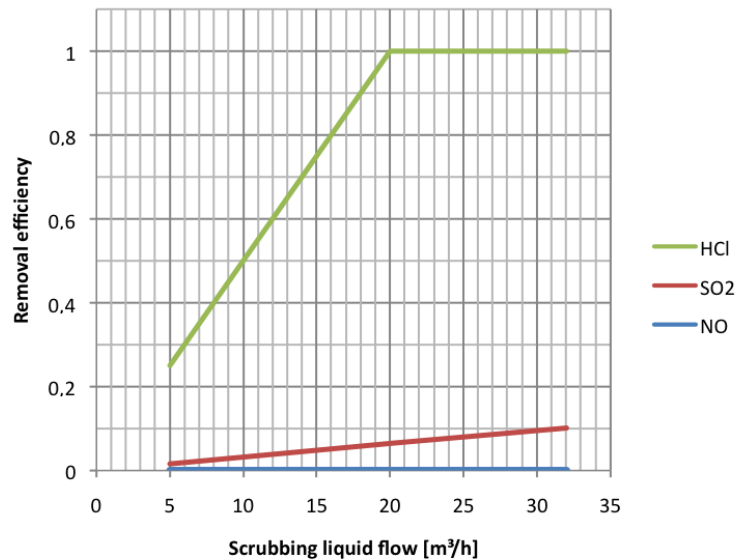


Figure 7.8: Calculated removal efficiencies depending on the scrubbing liquid flow

Based on a constant fresh scrubbing liquid flow, the removal efficiencies for SO_2 and NO are further increased, when scrubbing liquid recirculation is used. During the residence time of the scrubbing liquid in the scrubber basin, the reactions of SO_2 and NO to other components with a higher solubility have time to perform and the recirculated scrubbing liquid can absorb additional pollutants.

7.2.3 Flue gas reheater

After the quench cooler and the scrubber column the flue gas is saturated with water vapor. Further cooling and compression of the flue gas with an induced draft fan would lead to condensation of water. This could damage the equipment due to corrosion or the impact of the droplets on the fan blades. Therefore, a thermal oil / flue gas heat exchanger was installed, which uses a small part of the energy from the heat recovery system of the incinerator to reheat the flue gas. It is situated between the scrubber column and the fan, as shown in Figure 7.1 or, in detail, in Figure 7.9.

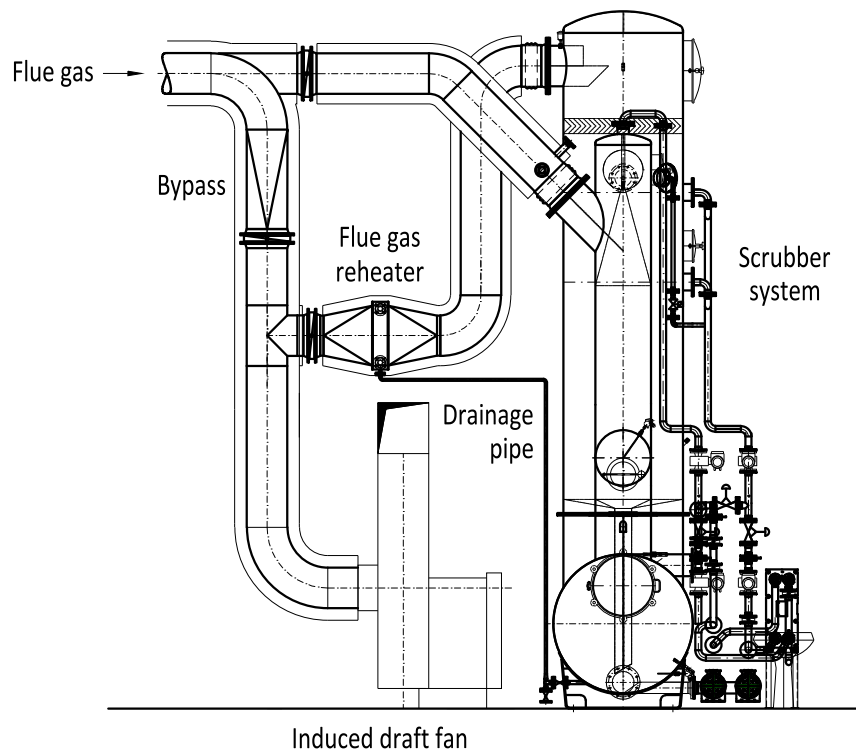


Figure 7.9: Installation of the flue gas reheater

7 Development of the scrubber system

Thermal design and construction Generally, a temperature of 10 - 20 K above the actual dew point temperature is estimated to be save in terms of condensation. Therefore, the flue gas reheater was designed to increase the flue gas temperature by 20 K. A thermal power of approx. 60 kW is needed for the design flue gas stream of 8500 Nm³/h (wet). This is provided by a thermal oil side stream of 5.4 m³/h with a temperature drop from 160°C to 140°C.

The heat exchanger is built as a finned tube heat exchanger with 0.4 mm fins on 25 x 1.5 mm core tubes. Condensation of water vapor can occur in the area of the heat exchanger, therefore, it is made of 1.4571 stainless steel like the columns. For the discharge of condensate, which might gather in the heat exchanger, a drainage pipe is installed, which leads the condensate to the scrubber basin. Both ends of the heat exchanger are equipped with cones for the connection of the rectangular heat exchanger to the circular flue gas channel.

7.2.4 Scrubber basin and liquid distribution system

Both columns are mounted on top of a common tank, the scrubber basin. The used scrubbing liquid is collected in this tank and led to the recirculation pump. Next to the recirculation pump the fresh scrubbing liquid pump is installed. Together these two pumps deliver the scrubbing liquid to the columns. Figure 7.10 shows the basin with the columns and the scrubbing liquid distribution system.

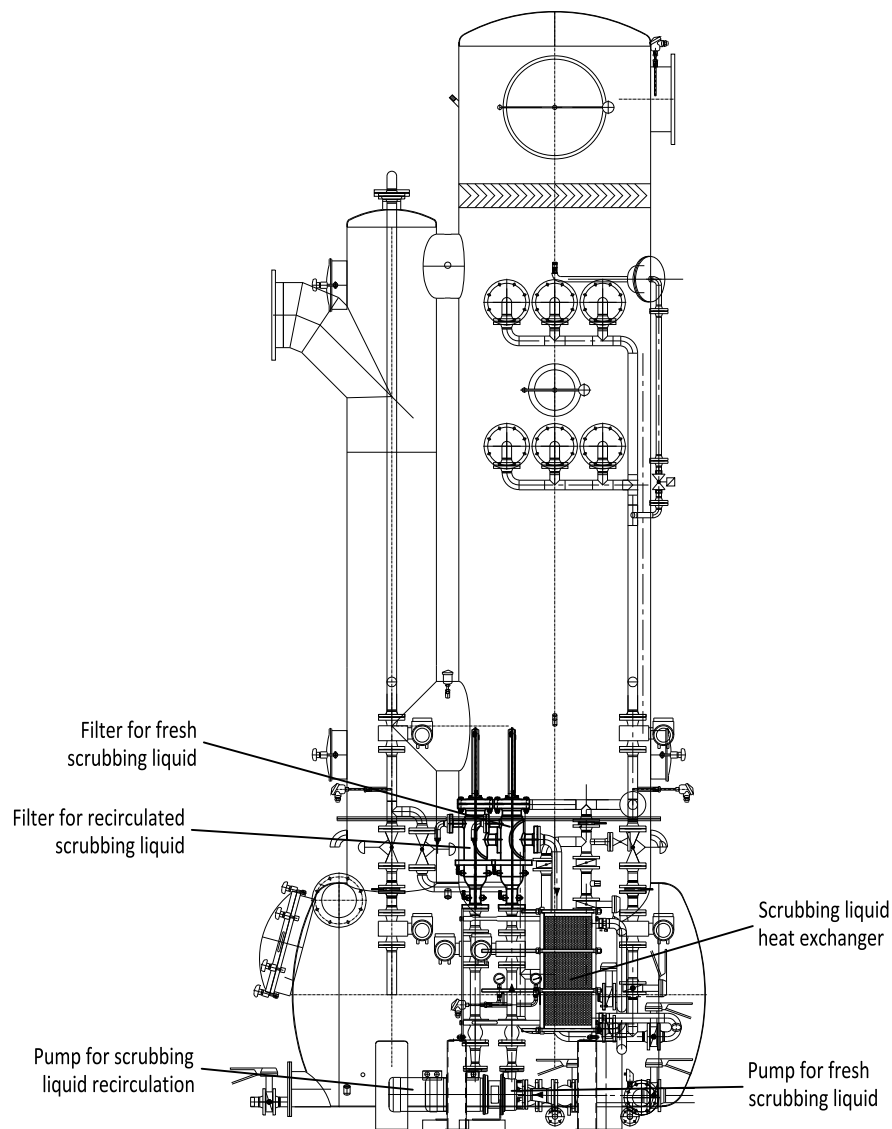


Figure 7.10: Scrubber basin with quench and scrubber column

7 Development of the scrubber system

Construction of the basin The basin has a diameter of 1400 mm, is 2600 mm long and can store a volume of approx. 3200 lt. It is made of 1.4571 stainless steel and is equipped with a manhole for internal inspection and cleaning. As previously mentioned, the used scrubbing solution from the quench and the scrubber column is led into the tank through dip pipes. They act as siphons and separate the gas inside the columns from the air above the liquid level in the scrubber basin. The air space in the basin is connected with the environment with two pipes at the top section, which allow pressure equalization. Excess scrubbing solution leaves the basin through an overflow pipe to the internal sewage drain of the plant. The residence time of the scrubbing liquid depends on the fresh scrubbing liquid flow and is between 20 minutes at a fresh scrubbing liquid flow rate of $10 \text{ m}^3/\text{h}$ and 6 minutes at a fresh scrubbing liquid flow rate of $32 \text{ m}^3/\text{h}$.

Construction of the scrubbing liquid distribution system In Figure 7.11 the flow of the fresh and the recirculated scrubbing liquid is shown.

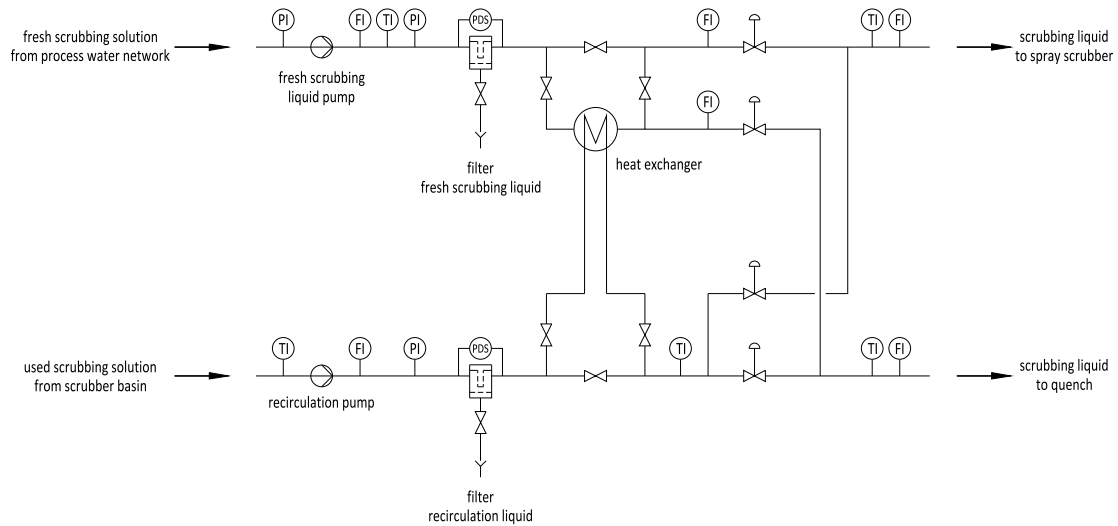


Figure 7.11: Scheme of the scrubbing liquid distribution system

7 Development of the scrubber system

The recirculated scrubbing liquid is pumped out at the bottom of the tank. For this purpose a centrifugal pump is used. After the pump a backwash filter is installed to remove particles from the scrubbing liquid. After the filter, the used scrubbing liquid is utilized to preheat the fresh scrubbing solution or it is led directly to the columns. The flow to each column is controlled with a membrane valve and measured with a magnetic-inductive flow meter.

The fresh scrubbing solution is pressurized in the fresh scrubbing solution pump, which is the same type like the recirculation pump. Like the recirculated fluid, also the fresh fluid is filtered as it might contain small parts of algae or other residues from the water treatment process. After the filter the scrubbing solution is either led directly to the scrubber column or to the quench column, or it is preheated in counterflow with the used scrubbing solution and afterwards led to the columns. Equal to the used scrubbing solution, the flow of fresh liquid is controlled with membrane valves and measured with magnetic-inductive flow meters.

The installed heat exchanger is a plate heat exchanger with a nominal thermal capacity of 800 kW. Depending on the flow, the temperature of the fresh scrubbing liquid can be regulated between 15 °C and 40 °C.

7.2.5 Scrubbing liquid regeneration

The scrubbing liquid regeneration takes place in the affiliate wastewater treatment plant. As previously described the used scrubbing liquid is led into the internal sewage drain, which is directed to the main bioreactor. There the solved pollutants are treated (Fig. 7.12).

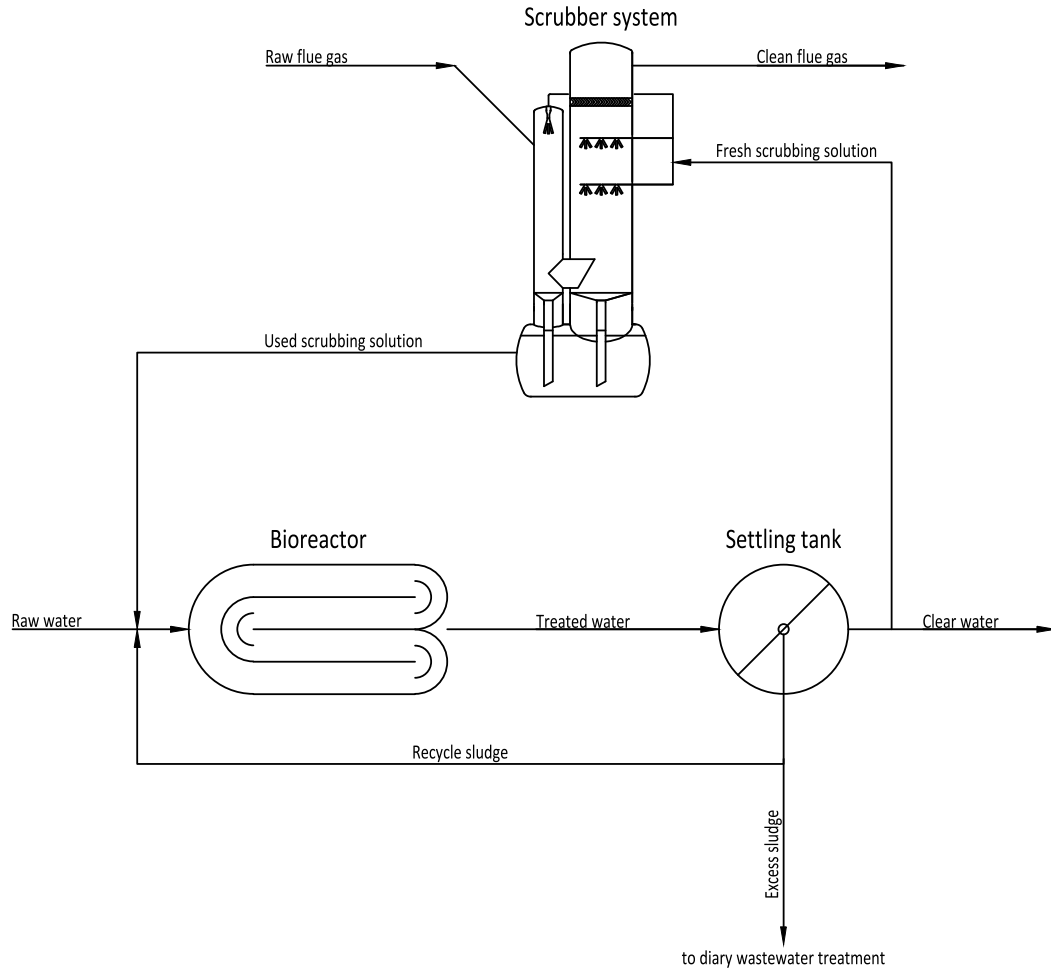
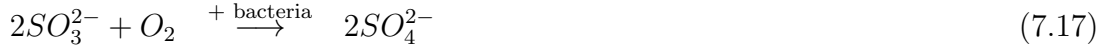


Figure 7.12: Regeneration of the scrubbing liquid

SO₂ treatment In the bioreactor the dissolved sulfur dioxide is oxidized to sulfite (SO_3^{2-}) and further to sulfate (SO_4^{2-}). In the first step, SO_2 reacts with water to sulfurous acid, which disproportionates to SO_3^{2-} and H^+ (eq. (7.15) and (7.16)).

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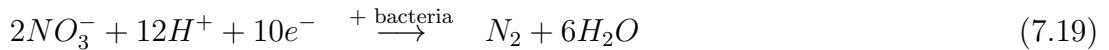
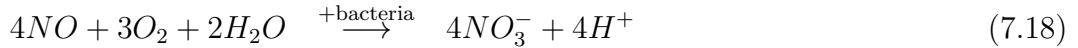
In a second step, bacteria (e.g. from genus *Thiobacillus*) oxidize the SO_3^{2-} to SO_4^{2-} (eq. (7.17)). This process consumes oxygen, which has to be provided by the aeration system of the bioreactor [37, 38, 39].



For oxidation of two mole sulfur dioxide one mole of oxygen is needed, which is equal to 0.25 kg oxygen for each 1 kg SO_2 . Based on the design concentration of 500 mg/Nm³ and a flue gas flow of 8500 Nm³/h, a maximal amount of 4.25 kg/h of SO_2 is led to the bioreactor. This means an oxygen consumption of approx. 1 kg/h, additional to the regular consumption of the bioreactor.

The produced sulfate remains in the effluent of the wastewater plant. Based on the effluent flow of 1200 - 2000 m³/h, an additional sulfate content of 3-5 mg/l is expected. Currently the effluent has a fluctuating sulfate content of 40 to 80 mg/l.

NO_x treatment The dissolved nitrogen oxides are oxidized to nitrite (NO_2^-) and nitrate (NO_3^-) and further reduced to gaseous nitrogen through different microbiological processes. The first step, called nitrification, is the oxidation of NO to NO_2^- and further to NO_3^- , which is executed by bacteria like *Nitrobacter* (eq. (7.18)). In this step oxygen is consumed, therefore, it happens in the oxic section of the bioreactor. In the second step, nitrate is reduced to nitrogen in the denitrification step. In the anoxic section of the bioreactor bacteria like *Paracoccus denitrificans* use the oxygen from the nitrate for their metabolism and produce nitrogen (eq. (7.19)) [37, 38, 39].



One mole of oxygen is spent for oxidation of each mole nitrogen monoxide, respectively 1.1 kg of oxygen for each 1 kg of NO. The assumed NO concentration of

7 Development of the scrubber system

200 mg/Nm³ and an estimated removal efficiency of 0.5 would lead to an NO flow of 0.85 kg/h to the bioreactor. This would cause an additional oxygen consumption of 0.9 kg/h.

HCL treatment The dissolved hydrochloric acid remains in the wastewater and leads to an increase of the salt level of the effluent. The wastewater plant effluent has a fluctuating chloride content of 60 to 100 mg/l. Based on the design value of 100 mg/Nm³ in the flue gas and the effluent flow of 1200-2000 m³/h, the chloride content in the effluent would be raised by approx. 1-2 mg/l.

Effect on the operation of the wastewater plant The aeration capacity of the bioreactor is around 14,000 kg of oxygen per day [40, 41], which means that the additional consumption of 46 kg/day is marginal. Also the increase of the salt level in the effluent is minor. Therefore no additional activities are necessary to treat the used scrubbing liquid from the scrubber system.

7.2.6 Control system

For the operation of the scrubber system a control system was installed. A central switching cabinet was built, which hosts not only the fuses, power switches and frequency converters but also a PLC. This system allows to control the pumps and valves and to collect all the data recorded from the different measurements. For this purpose the PLC is connected to the SCADA system of the wastewater plant, where a separate tab was programmed to visualize the scrubber system (Fig. 7.13).

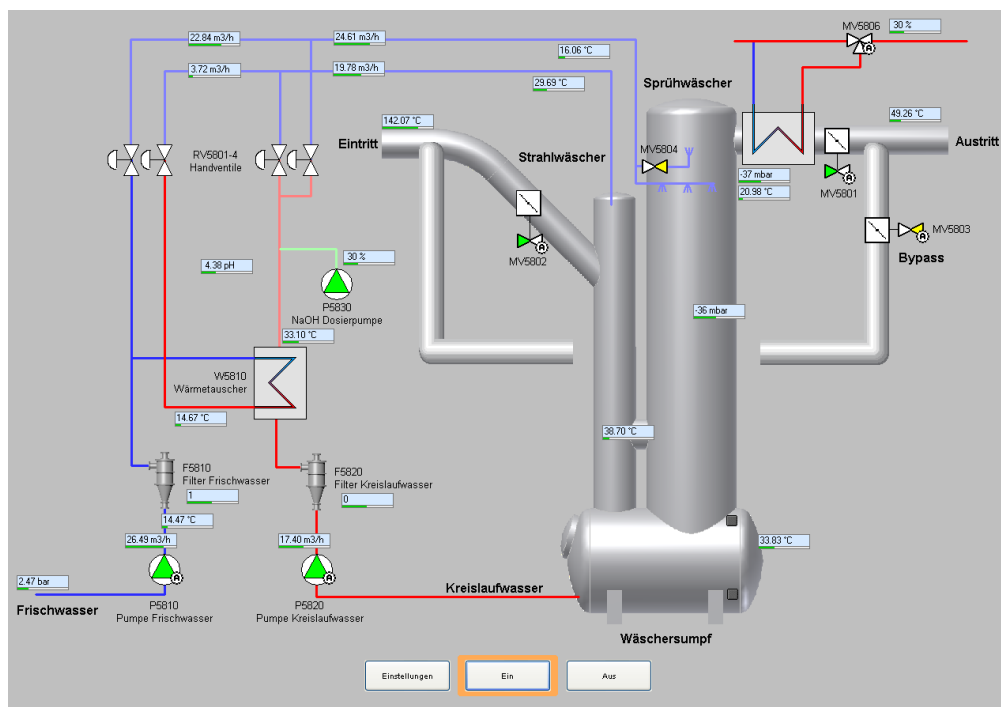


Figure 7.13: Main SCADA tab of the scrubber system

In this way the scrubber system is operated from a PC screen in the control room of the incineration plant, but it is also possible to operate it from any other PC where the SCADA system is installed. The tap shows all measured volume flows, temperatures and pressures, the current operating state of the pumps and the po-

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sition of the valves. All these data are recorded, stored in a database and can be shown in trends (Fig. 7.14) or exported to excel sheets.

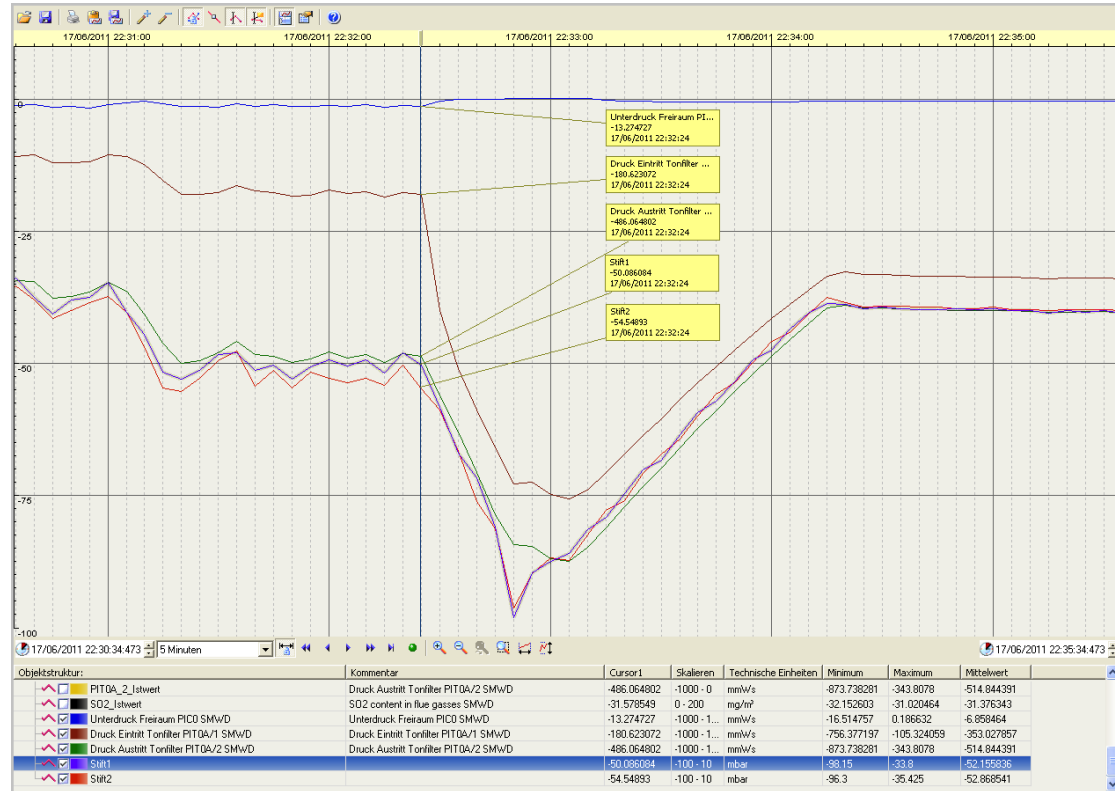


Figure 7.14: Trend tab of the scrubber system

The PLC contains also programmed sequences for the startup and shutdown of the scrubber system. Interlock sequences ensure a safe operation, as they prevent operating states which could damage the equipment or be harmful to the environment or the operating personnel. Depending on the occurred failure, these sequences direct the system back to a safe operational state.

The previously mentioned frequency converters allow to control the speed of the pumps. Therefore, the membrane valves are only used for fine adjustment of the particular flow. In this way the pressure drop and the power consumption of the pumps is reduced.

8 Experiments

Beside the development of the new scrubber system the main focus of the work was to determine the mechanism of the absorption processes. Therefore an extensive test program was run at the scrubber pilot plant and both, flue gas and scrubbing liquid were analysed before and after the scrubber.

The test run schedule had to be coordinated with the operation hours and process requirements of the sludge incineration and wastewater treatment plant Bad Vöslau. The sludge incineration plant had several unplanned stops during the experimental series, which made continuous measurements quite difficult. As the fuel composition and the combustion conditions in the sludge incineration plant could not be modified, it was not possible to adjust the flue gas composition. Also the properties of the fresh scrubbing liquid could not be changed, because it is taken directly from the effluent of the wastewater plant.

8.1 Description of the measuring equipment

8.1.1 Gas analysis

For simultaneous analysis of the raw and clean flue gas two sets of gas measuring equipment are used which are both connected to a data recorder. The positions of both gas measuring systems are shown in Figure 8.1, detailed flowsheets with all measuring points for temperature, flow, pressure etc. are enclosed in Appendix C and D.

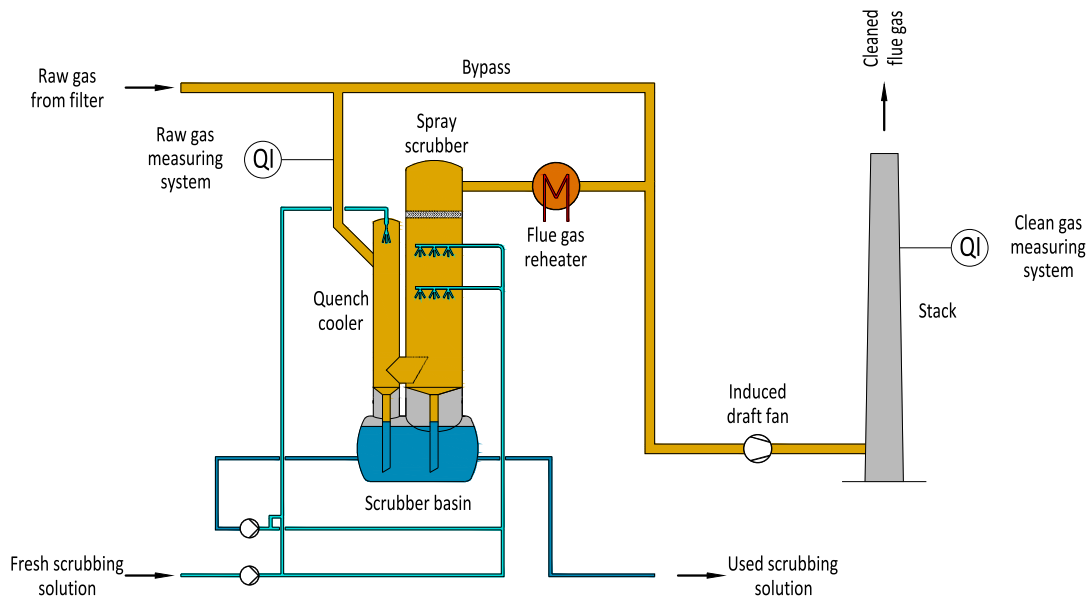


Figure 8.1: Positions of the raw and clean gas measuring system

Raw gas analysis The raw gas analysis system consists of three main parts: the probe with an integrated gas filter to remove particles, the gas pump and the gas analyser. The gas pump is a diaphragm pump which draws the gas through the probe and a heated gas tube to the analyser. At the gas pump also an inlet for calibration gas is installed (Fig. 8.2).

As gas analyzer a SICK MCS 100 E is used, which is equipped for measuring of NO, NO₂, N₂O, CO, SO₂, HCl, H₂O, NH₃, HF und O₂. The measuring limits and the measuring principles are shown in Table 8.1.

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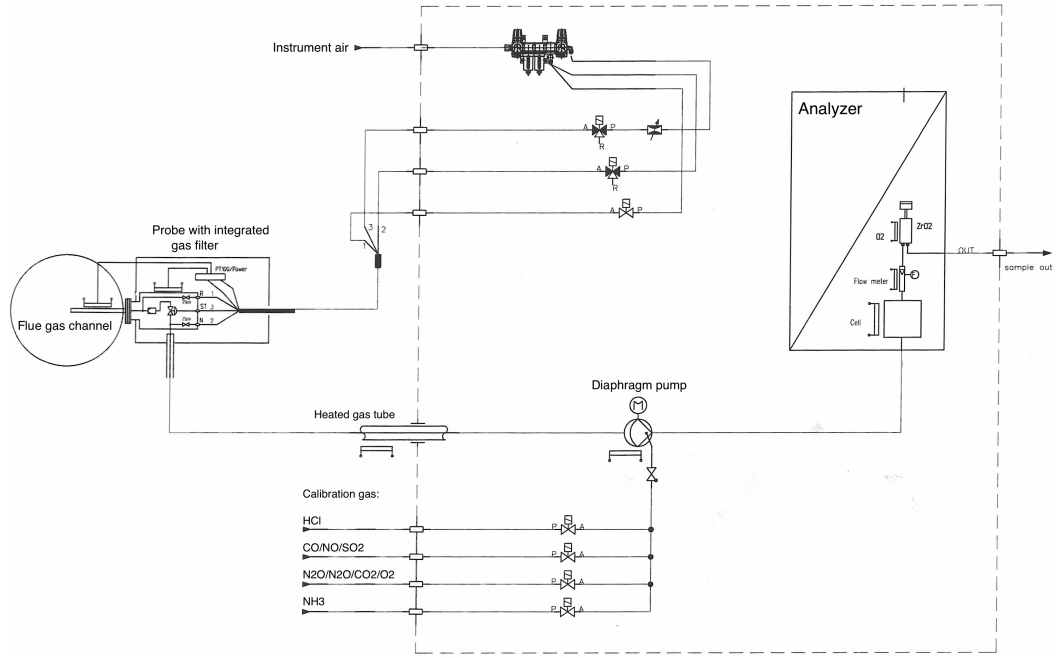


Figure 8.2: Scheme of the raw gas measuring equipment [42]

Component		concentration range		measuring principle
Nitrogen monoxide	NO	0-400	mg/Nm ³	NDIR
Nitrogen dioxide	NO ₂	0-100	mg/Nm ³	NDIR
Nitrous oxide	N ₂ O	0-100	mg/Nm ³	NDIR
Sulfur dioxide	SO ₂	0-300	mg/Nm ³	NDIR
Carbon monoxide	CO	0-150	mg/Nm ³	NDIR
Carbon dioxide	CO ₂	0-20	mg/Nm ³	NDIR
Hydrogen chloride	HCl	0-90	mg/Nm ³	NDIR
Ammonia	NH ₃	0-20	mg/Nm ³	NDIR
Water vapor	H ₂ O	0-40	vol%	NDIR
Oxygen	O ₂	0-21	vol%	zirconium dioxide sensor

Table 8.1: Measuring ranges and principles of the Sick MCS 100 E gas analyser

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Clean gas analysis Similar to the raw gas the clean gas analysis system consists of a probe with an integrated gas filter, a gas pump and the gas analyser. The gas pump is an injector pump which draws the gas not only through the gas probe and the heated tube but also through the gas analyser (Fig. 8.3).

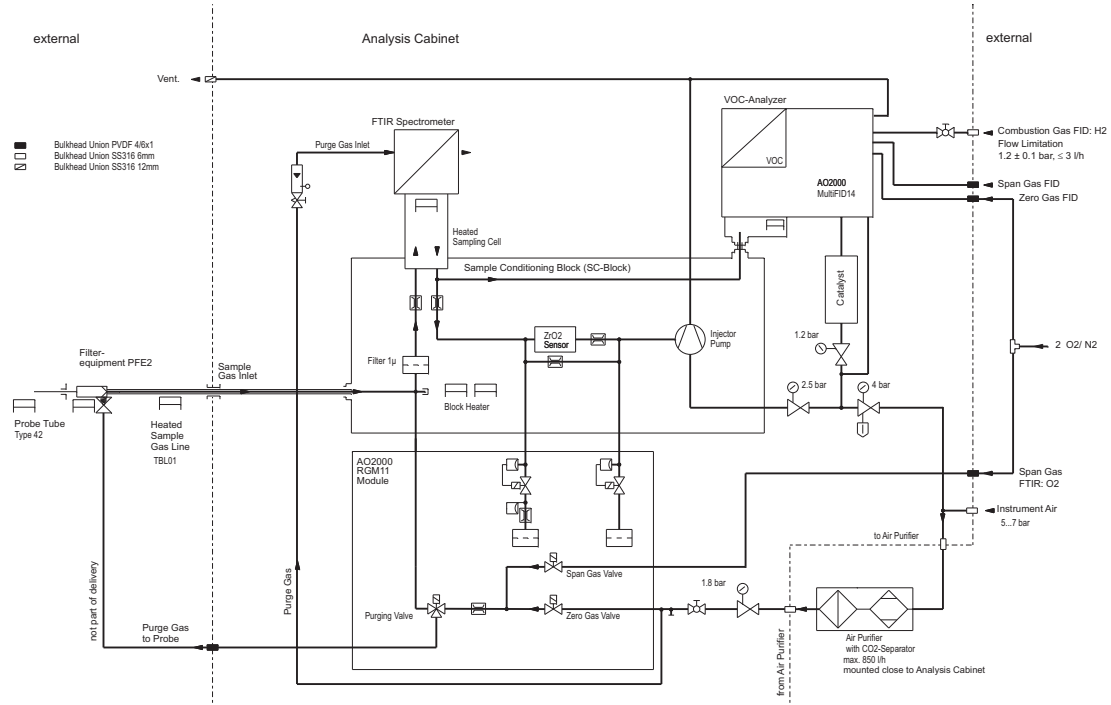


Figure 8.3: Scheme of the clean gas measuring equipment [43]

For the clean gas analysis an ABB Advance CEMAS FTIR is used. It is equipped with a FTIR spectrometer, a zirconium dioxide sensor and a FID analyzer for hydrocarbons. Table 8.2 shows the concentration limits and the measuring principles of all measured components.

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Component		concentration range		measuring principle
Nitrogen monoxide	NO	0-600	mg/Nm ³	FTIR
Nitrogen dioxide	NO ₂	0-200	mg/Nm ³	FTIR
Sulfur dioxide	SO ₂	0-150	mg/Nm ³	FTIR
Carbon monoxide	CO	0-100	mg/Nm ³	FTIR
Hydrogen chloride	HCl	0-30	mg/Nm ³	FTIR
Water vapor	H ₂ O	0-40	vol%	FTIR
Oxygen	O ₂	0-25	vol%	zirconium dioxide sensor
Total hydrocarbons	THC	0-15	mg/Nm ³	FID

Table 8.2: Measuring ranges and principles of the ABB Advance CEMAS FTIR gas analyser

8.1.2 Water analysis

For analysis of the water samples of the scrubbing liquid standardized cuvette tests, produced by Hach-Lange, are used. After preparation of the samples they are analysed with a LANGE Xion 500 spectrophotometer. Additionally, the oxygen content and the pH-value of the samples are measured with a HACH HQ40d multimeter. In Table 8.3 the type of test and the measuring limits for the analysed components are shown.

Component		range		used equipment
Oxygen	O ₂	0-20	mg/l	O ₂ -probe
pH-value		2-14		pH-probe
Nitrite	NO ₂ ⁻	0.05-5	mg/l	LCK341 + spectrophotometer
Nitrate	NO ₃ ⁻	1-60	mg/l	LCK339 + spectrophotometer
Sulfite	SO ₃ ²⁻	0.1-5	mg/l	LCK654 + spectrophotometer
Sulfate	SO ₄ ²⁻	40-150	mg/l	LCK153 + spectrophotometer
Chloride	Cl ⁻	1-1000	mg/l	LCK311 + spectrophotometer

Table 8.3: Measuring ranges and test types for the water samples

8.1.3 Test rig for investigation of the absorption mechanism

As the scrubber water for the scrubber pilot plant comes directly from the sewage plant outlet it is not possible to change the scrubbing liquid composition. Therefore, a small test rig was built to perform tests with a variable scrubbing liquid composition (Fig. 8.4). The test rig consists of four fritted gas washing bottles, which are connected via silicon tubes. Each bottle can be filled with up to 400 ml scrubbing liquid.



Figure 8.4: Test rig for investigation of the absorption mechanism

This test rig can be connected to the raw gas analysis system and be fed either with gas from the plant or from a gas bottle, which is filled with a synthetic flue gas mixture. To avoid under- or overpressure in the system during the setup of an experiment a bypass can be opened, which is closed as soon as the correct gas flow is set and the measurement starts (Fig. 8.5).

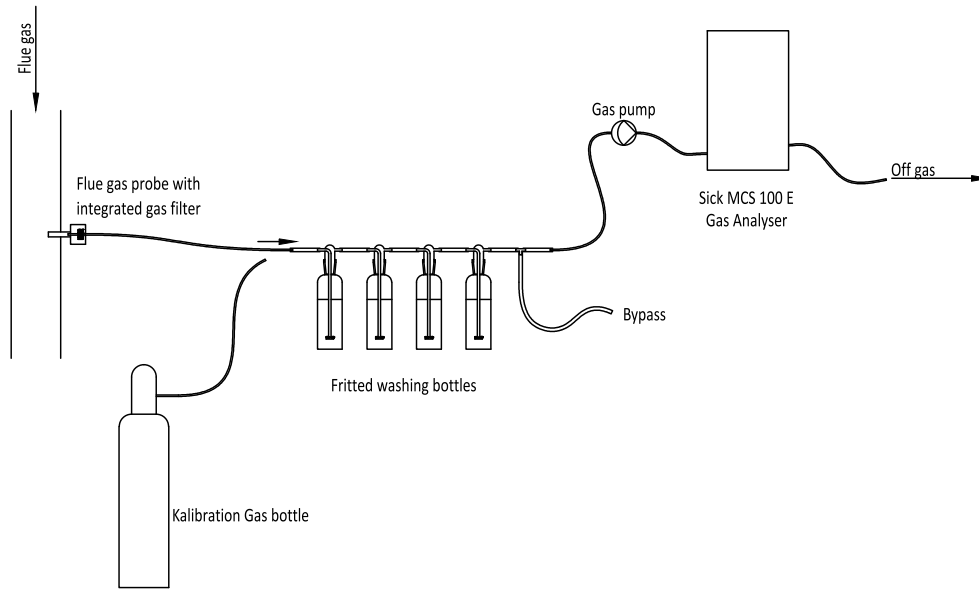


Figure 8.5: Scheme of the test rig

8.2 Experiments

As already mentioned before the experimental part of this work should deliver data about the capabilities and potential of the scrubber system at different operating states and should help to investigate and better understand the absorption mechanisms, especially for NO_x -absorption. These experiments were mainly carried out at the newly designed scrubber pilot plant described in chapter 7, for some additional experiments the small test rig which is described in section 8.1.3 was used.

For all experimental runs the sludge incinerator was brought to a steady state and for all experiments fired with pure sewage sludge. The automatic control system of the sludge incineration plant is active during the test runs and controls the process parameters for the combustion reactor, the heat recovery system

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and the flue gas cleaning system. An analysis of the dried sewage sludge (DSS) is shown in Table 8.4. Especially the contents of sulfur and chlorine vary quite a lot, which leads also to a variation of the flue gas concentrations of SO_2 and HCl .

Analysis		typical	range
<i>proximate</i> (wt%)			
moisture		20	18 - 24
ash		36	30 - 38
<i>ultimate</i> (wt% dry basis)			
Carbon	C	29.2	29.1 - 30.0
Hydrogen	H	4.3	4.2 - 4.3
Nitrogen	N	4.0	3.9 - 5.1
Sulfur	S	0.9	0.13 - 1.70
Chlorine	Cl	0.2	0.12 - 0.34
Oxygen	O	16.4	15.8 - 16.7
<i>LHV</i> (MJ/kg dry basis)		10.8	9.9 - 11.7
<i>ash softening temperature</i> ($^{\circ}\text{C}$)		1080	n.a.

Table 8.4: Fuel characterization of dried sewage sludge from Bad Vöslau

During the experimental runs the liquid-to-gas ratio, the inlet temperature of the flue gas, the recirculation ratio of the scrubbing liquid and the temperature of the scrubbing liquid were varied. A detailed list of all experiments is shown in Table 8.5. A scrubbing liquid rate of $10 \text{ m}^3/\text{h}$, according to a liquid-to-gas ratio of approx. 1.6, a flue gas temperature of approx. 60°C , a recirculation ratio of 0 and a scrubbing liquid temperature of about 19°C was chosen as base case and repeated within each series of experiments. In Table 8.5 these "base case runs" are marked bold. The letter in the experiment number stands for the variated parameter (F: scrubbing liquid flow, R: recirculation ratio, T: scrubbing liquid temperature), the number after the letter stands for the value of the parameter.

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Experiment	liquid-to-gas ratio	recirculation ratio	inlet temperature flue gas	inlet temperature scrubbing liquid
	[-]	[-]	[°C]	[°C]
01-a-F10	1.6	0	56.7	19.5
01-a-F15	2.4	0	53.9	18.9
01-a-F20	3.2	0	51.5	18.5
01-b-F10	1.6	0	60.4	18.5
01-b-F15	2.4	0	60.6	18.0
01-b-F16	2.4	0	61.0	13.0
01-b-F19	2.8	0	59.0	13.0
01-b-F20	3.2	0	61.7	17.7
02-R00	1.8	0	57.2	19.4
02-R50	2.6	0.5	60.4	33.8
02-R100	3.3	1	59.7	39.6
02-R150	4.0	1.5	59.9	42.5
03-a-T19	2.4	0	53.9	18.9
03-a-T24	2.4	0	55.7	24.3
03-a-T30	2.4	0	53.1	30.0
03-a-T37	2.4	0	49.6	37.0
03-b-T18	3.2	0	51.6	18.5
03-b-T30	3.1	0	50.6	30.1

Table 8.5: List of test conditions at the experiments

The *liquid-to-gas ratio* is defined as the total molar flow of the scrubbing liquid divided by the molar flow of the flue gas.

$$LGR = \frac{\dot{L}}{\dot{G}} \quad (8.1)$$

It is adjusted by variation of the scrubbing liquid stream as the flue gas stream is given by the combustion conditions in the incineration reactor.

8 Experiments

The *recirculation ratio* is defined as the molar flow of the recirculated scrubbing liquid divided by the molar flow of the fresh scrubbing liquid.

$$R = \frac{\dot{L}_r}{\dot{L}_f} \quad (8.2)$$

For variation of the recirculation ratio the flow of recirculated scrubbing liquid is varied, while the fresh liquid intake is held constant.

The *inlet temperature of the flue gas* is measured in the connection pipe between the quench column and the scrubber column and controlled by means of the liquid flow in the quench.

The *scrubbing liquid temperature* is measured right before the scrubbing liquid distributor of the spray scrubber column. It can be modified by preheating the fresh scrubbing liquid in countercurrent flow to the used scrubbing liquid as described in chapter 7.2.

Beside the parameter variation the effect of the continuous changing composition of the scrubbing liquid on the scrubbing process was monitored. As the composition of the scrubbing fluid cannot be modified the already mentioned test rig (chapter 8.1.3) was used for further investigations and clarification of the observations.

As the raw gas concentrations of the observed components were not constant during the series of experiments and the changing water content caused dilution effects, the clean gas concentrations could not be compared directly. Therefore, the flue gas concentrations for all components were converted to dry gas (eq. 8.3) and based on this the *removal efficiency* was calculated with equation (8.4).

$$c_{x,dry} = \frac{c_x}{1 - c_{water}} \quad (8.3)$$

$$\eta_{rem,x} = 1 - \frac{c_{x,dry,clean}}{c_{x,dry,raw}} \quad (8.4)$$

8.3 Results and discussion

In this section the results of the experimental series are summarized, the detailed results of all experimental runs are enclosed in Appendix B. First the influence of the liquid-to-gas ratio and the gas temperature are discussed. These effects have been investigated during the first experimental series (experiments 01-a-F10 to 01-b-F20). Afterwards the influence of scrubbing liquid recirculation (experiments 02-R00 to 02-R20) and the influence of the scrubbing liquid temperature (experiments 03-a-T19 to 03-b-T30) are discussed. Finally a qualitative summary of the observed behavior is given.

8.3.1 Influence of the liquid-to-gas ratio and gas temperature

During the first series of experiments the influence of the liquid-to-gas ratio (LGR) and the flue gas temperature at the inlet of the scrubber column on the removal efficiency of NO, SO₂ and HCl was investigated.

Figure 8.6 shows the effect of the LGR on the NO removal efficiency for two different gas temperatures. It is visible that the removal efficiency for NO increases with higher liquid-to-gas ratios reaching values of nearly 0.35 and even more. This meets the expectations as the increasing fresh scrubbing liquid flow, which causes the higher liquid-to-gas ratios, absorbs more NO and leads also to a higher contact surface between gas and scrubbing liquid and, therefore, a higher mass transfer. On the other hand the gas temperature at the inlet of the scrubber column has no significant influence. The big scatter of the results is related to changing compositions of the scrubbing fluid during the experimental series and will be explained in chapter 9.

In Figure 8.7 the effect of the liquid-to-gas ratio on the removal efficiency of SO₂ is shown. Similar to NO the efficiency increases with higher liquid-to-gas ratios and reaches a maximum of 0.95. At lower LGRs a significant influence of the inlet temperature is visible which is reduced at higher values. This effect is caused through the reduced solubility of the gas at higher temperatures, with

8 Experiments

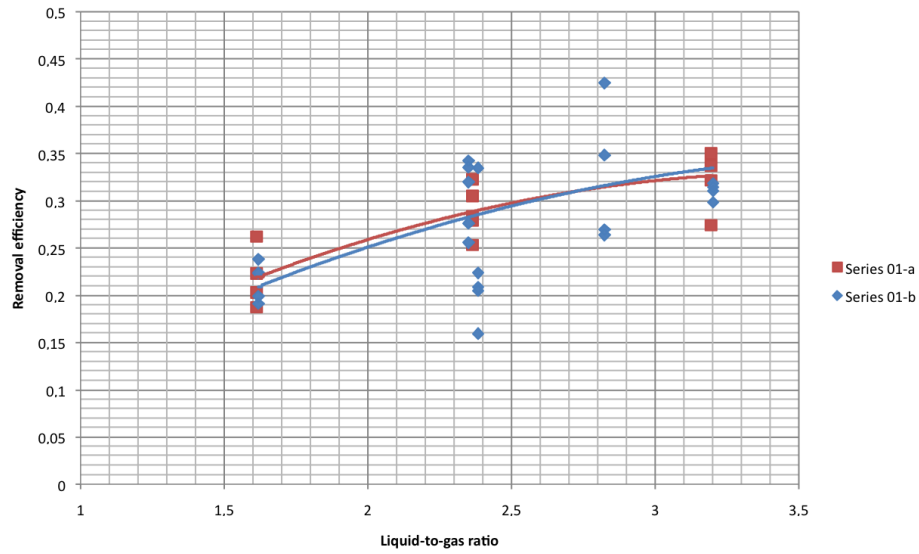


Figure 8.6: Effect of the LGR on the NO removal efficiency

higher scrubbing liquid flows the average contact temperature between gas and liquid is more and more dominated by the temperature of the scrubbing liquid, which was the same for both series.

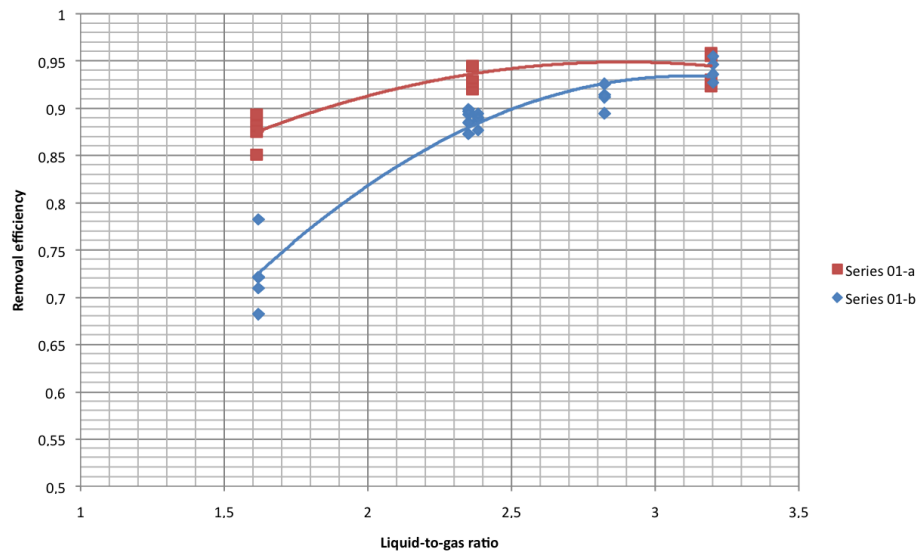


Figure 8.7: Effect of the LGR on the SO₂ removal efficiency

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Figure 8.8 illustrates the effect of the liquid-to-gas ratio on the HCl removal. Similar to NO and SO₂, the removal efficiency is increasing with higher liquid-to-gas ratios. The removal efficiency is close to 1 in all experiments, with the lower inlet temperature removal efficiencies of 0.99 are reached while the efficiency at the higher temperature is about 0.98. Unlike to SO₂ the gap is not closing with higher LGRs, which could be explained with very high solubility of HCl in water: As the water partly condenses in the quench, some of the HCl is already removed there. The lower the inlet temperature is, the more water condenses and also more cooling water gets in contact with the gas. Therefore, at the lower temperature the quench works as a second stage for HCl removal, which leads to higher removal efficiencies.

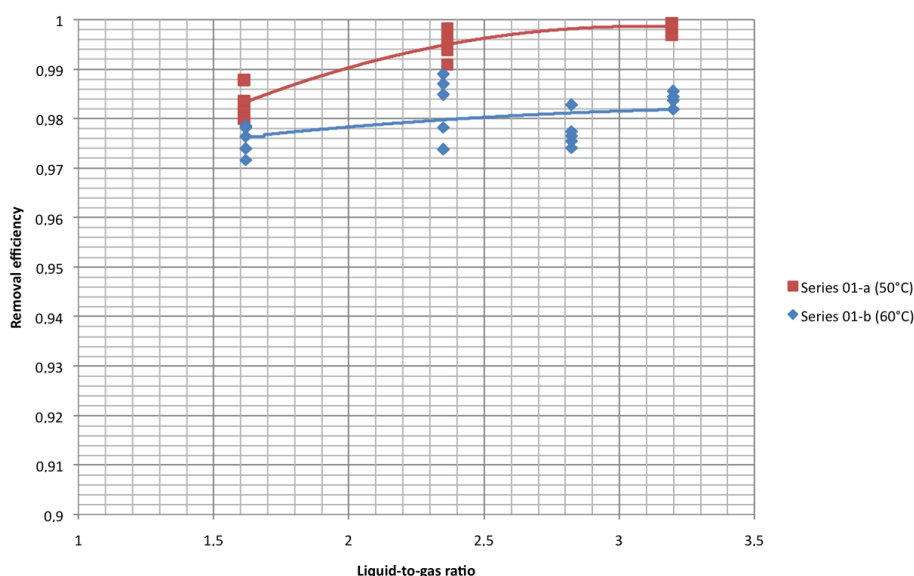


Figure 8.8: Effect of the LGR on the HCl removal efficiency

Table 8.6 shows a summary of the average gas concentrations of water, nitrogen monoxide, sulfur dioxide and hydrochloric acid during the test runs for the raw and the clean gas sample point. The detailed data for all components is enclosed

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in Appendix B. While the raw gas concentration of NO was quite constant during the series, the concentrations of SO₂ and HCl varied a lot. The latter is related to the varying concentrations of sulfur and chlorine in the sewage sludge. The clean gas concentrations of H₂O correspond with the dew point of the flue gas after the scrubber and varies according to the different exit temperatures.

Experiment	Sample point	H ₂ O [vol%]	NO [mg/Nm ³]	SO ₂ [mg/Nm ³]	HCl [mg/Nm ³]
01-a-F10	raw gas	19.2	33.0	66.1	3.3
	clean gas	12.1	25.8	8.1	0.1
01-a-F15	raw gas	19.6	32.9	135.7	16.8
	clean gas	8.6	23.4	9.0	0.1
01-a-F20	raw gas	19.6	33.9	213.3	63.6
	clean gas	6.1	22.8	11.2	0.1
01-b-F10	raw gas	18.4	31.3	52.2	2.6
	clean gas	12.5	24.6	14.1	0.1
01-b-F15	raw gas	18.5	29.9	92.7	4.2
	clean gas	9.7	20.8	10.4	0.1
01-b-F16	raw gas	20.6	45.8	96.0	1.4
	clean gas	9.2	35.4	10.9	0.1
01-b-F19	raw gas	18.6	37.9	122.2	4.8
	clean gas	6.9	24.4	10.7	0.1
01-b-F20	raw gas	18.7	34.1	145.5	3.0
	clean gas	7.1	23.5	8.8	0.0

Table 8.6: Average gas concentrations during the first series of test runs

In Table 8.7 the results of the analysis of the pH-value, oxygen, nitrite, nitrate, sulfite and sulfate in the raw and used scrubbing liquid are shown. It is visible that pH-value dropped slightly during the absorption process, which is related to

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the acidic behavior of the absorbed gases SO_2 , HCl and also NO . Also the oxygen content dropped, which is caused by the oxidation process of SO_2 to SO_3^{2-} and SO_4^{2-} respectively NO to NO_2^- and NO_3^- . The increase of SO_3^{2-} and SO_4^{2-} displays the amount of removed SO_2 quite accurate, while the change of the NO_2^- and NO_3^- concentrations reflect only 10 - 20% of the absorbed NO . This is caused by the fact, that only a small amount of the absorbed NO is oxidized to NO_2^- or NO_3^- during the absorption process itself. A bigger part is still in an intermediate step of the oxidation process, which will be further oxidized in the bioreactor of the sewage plant. The steps of NO absorption mechanism are explained in detail in chapter 9.

Experiment	Sample point	pH	O_2	NO_2^-	NO_3^-	SO_3^{2-}	SO_4^{2-}
		[-]	[mg/l]	[mg/l]	[mg/l]	[mg/l]	[mg/l]
01-a-F15	raw liquid	7.6	7.6	0.056	2.3	2.0	65.7
	used liquid	7.1	0.3	0.127	4.2	43.5	101.0
01-a-F20	raw liquid	7.6	7.6	0.056	2.3	2.0	65.7
	used liquid	7.2	3.8	0.101	3.3	25.4	76.6
01-b-F10	raw liquid	7.7	7.9	0.074	2.7	2.1	57.6
	used liquid	7.4	4.2	0.129	4.3	10.7	68.8
01-b-F15	raw liquid	7.7	7.9	0.074	2.7	2.1	57.6
	used liquid	7.3	4.5	0.113	3.2	18.8	68.0
01-b-F16	raw liquid	7.4	5.7	0.572	7.0	1.7	63.8
	used liquid	7.4	4.5	0.512	6.9	16.8	80.2
01-b-F20	raw liquid	7.7	7.9	0.074	2.7	2.1	57.6
	used liquid	7.2	3.9	0.065	2.9	35.8	70.9

Table 8.7: Water samples of raw and used scrubbing liquid during the first series of test runs

8.3.2 Influence of the scrubbing liquid properties

In the second and third series of experiments the influence of scrubbing liquid recirculation and the scrubbing liquid temperature was investigated.

Figure 8.9 shows that the NO removal efficiency increases with higher recirculation ratios. Nevertheless the removal efficiencies are much lower than during the first series of experiments. This can be explained with the low nitrite content of the scrubbing solution as described in chapter 9.

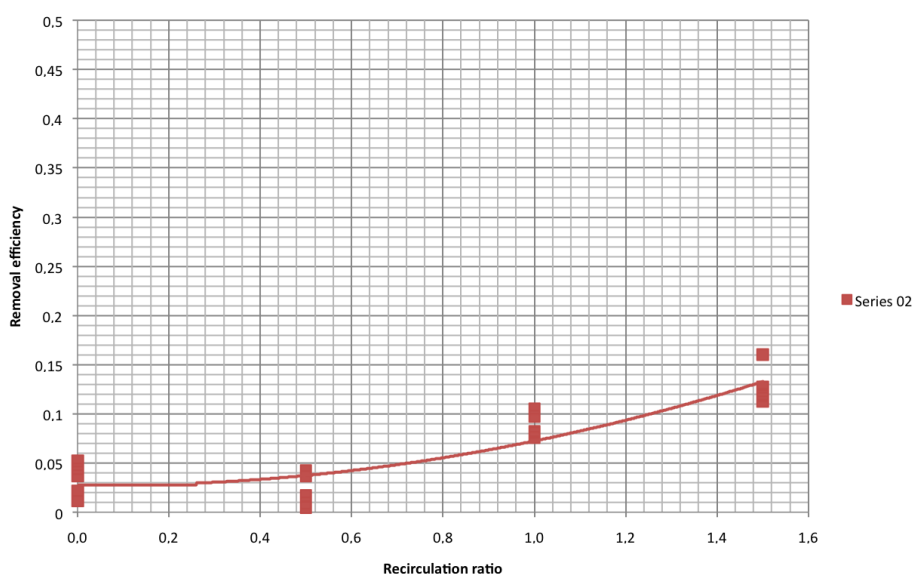


Figure 8.9: Effect of the recirculation ratio on the NO removal efficiency

In Figure 8.10 the effect of scrubbing liquid recirculation on the SO_2 removal efficiency is displayed. The efficiency increases with rising recirculation ratios, and reaches values of up to 0.95. This is comparable to the first series and indicates that the scrubbing liquid was not saturated with SO_2 . Scrubbing liquid recirculation leads to a bigger contact surface between gas and liquid with a constant consumption of fresh scrubbing liquid which helps to reach higher saturation

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levels. Another fact is, that during the residence time of the recirculated scrubbing liquid in the scrubber basin SO_2 is oxidized to SO_3^- and SO_4^- . Therefore, the recirculated scrubbing liquid can absorb SO_2 again.

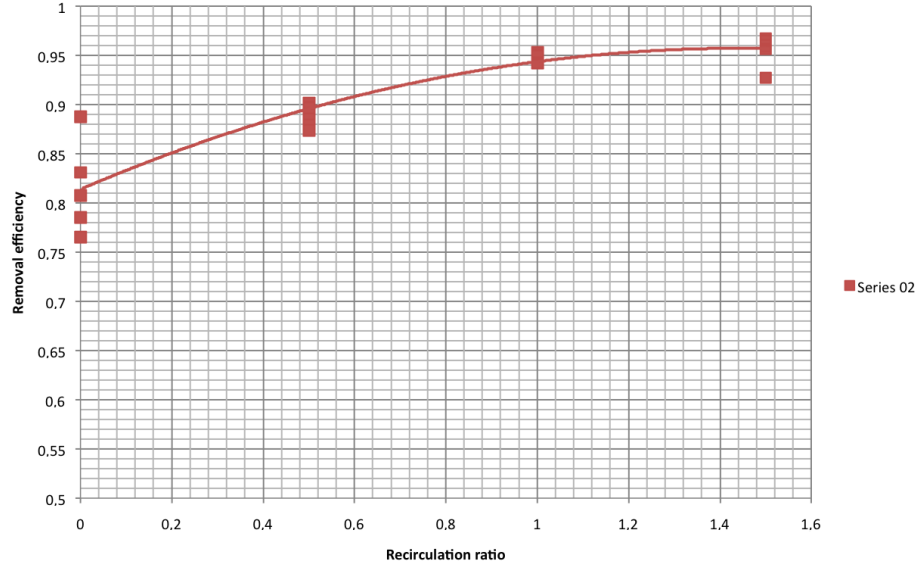


Figure 8.10: Effect of the recirculation ratio on the SO_2 removal efficiency

Figure 8.11 shows a constant removal efficiency of 1 for all recirculation ratios. This is caused by the very low HCl content of the raw gas during this experiment, which led to clean gas concentrations below the detection limit of the gas analysis system.

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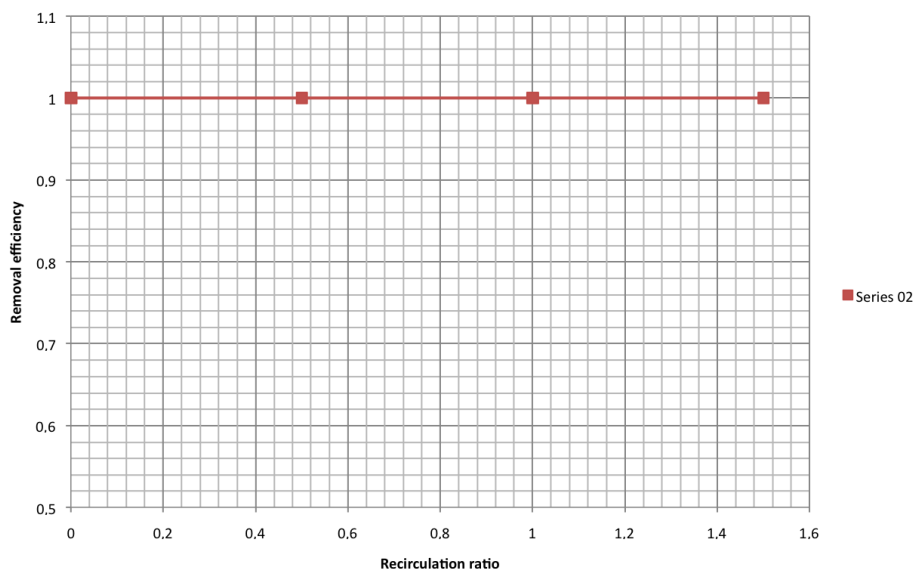


Figure 8.11: Effect of the recirculation ratio on the HCl removal efficiency

The influence of the scrubbing liquid temperature on the NO removal efficiency is illustrated in Figure 8.12. Although the data scatters quite a lot a slight tendency to higher removal efficiencies at higher temperatures is visible. This effect is caused by the higher reaction rates during the absorption process at higher temperatures and exceeds the effect of lower physical solubility at increasing temperatures. With a raised liquid-to-gas ratio (LGR was 2.4 during series 03-a and 3.2 in series 03-b) the effect is even stronger.

In Figure 8.13 the effect of increasing temperatures on the SO₂ removal efficiency is displayed. As the physical solubility of SO₂ decreases with higher temperatures, the removal efficiency decreases also. This effect is the same for both tested liquid-to-gas ratios.

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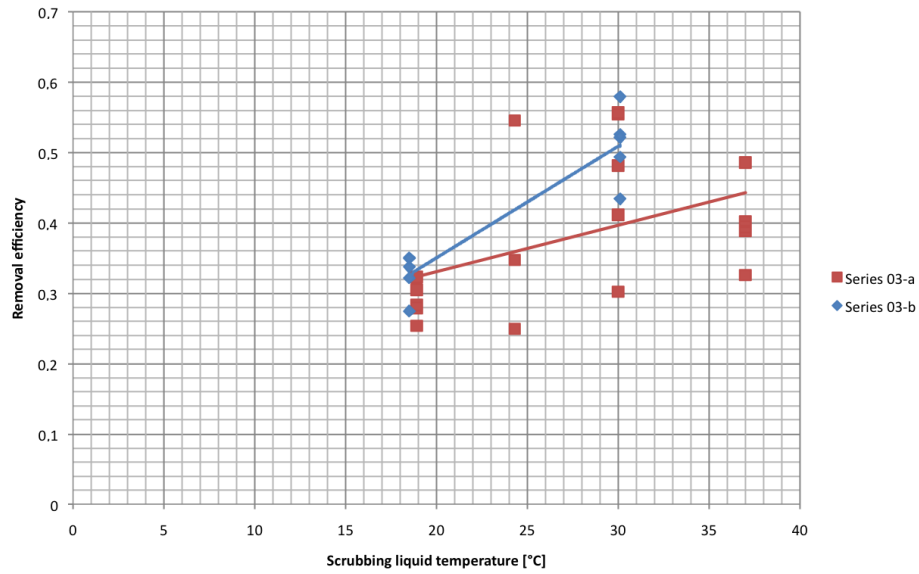


Figure 8.12: Effect of the scrubbing liquid temperature on the NO removal efficiency

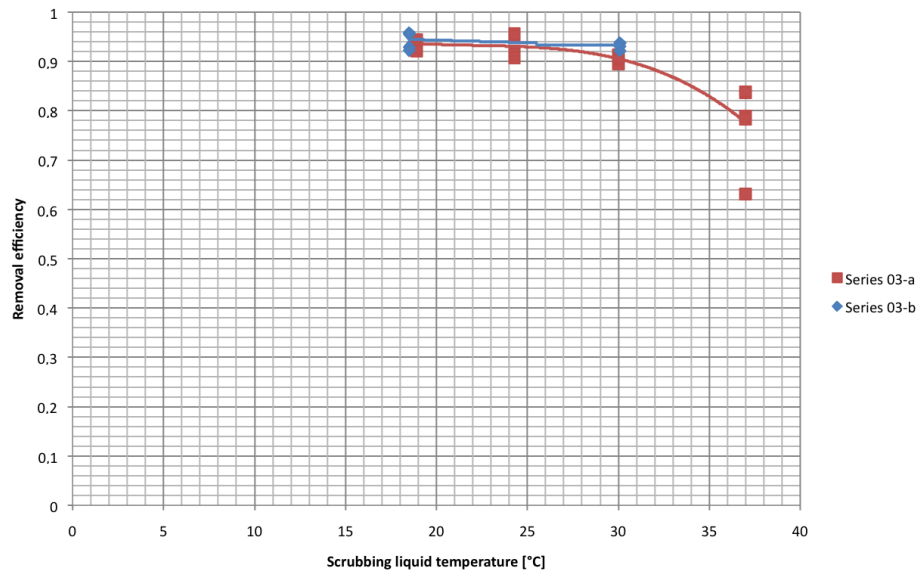


Figure 8.13: Effect of the scrubbing liquid temperature on the SO₂ removal efficiency

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Similar to SO_2 also the HCl removal efficiency decreases with increasing temperatures, which is shown in Figure 8.14. Like before, the reason for this behavior is the reduced solubility of HCl in water at higher temperatures.

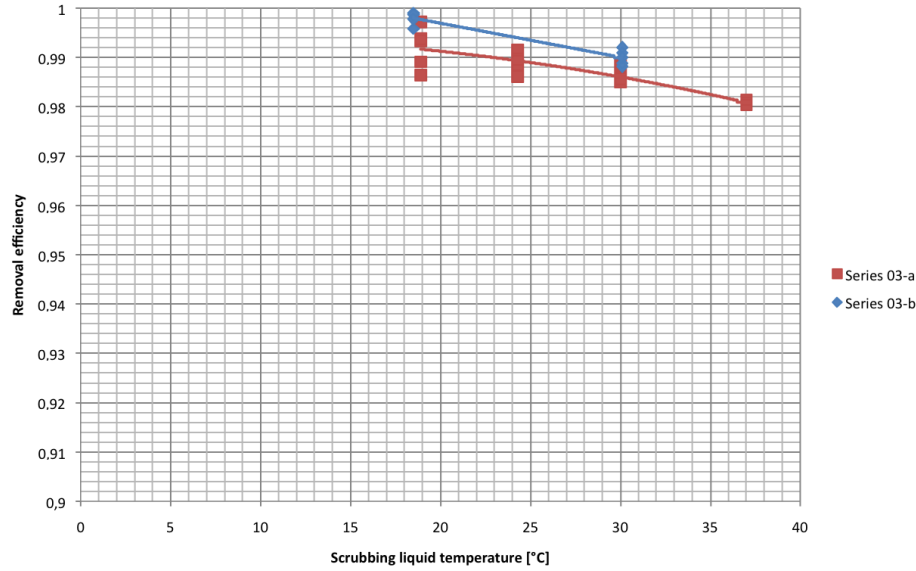


Figure 8.14: Effect of the scrubbing liquid temperature on the HCl removal efficiency

Table 8.8 shows the concentrations of water, nitrogen monoxide, sulfur dioxide and hydrochloric acid during the second and third series of test runs. Similar to the first series (Tab. 8.6), the NO concentration in the raw gas does not differ a lot, while the SO_2 and HCl concentration vary in a big range. The humidity of the clean flue gas varies due to the different outlet temperatures of the gas and the corresponding dew point.

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Experiment	Sample point	H ₂ O	NO	SO ₂	HCl
		[vol%]	[mg/Nm ³]	[mg/Nm ³]	[mg/Nm ³]
02-R00	raw gas	20.6	45.1	77.3	0.1
	clean gas	19.0	43.6	14.1	0.0
02-R50	raw gas	19.3	43.0	105.3	0.1
	clean gas	12.6	42.2	11.5	0.0
02-R100	raw gas	18.1	43.4	145.7	0.1
	clean gas	11.8	39.6	7.3	0.0
02-R150	raw gas	18.4	41.7	59.4	0.1
	clean gas	12.7	36.4	2.5	0.0
03-a-T19	raw gas	19.6	32.9	135.7	16.8
	clean gas	8.6	23.4	9.0	0.1
03-a-T24	raw gas	20.9	36.8	91.1	8.9
	clean gas	10.3	19.4	6.1	0.1
03-a-T30	raw gas	20.4	33.7	102.6	7.7
	clean gas	11.3	18.4	9.7	0.1
03-a-T37	raw gas	21.5	33.5	37.0	5.2
	clean gas	12.9	20.2	7.6	0.1
03-b-T18	raw gas	19.6	33.9	213.3	63.6
	clean gas	6.1	22.8	11.2	0.1
03-b-T30	raw gas	20.4	31.4	133.7	10.1
	clean gas	9.7	15.4	8.9	0.1

Table 8.8: Average gas concentrations during the second and third series of test runs

In Table 8.9 the results of the scrubbing liquid analysis of the pH-value, oxygen, nitrite, nitrate, sulfite and sulfate are displayed. Like in the first series the pH-value and the oxygen concentration drop. The SO_3^{2-} and SO_4^{2-} concentrations rise again, reflecting the absorbed amount of SO_2 . As already mentioned the NO

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removal efficiency was low during the second series, which also led to only minor increases of NO_2^- and NO_3^- . Like before NO_2^- and NO_3^- represent only 10-20% of the absorbed NO.

Experiment	Sample point	pH	O_2	NO_2^-	NO_3^-	SO_3^{2-}	SO_4^{2-}
		[-]	[mg/l]	[mg/l]	[mg/l]	[mg/l]	[mg/l]
02-R00	raw liquid	7.8	7.8	0.048	5.8	1.3	76.7
	used liquid	7.1	0.7	0.078	5.7	25.9	88.8
02-R50	raw liquid	7.8	8.1	0.056	4.9	0.7	42.4
	used liquid	7.3	2.6	0.064	5.1	21.1	69.9
02-R100	raw liquid	7.7	7.9	0.068	2.9	0.2	36.0
	used liquid	7.2	0.7	0.084	3.8	27.3	92.1
02-R150	raw liquid	7.7	8.0	0.083	4.2	1.1	54.4
	used liquid	7.5	3.4	0.090	4.3	7.0	70.2
03-a-T19	raw liquid	7.6	7.6	0.056	2.3	2.0	65.7
	used liquid	7.1	0.3	0.127	4.2	43.5	101.0
03-a-T24	raw liquid	7.7	7.9	0.068	2.9	0.2	36.0
	used liquid	7.3	2.0	0.083	3.5	7.3	80.9
03-a-T30	raw liquid	7.7	7.9	0.068	2.9	0.2	36.0
	used liquid	7.2	3.5	0.103	6.1	18.0	42.6
03-a-T37	raw liquid	7.7	7.9	0.068	2.9	0.2	36.0
	used liquid	7.3	4.6	0.095	3.5	5.6	51.1
03-b-T18	raw liquid	7.6	7.6	0.056	2.3	2.0	65.7
	used liquid	7.2	3.8	0.101	3.3	25.4	76.6
03-b-T30	raw liquid	7.7	7.9	0.068	2.9	0.2	36.0
	used liquid	7.1	3.1	0.093	5.2	20.1	51.8

Table 8.9: Water samples of raw and used scrubbing liquid during the second and third series of test runs

8.3.3 Summary

During the experimental series the influence of the liquid-to-gas ratio, the flue gas temperature at the scrubber inlet, the recirculation ratio and the temperature of the scrubbing liquid have been investigated. It was found that both, higher liquid-to-gas ratios and higher recirculation ratios, have a positive effect on the removal efficiency of all examined pollutants (NO, SO₂, HCl). A reduced flue gas temperature has a positive influence on the removal efficiencies of SO₂ and HCl, but no visible influence on the removal efficiency of NO. Higher scrubbing liquid temperatures have a negative effect on the removal efficiencies of SO₂ and HCl, which is related to the reduced physical solubility. Contrary to this, it was observed, that a higher scrubbing liquid temperature has a positive effect on the NO removal efficiency. This behavior is related to the increased reaction rate of the chemical reactions which occur during the absorption process and will be explained in detail in the next chapter. Table 8.10 shows a summary of the qualitative influence of the varied parameters.

Varied parameter	Removal efficiency		
	NO	SO ₂	HCl
Liquid-to-gas ratio ↑	↑	↑	↑
Recirculation ratio ↑	↑	↑	↑
Flue gas temperature ↓	→	↑	↑
Scrubbing liquid temperature ↑	↑	↓	↓

Table 8.10: Influence of the varied parameters on the removal efficiencies of NO, SO₂ and HCl

9 Investigation of the NO_x absorption mechanism

The absorption mechanisms of SO_2 and HCl are quite clear and the measurements during the experimental series met the expectations. The total amount of dissolved SO_3^{2-} and SO_4^{2-} in the scrubbing solution represents the amount of SO_2 removed from the flue gas, as mentioned in the previous chapter. These observations confirm the estimated mechanisms as described in 7.2.2 and 7.2.5, therefore, no further investigations on the SO_2 and HCl absorption mechanisms have been carried out. Compared to this, the amount of NO_2^- and NO_3^- in the scrubbing liquid did not correspond with the amount of NO removed from the flue gas. This indicates that NO is not only converted to NO_2^- or NO_3^- , but also other nitrogen oxides, which is investigated in this chapter.

As already mentioned a quite big scatter of the NO removal efficiencies was observed during the series of experiments in the pilot plant, while the removal efficiencies of SO_2 and HCl have been stable. The reason for this behavior was found in the continuously changing composition of the scrubbing liquid. Especially the nitrite (NO_2^-) content is a major influencing parameter, while no or only minor influence of the concentrations of the other measured parameters (nitrate, sulfite, sulfate, oxygen and pH-value) was observed. A comparison with the online measurements of ammonium in the waste water plants effluent stream also indicated no connection with the removal efficiency.

9.1 Influence of the nitrite content of the scrubbing solution on the NO_x absorption

Figure 9.1 shows the removal efficiency of NO in connection to the NO_2^- content of the fresh scrubbing solution. It turns out that below a NO_2^- concentration of 0.05 mg/l the removal efficiency drops to zero and that at concentration levels of around 0.1 mg/l removal efficiencies of 0.7 and more were achieved.

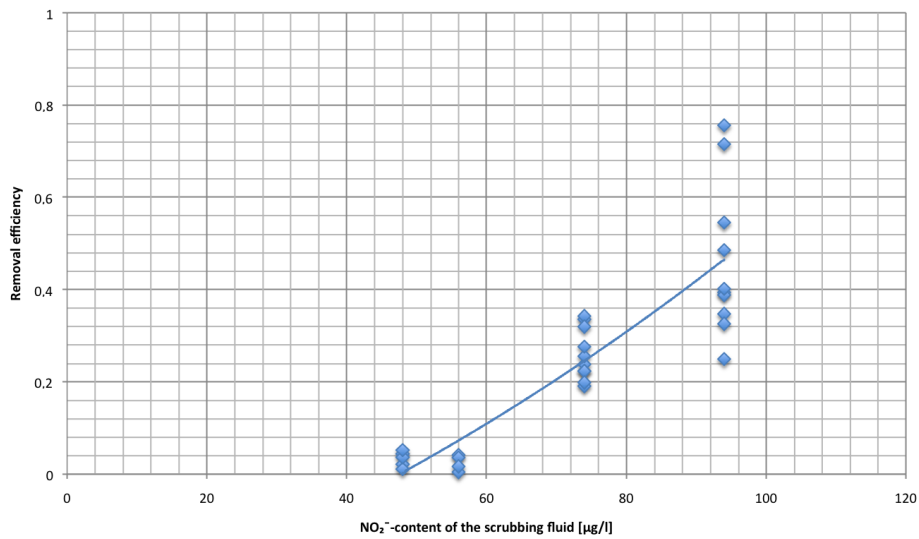


Figure 9.1: Effect of the NO_2^- -content of the scrubbing liquid on the NO removal efficiency

As the reactions (5.1) to (5.9) do not consider an influence of nitrite a literature research based on these findings was carried out and brought up additional eligible reactions of NO_x . Figure 9.2 shows the different oxidation levels of nitrogen and its reaction relations between the compounds. In compounds with oxygen, nitrogen has oxidation levels of +2 to +4. Compared to this, in compounds with hydrogen, nitrogen has oxidation levels of -2 to -3 [44]. In the absorption process only compounds with oxygen occur.

9 Investigation of the NO_x absorption mechanism

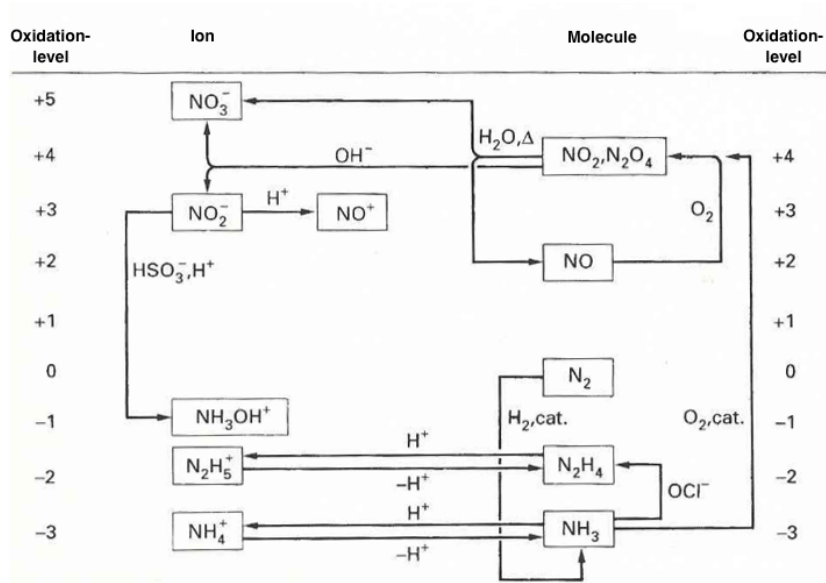


Figure 9.2: The oxidation levels of nitrogen in compounds [44]

Nitrogen monoxide has an odd number of electrons and is, therefore, a radical molecule with different possible electron configurations (Fig. 9.3b). It can easily convert to an "even" molecule by accepting or donating an electron [12, 45, 46].

By the following reaction (9.1) the *nitrosyl cation* (NO^+) is formed, which has a triple bond and is isoelectronic with nitrogen (Fig. 9.3a) [46].



The *nitroxyl anion* (NO^-) is formed from nitrogen monoxide by acceptance of an electron as stated in equation (9.2). This molecule has a double bond and is isoelectronic with oxygen (Fig. 9.3c) [46].



Dimerisation of NO as shown in (9.3) does not happen at ambient temperature in the gas phase, it occurs only in liquid or solid condition. Figure 9.3d and e

9 Investigation of the NO_x absorption mechanism

shows two structural isomers of *dinitrogen dioxide* (N_2O_2), where the latter is only formed in presence of acids like HCl or SO_2 [46].

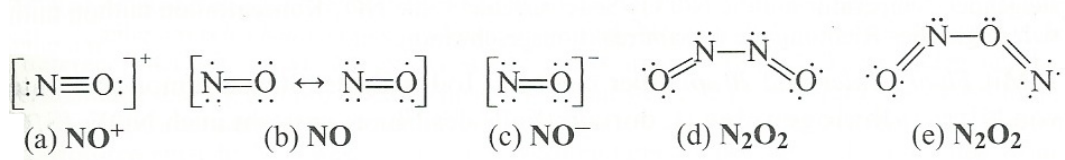


Figure 9.3: The nitrogen monoxide radical and different modified states [46]

Decomposition of nitrogen monoxide to *nitrogen dioxide* (NO_2) and *nitrous oxide* (N_2O) (eq. 9.4) occurs only at high pressures nitrogen and temperatures of more than 50°C [46].



Formation of *dinitrogen trioxide* (N_2O_3) is known from mixtures of NO and NO_2 , as explained in chapter 5.1. N_2O_3 has two isomer structures (Fig. 9.4a), for the solid state an ionogenic structure could also occur (Fig. 9.4b).

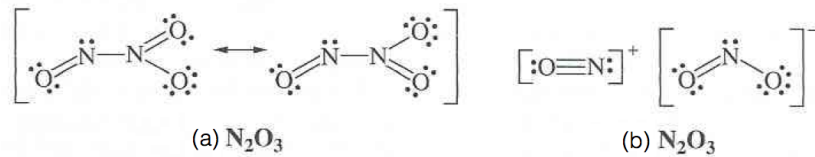


Figure 9.4: Isomers of dinitrogen trioxide [46]

Another possibility to form N_2O_3 is the reaction of the strong Lewis acid NO^+ with the weak Lewis base HNO_2 as stated in equation (9.5) [46].



9 Investigation of the NO_x absorption mechanism

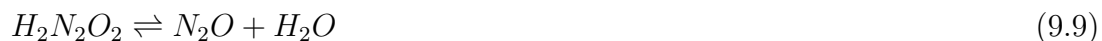
By the following equation (9.6) dinitrogen trioxide forms *nitrite* (NO_2) in an alkaline milieu [12].



The nitrosyl cation also forms *nitrous acid* (HNO_2) in aqueous solution as described in equation (9.7) [12].



Formation of *nitroxyl* (HNO) through protonation of the nitroxyl anion happens quite slow (eq. (9.8)), but the formed acid is very reactive to nucleophiles. It also dimerizes to *hyponitrous acid* ($\text{H}_2\text{N}_2\text{O}_2$), which is not stable and dehydrates to nitrous oxide and water (eq. (9.9)) [46].



Oxidation of the nitrosyl anion forms *peroxynitrite* (OONO^-). In acidic solution the peroxynitrite would further react to peroxynitrous acid which is isomer to nitric acid (eq. (9.10) and (9.11)) [46].



Summary of the mechanism review In conventional gas cleaning facilities NO is usually oxidized to NO_2 before it is removed in a wet scrubber as the solubility of NO_2 is considerably higher than of NO (Tab. 9.1). This conversion of NO to NO_2 takes place at temperatures below 50°C in presence of oxygen. But during the experiments this conversion was not observed, which might be related to the short residence time of the flue gas at temperatures below 50°C .

The observed influence of the nitrite content indicates another absorption mechanism. As nitrite is the dissociated form of nitrous acid, the formation of dinitrogen trioxide as shown in reaction (9.5) seems more probable. The solubility of N_2O_3 is more than two orders of magnitude higher than of NO and close to N_2O_4 (Tab. 9.1).

According to equation (9.6) the nitrite can be newly generated from N_2O_3 if it gets in touch with an alkaline solution. This happens when the used scrubbing solution is mixed with fresh scrubbing solution during recirculation, as the pH-value of the fresh scrubbing solution is approximately 8. This behavior explains the up to 3 times higher removal efficiencies during the experiments with very low nitrite content when recirculation was used (compare Figure 8.9).

Component		Solubility $\left[\frac{\text{mol}}{\text{kg}\cdot\text{bar}}\right]$
Nitrogen monoxide	NO	1.9×10^{-3}
Nitrogen dioxide	NO_2	1.2×10^{-2}
Nitrous oxide	N_2O	2.5×10^{-2}
Dinitrogen trioxide	N_2O_3	0.6
Dinitrogen tetroxide	N_2O_4	1.4

Table 9.1: Solubilities of nitrogen oxides in water [36]

9.2 Additional experiments with focus on the nitrite content of the scrubbing solution - Proof of hypothesis

Based on the findings described in the previous section, experiments with the washing bottle setup described in section 8.1.3 were performed. During these experiments different mixtures of process water from the wastewater plant and distilled water were used to obtain different NO_2^- concentrations. The bottles were aerated for 10 min with flue gas from the incineration plant. This quite long time, compared to the normal contact time of only few seconds in the scrubber, should simulate the recirculation process. Compared to the aeration time of the water, the contact time of the gas phase was only some hundreds of a second. Therefore, the removed part of NO from the gas stream was below the accuracy limit of the gas measuring equipment and not evaluated. As the plant was stopped during the experimental series, also synthetic flue gas from a gas bottle was used to continue with the experiments.

The first bottle was used to remove SO_2 from the gas stream. This causes over the long aeration time a significant drop of the pH-value in the bottle, which does not occur in the pilot plant. The scrubbing liquid in the other three bottles was analyzed. Figure 9.5 shows the NO_2^- concentration in the washing bottles before and after aeration. The contact time and surface are different to the pilot plant, therefore, it is not possible to transfer the results directly to the pilot plant.

9 Investigation of the NO_x absorption mechanism

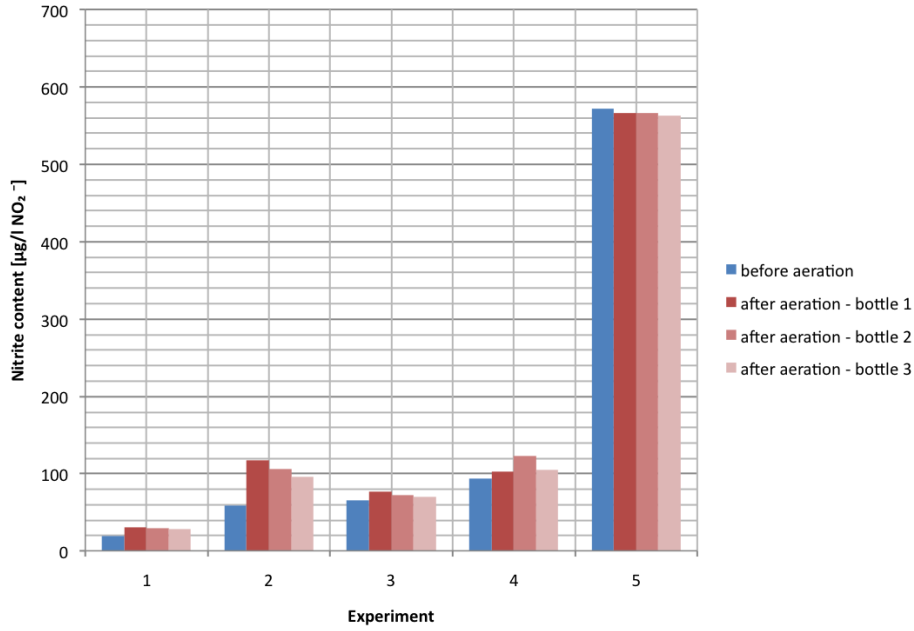


Figure 9.5: NO_2^- concentration before and after aeration

It was observed that the NO_2^- concentration increases during all experiments except the last. The biggest increase was observed during the second experiment. The equilibrium concentration of NO_2^- seems to be around $100 \mu\text{g/l}$. During the first experiment the equilibrium concentration was not reached within the aeration time due to the very low initial concentration of NO_2^- . This is the same behavior as it was observed in the pilot plant. As the concentration in the fifth experiment was much about the equilibrium concentration, the reaction was reversed and the concentration dropped.

The nitrite concentration of almost $600 \mu\text{g/l}$ was the actual concentration in the process water system of the wastewater plant, therefore an additional experimental run with the pilot plant was conducted. Figure 9.1 showed that the NO removal efficiency raised with higher NO_2^- contents of the scrubbing liquid. During the additional run, the NO removal efficiency was around 0.25, which is much lower

9 Investigation of the NO_x absorption mechanism

than the efficiencies reached with nitrite concentrations around $100\text{ }\mu\text{g/l}$. Therefore, it can be assumed, that a maximum in NO removal efficiency would be found between $100\text{ }\mu\text{g/l}$ and $600\text{ }\mu\text{g/l}$ (Fig. 9.6).

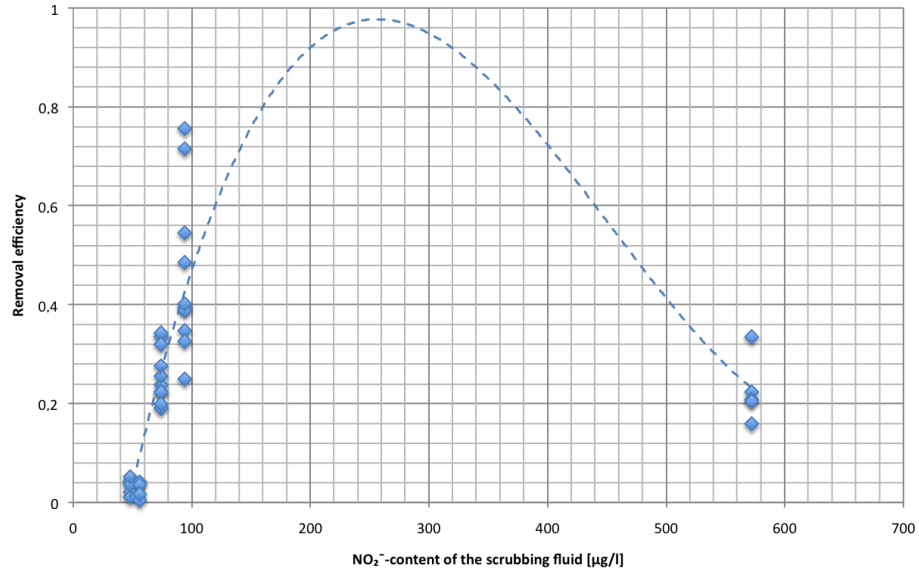


Figure 9.6: NO removal efficiency of the original experiments with nitrite concentrations between 50 and $100\text{ }\mu\text{g/l}$ and the additional experiment with a nitrite concentration of $570\text{ }\mu\text{g/l}$

9.3 Relevance of the findings for the operation of the scrubber system

The relations found during this part of the work have added some new parameters for the operation of the scrubber. As it is very difficult to change the composition of the scrubbing liquid, other methods to deal with the fluctuating nitrite content of the process water and the changing removal efficiency are necessary.

Low concentrations of nitrite could be balanced by recirculation of the scrubbing fluid, as this leads to an accumulation of nitrite in the scrubber as described in section 9.1 (compare also Fig. 8.9). High concentrations of nitrite could be adjusted by mixing the scrubbing solution with drinking water. In the actual scrubber pilot plant only recirculation is possible as there is no drinking water supply at the site of the plant, which could deliver enough water (Fig. 9.7, option A).

Another parameter to increase the removal efficiency is to preheat the scrubbing liquid (Fig. 8.12) as this leads to an increasing reaction rate and, therefore, a higher removal efficiency. In the pilot plant this can be done using the heat exchanger between the recirculation system and the fresh scrubbing liquid supply, but this is limited to the temperature of the used scrubbing solution. This method can be used only in a limited extend, as it has a negative influence to the removal efficiencies of SO_2 and HCl (Fig. 9.7, option B).

Independent from the absorption process itself, but nevertheless very effective is the increase of the liquid-to-gas ratio by increasing the fresh scrubbing liquid flow (Fig. 8.6). The bigger amount of scrubbing liquid can absorb a bigger amount of pollutants, which leads to a higher removal efficiency. This has also a positive effect on the removal efficiencies of the other pollutants (Fig. 9.7, option C).

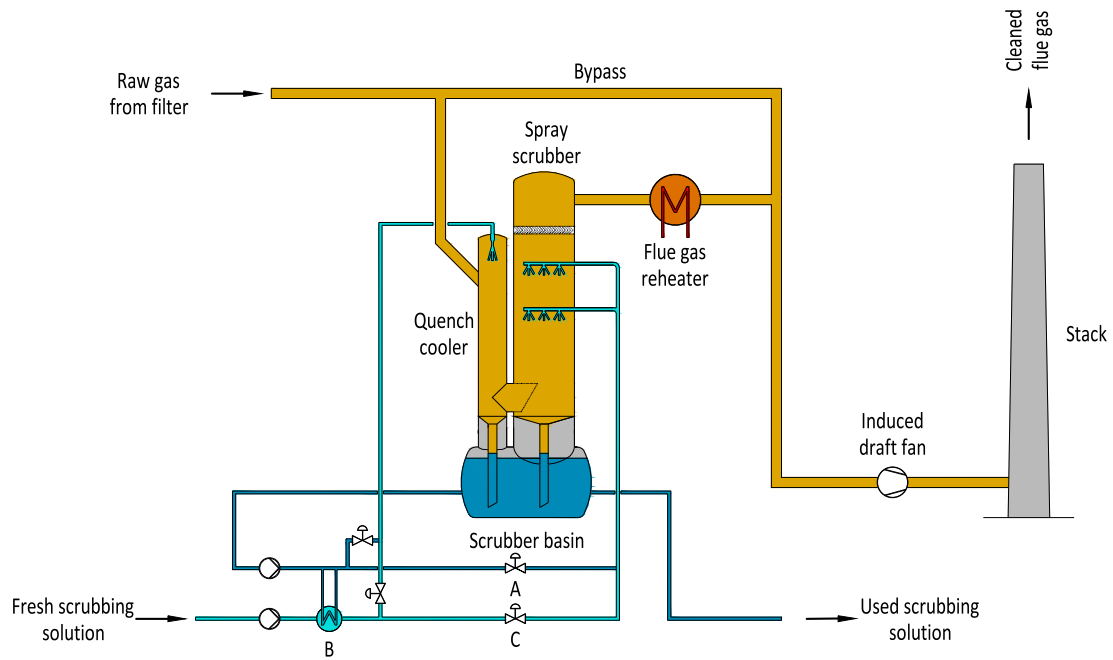


Figure 9.7: Parameters for optimization of the NO removal efficiency in the scrubber system: A - scrubbing liquid recirculation, B - scrubbing liquid preheating, C - fresh scrubbing liquid flow rate

10 Conclusion

A new kind of a scrubber system for treatment of the flue gas from a sewage sludge incineration plant was designed, built, commissioned and tested during this thesis. The developed system uses effluent water from a wastewater treatment plant as absorption fluid in a spray scrubber. Afterwards the liquid is sent back to the bioreactor of the plant for regeneration.

The scrubber system was tested with different liquid-to-gas ratios, different gas and scrubbing liquid temperatures and different scrubbing liquid recirculation ratios. During these experimental series the HCl removal efficiency reached up to 100%, also the SO₂ removal efficiency reached values of up to 95%. NO emissions were reduced by up to 55%, which is comparable to conventional NO_x treatment technologies.

As expected the removal efficiencies for all components grew when the liquid-to-gas ratio was increased. Like the liquid-to-gas ratio, also scrubbing liquid recirculation had also a positive effect on the removal efficiencies, which means that the scrubbing liquid is not saturated with the pollutants.

Reduced flue gas temperatures had a positive effect on the removal efficiency of SO₂ and HCl, as this led to a lower temperature of the scrubbing fluid and, therefore, to an increased physical solubility. With increasing liquid-to-gas ratio this effect was reduced. On the other hand the flue gas temperature had no visible influence on the NO removal efficiency. Similar to the gas temperature lower scrubbing liquid temperatures had also a positive effect on the removal efficiency of SO₂ and HCl, which is related to the increased physical solubility at lower

10 Conclusion

temperatures. Contrary to this, the removal efficiency for NO was remarkable higher when the liquid temperature was increased. This is related to the fact, that the reaction rate of chemical reactions raise at higher temperature, and these dominate the absorption mechanism.

Another finding is, that a nitrite content of the scrubbing liquid within a specific range had a positive effect to the NO absorption rate. This effect will be enforced by scrubbing liquid recirculation, if the nitrite content is low.

Analysis of the used scrubbing liquid had shown that sulfur dioxide is converted to sulfite within the residence time in the scrubber system. Some of the sulfite was further oxidized to sulfate, which might indicate a biological activity in the scrubber basin. Similar to SO₂ also NO led to an increase of the corresponding ions nitrite and nitrate, but not in the same dimension as SO₂.

Investigations of the NO absorption mechanism have shown, that not only formation of nitric and nitrous acid is possible, but also formation of nitrous oxide, dinitrogen trioxide and dinitrogen tetraoxide. The latter are mainly intermediate products in the formation process of the two acids, but they could also remain in this state or react with other intermediate products. Conversion of NO to NO₂ was not observed in the gas phase, which is normally a necessary step to form nitric or nitrous acid. Generally, NO can also react with in water dissolved nitrous acid (or nitrite) and form N₂O₃. This formation of N₂O₃ would explain the influence of dissolved nitrite and the relative high absorption rate without formation of NO₂ in the gas phase.

Based on these observations, the scrubbing liquid composition is of special importance to gain high NO elimination rates, while this is not critical for the SO₂ and HCl elimination.

From a general point of view the installation is suitable for the intended purpose and that the emissions from the incineration plant can be controlled according to the legal regulations. The used technology is quite simple and should be transferable to other installation sites, as long as the effluent composition (especially

10 Conclusion

the nitrite content) of the wastewater plant is kept in mind. As effluent water is cheap and plentiful available in a wastewater treatment plant, this technology is an interesting choice for incineration plants at locations of wastewater treatment plants. If only SO_2 and HCl elimination are important, the water consumption can be reduced by extensive scrubbing fluid recirculation.

From the scientific view further investigations of the NO absorption mechanism would be desirable. Especially the intermediate steps of the reaction chain and possible biological effects are of major interest and should be topics for future research projects.

Abbreviations

DSS	Dried sewage sludge
EDTA	Ethylenediaminetetraacetic acid
FID	Flame ionization detector
FTIR	Fourier transform infrared spectroscopy
NDIR	Nondispersive infrared sensor
n.a.	not available
PCDD	polychlorinated dibenzo-p-dioxine
PCDF	polychlorinated dibenzo-p-furan
PE	Population equivalent
PLC	Programmable Logic Controller
SCADA	Supervisory control and data acquisition
SCR	Selective catalytic reduction
SNCR	Selective non-catalytic reduction
TEG-DME	Tetraethylenglykoldimethylether
THC	Total hydrocarbons
TIBP	Tributylphosphate
TIEG-DME	Triethylenglykoldimethylether
WWTP	Wastewater treatment plant

Symbols

c_{CO_2}	Volume fraction of carbon dioxide	[-]
c_{N_2}	Volume fraction of nitrogen	[-]
c_{O_2}	Volume fraction of oxygen	[-]
c_{water}	Volume fraction of water vapor	[-]
$c_{water,sat}$	Saturation volume fraction of water vapor	[-]
c_x	Concentration of component x in the flue gas	[mg/Nm ³]
$c_{x,l}$	Concentration of component x in the scrubbing liquid	[mol/kg]
\dot{G}	Gas flow	[kmol/h]
$h_{sol,x}$	Heat of solution of component x	[kJ/kg]
h_g	Specific enthalpy of the flue gas	[kJ/kg]
h_l	Specific enthalpy of the scrubbing liquid	[kJ/kg]
h_v	Specific enthalpy of water vapor	[kJ/kg]
h_{CO_2}	Specific enthalpy of carbon dioxide	[kJ/kg]
h_{N_2}	Specific enthalpy of nitrogen	[kJ/kg]
h_{O_2}	Specific enthalpy of oxygen	[kJ/kg]
$k_{H,x}$	Henry coefficient of component x	[mol/(kg·bar)]
LGR	Liquid-to-gas ratio	[-]
\dot{L}	Liquid flow	[kmol/h]
\dot{L}_f	Flow of fresh scrubbing liquid	[kmol/h]
\dot{L}_r	Flow of recirculated scrubbing liquid	[kmol/h]
\dot{m}	Mass flow	[kg/h]
\dot{m}_g	Flue gas mass flow	[kg/h]

Symbols

\dot{m}_l	Scrubbing liquid mass flow	[kg/h]
\dot{m}_v	Water vapor mass flow	[kg/h]
$\dot{m}_{v,con}$	Mass flow of condensing water vapor	[kg/h]
$\dot{m}_{x,t}$	Mass flow of component x, transferred to the scrubbing liquid	[kg/h]
$\dot{n}_{x,t}$	Molar flow of component x, transferred to the scrubbing liquid	[mol/h]
p_{abs}	Absolute pressure	[bar]
p_{sat}	Saturation partial pressure of water vapor	[bar]
p_x	Partial pressure of component x	[bar]
R	Recirculation ratio	[-]
$\dot{V}_{g,n}$	Flue gas volume flow at normal conditions	[Nm ³ /h]
$\dot{V}_{v,n}$	Water vapor volume flow at normal conditions	[Nm ³ /h]
η_{rem}	Removal efficiency	[-]
$\rho_{g,n}$	Density of the flue gas at normal conditions	[kg/m ³]
ρ_n	Density at normal conditions	[kg/m ³]
$\rho_{CO_2,n}$	Density of carobon dioxide at normal conditions	[kg/m ³]
$\rho_{N_2,n}$	Density of nitrogen at normal conditions	[kg/m ³]
$\rho_{O_2,n}$	Density of oxygen at normal conditions	[kg/m ³]

Additionally used indices

<i>dry</i>	Dry conditions
<i>wet</i>	Wet conditions
<i>n</i>	Normal conditions (273.15 K, 1.01325 bar)
<i>in</i>	Conditions at inlet
<i>out</i>	Conditions at outlet
<i>raw</i>	Conditions in raw flue gas
<i>clean</i>	Conditions in clean flue gas

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Appendix A. Calculation of the design flue gas composition

The flue gas composition, which was used for the design of the quench column and the scrubber column, was calculated based on the sewage sludge composition (Tab.8.4), combined with data gained from the operation of the sewage sludge incineration plant Bad Vöslau.

The flue gas composition was calculated according to equation (A.1) to (A.10), the air / fuel equivalence ratio was 1.8.

$$V_{O_2,min} = 1,867 \cdot c_C + 5.603 \cdot c_H + 0.7 \cdot c_S - 0.7 \cdot c_O - 0.158 \cdot c_{Cl} \quad (A.1)$$

$$V_{air,min} = V_{O_2,min}/0.21 \quad (A.2)$$

$$V_{CO_2} = 1.867 \cdot c_C \quad (A.3)$$

$$V_{SO_2} = 0.7 \cdot c_S \quad (A.4)$$

$$V_{HCl} = 0.631 \cdot c_{Cl} \quad (A.5)$$

$$V_{H_2O} = 11.21 \cdot c_H - 0.316 \cdot c_{Cl} + 1.24 \cdot c_W \quad (A.6)$$

$$V_{O_2} = 0.21 \cdot (\lambda - 1) \cdot V_{air,min} \quad (A.7)$$

$$V_{N_2} = 0.79 \cdot \lambda \cdot V_{air,min} \quad (A.8)$$

$$V_{g,dry} = V_{CO_2} + V_{SO_2} + V_{HCl} + V_{O_2} + V_{N_2} \quad (A.9)$$

$$V_{g,wet} = V_{CO_2} + V_{SO_2} + V_{HCl} + V_{H_2O} + V_{O_2} + V_{N_2} \quad (A.10)$$

Division of the volume of the particular component through the volume of the wet or dry flue gas delivers the wet respectively dry volumetric concentration of the component. The calculated concentrations of carbon dioxide, nitrogen, oxygen and water are shown in Table 7.4 and 7.7, the concentrations of the other components listed in Table 7.7 have been estimated based on the operational experience.

V_X	Volume of component X per kg fuel	[m ³ /kg]
c_X	Mass fraction of component X in the fuel	[-]
λ	air / fuel equivalence ratio	[-]

Appendix B. Result sheets of the experimental runs

19.06.2013		Test run 01-A-F10		Page 1 of 2
Operating conditions of the incineration plant:				
Primary fuel:	sewage sludge	feed rate:	589 kg/h	
Secondary fuel:	screenings	feed rate:	0 kg/h	
Bed material:	silica sand	amount:	2000 kg	
Start Test Run:	15:00	Stable:	16:00	End Test Run: 18:00
Measurements flue gas:				
		Begin stable:	End stable:	
Flue gas volume flow	FI0A	7810	7790	Nm³/h
Fluidized bed temp.	TC2	730	731	°C
Free board temp.	TC3.1	847	843	°C
Afterburner chamber temp.	TC4.1	922	923	°C
Outlet TO-Heater temp.	TC7	278	276	°C
Inlet filter temp.	TC9	159	160	°C
Inlet quench cooler temp.	TI5805	141	142	°C
Outlet quench cooler temp.	TI5810	56.4	56.9	°C
Outlet spray scrubber temp.	TI5803	48.7	48.9	°C
Stack temp.	TC0A	66.5	66.5	°C
Measurements scrubbing liquid:				
		Begin stable:	End stable:	
Fresh liquid flow quench	FIS5803	0.0	0.0	m³/h
Recirculation liquid flow quench	by calc.	20.8	20.7	m³/h
Total liquid flow quench	FIS5806	20.8	20.7	m³/h
Mix liquid temp. quench	TI5804	51.0	51.4	°C
Fresh liquid flow spray scrubber	FI5801	10.1	10.1	m³/h
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m³/h
Total liquid flow spray scrubber	FI5802	10.1	10.1	m³/h
Liquid to gas ratio	by calc.	1.61	1.61	kmol/kmol
Recirculation ratio	by calc.	0.0	0.0	-
Mix liquid temp. spray scrubber	TI5802	19.4	19.5	°C
Fresh liquid temperature	TI5801	18.5	18.7	°C
Used liquid temperature	TI5808	51.0	51.1	°C

Flue gas analysis - Raw gas:

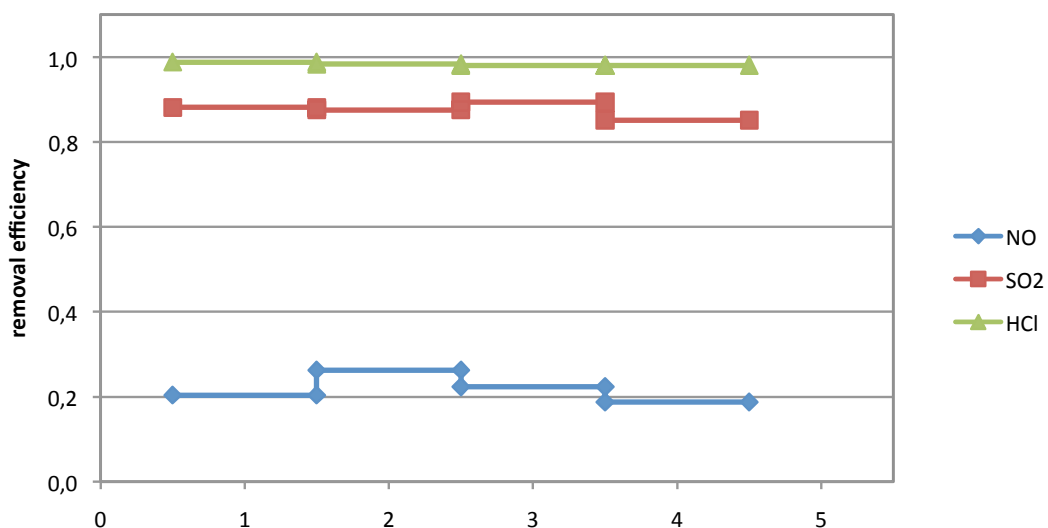
	HMW 1	HMW 2	HMW 3	HMW 4	
O ₂	15.4%	14.5%	14.4%	14.5%	vol%
H ₂ O	18.9%	19.2%	19.3%	19.5%	vol%
NO	33.9	33.9	32.2	32.0	mg/Nm ³ (dry)
NO ₂	2.9	0.7	0.4	0.0	mg/Nm ³ (dry)
SO ₂	72.6	67.5	72.5	51.7	mg/Nm ³ (dry)
HCl	4.1	3.3	3	2.7	mg/Nm ³ (dry)
CO	4.6	4.4	4.4	4.2	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:

	HMW 1	HMW 2	HMW 3	HMW 4	
O ₂	15.5%	15.4%	15.4%	15.4%	vol%
H ₂ O	12.0%	12.1%	12.1%	12.1%	vol%
NO	27.0	25.0	25.0	26.0	mg/Nm ³ (dry)
NO ₂	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
SO ₂	8.6	8.4	7.7	7.7	mg/Nm ³ (dry)
HCl	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
CO	3.5	3.6	3.1	3.4	mg/Nm ³ (dry)

Removal efficiency:

	HMW 1	HMW 2	HMW 3	HMW 4	AV
NO	0.204	0.263	0.224	0.188	0.220
SO ₂	0.882	0.876	0.894	0.851	0.877
HCl	0.988	0.984	0.980	0.981	0.984



19.06.2013		Test run 01-A-F15		Page 1 of 2	
Operating conditions of the incineration plant:					
Primary fuel:	sewage sludge	feed rate:	574 kg/h		
Secondary fuel:	screenings	feed rate:	0 kg/h		
Bed material:	silica sand	amount:	2000 kg		
Start Test Run: 12:10		Stable:	12:35	End Test Run: 15:00	
Measurements flue gas:					
		Begin stable:	End stable:		
Flue gas volume flow	FI0A	7810	7830	Nm³/h	
Fluidized bed temp.	TC2	729	727	°C	
Free board temp.	TC3.1	845	842	°C	
Afterburner chamber temp.	TC4.1	915	915	°C	
Outlet TO-Heater temp.	TC7	276	278	°C	
Inlet filter temp.	TC9	154	162	°C	
Inlet quench cooler temp.	TI5805	137	141	°C	
Outlet quench cooler temp.	TI5810	53.7	54	°C	
Outlet spray scrubber temp.	TI5803	39.2	39.6	°C	
Stack temp.	TC0A	58.5	59.7	°C	
Measurements scrubbing liquid:					
		Begin stable:	End stable:		
Fresh liquid flow quench	FIS5803	0.0	0.0	m³/h	
Recirculation liquid flow quench	by calc.	20.2	20.7	m³/h	
Total liquid flow quench	FIS5806	20.2	20.7	m³/h	
Mix liquid temp. quench	TI5804	45.7	46.8	°C	
Fresh liquid flow spray scrubber	FI5801	14.8	14.9	m³/h	
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m³/h	
Total liquid flow spray scrubber	FI5802	14.8	14.9	m³/h	
Liquid to gas ratio	by calc.	2.36	2.37	kmol/kmol	
Recirculation ratio	by calc.	0.0	0.0	-	
Mix liquid temp. spray scrubber	TI5802	18.8	19.0	°C	
Fresh liquid temperature	TI5801	18.1	18.4	°C	
Used liquid temperature	TI5808	45.8	47.1	°C	

Flue gas analysis - Raw gas:

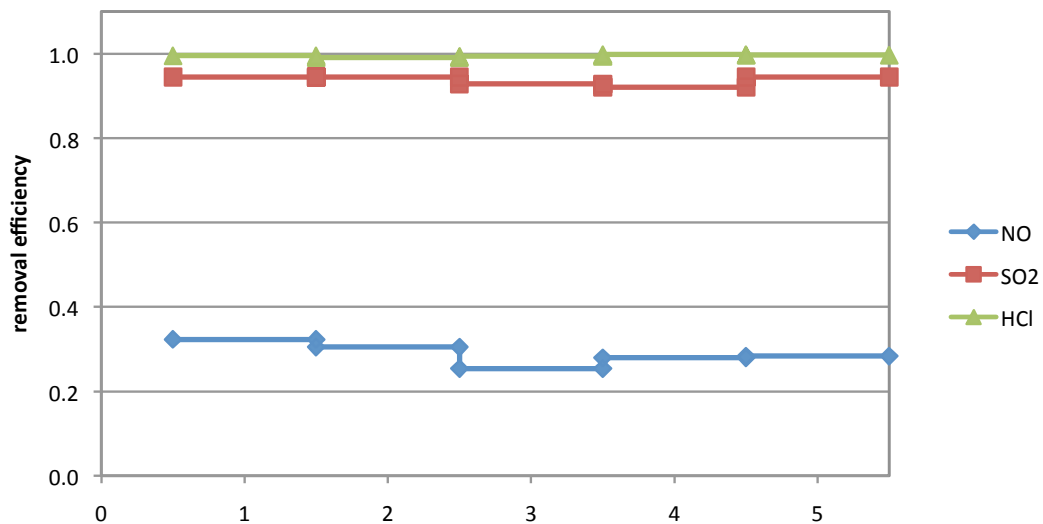
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.5%	14.5%	14.3%	14.4%	14.5%	vol%
H ₂ O	19.2%	19.6%	20.1%	19.7%	19.4%	vol%
NO	32.5	33.1	33.5	33.3	32.1	mg/Nm ³ (dry)
NO ₂	0.1	0.0	0.3	0.2	0.0	mg/Nm ³ (dry)
SO ₂	99.4	97.7	195.3	185.1	101.1	mg/Nm ³ (dry)
HCl	15.1	7.4	9.2	36.1	16.3	mg/Nm ³ (dry)
CO	4.2	4.4	4.5	4.5	4.4	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	15.5%	15.5%	15.3%	15.3%	15.5%	vol%
H ₂ O	8.5%	8.7%	8.6%	8.5%	8.7%	vol%
NO	22.0	23.0	25.0	24.0	23.0	mg/Nm ³ (dry)
NO ₂	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
SO ₂	5.5	5.4	14	14.7	5.6	mg/Nm ³ (dry)
HCl	0.07	0.067	0.056	0.065	0.058	mg/Nm ³ (dry)
CO	1.3	2.1	2.4	1.8	1.9	mg/Nm ³ (dry)

Removal efficiency:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.323	0.305	0.254	0.279	0.283	0.289
SO ₂	0.945	0.945	0.928	0.921	0.945	0.933
HCl	0.995	0.991	0.994	0.998	0.996	0.996



19.06.2013		Test run 01-A-F20		Page 1 of 2	
Operating conditions of the incineration plant:					
Primary fuel:	sewage sludge	feed rate:	577 kg/h		
Secondary fuel:	screenings	feed rate:	0 kg/h		
Bed material:	silica sand	amount:	2000 kg		
Start Test Run: 08:30		Stable:	09:30	End Test Run: 12:00	
Measurements flue gas:					
		Begin stable:	End stable:		
Flue gas volume flow	FI0A	7780	7810	Nm³/h	
Fluidized bed temp.	TC2	727	730	°C	
Free board temp.	TC3.1	835	843	°C	
Afterburner chamber temp.	TC4.1	907	915	°C	
Outlet TO-Heater temp.	TC7	278	275	°C	
Inlet filter temp.	TC9	145	157	°C	
Inlet quench cooler temp.	TI5805	134	141	°C	
Outlet quench cooler temp.	TI5810	52	51.1	°C	
Outlet spray scrubber temp.	TI5803	32	31.6	°C	
Stack temp.	TC0A	53.3	53.6	°C	
Measurements scrubbing liquid:					
		Begin stable:	End stable:		
Fresh liquid flow quench	FIS5803	0.0	0.0	m³/h	
Recirculation liquid flow quench	by calc.	20.2	20.2	m³/h	
Total liquid flow quench	FIS5806	20.2	20.2	m³/h	
Mix liquid temp. quench	TI5804	42.5	41.5	°C	
Fresh liquid flow spray scrubber	FI5801	20.0	20.0	m³/h	
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m³/h	
Total liquid flow spray scrubber	FI5802	20.0	20.0	m³/h	
Liquid to gas ratio	by calc.	3.20	3.19	kmol/kmol	
Recirculation ratio	by calc.	0.0	0.0	-	
Mix liquid temp. spray scrubber	TI5802	18.4	18.6	°C	
Fresh liquid temperature	TI5801	17.9	18.0	°C	
Used liquid temperature	TI5808	42.8	41.6	°C	

Flue gas analysis - Raw gas:

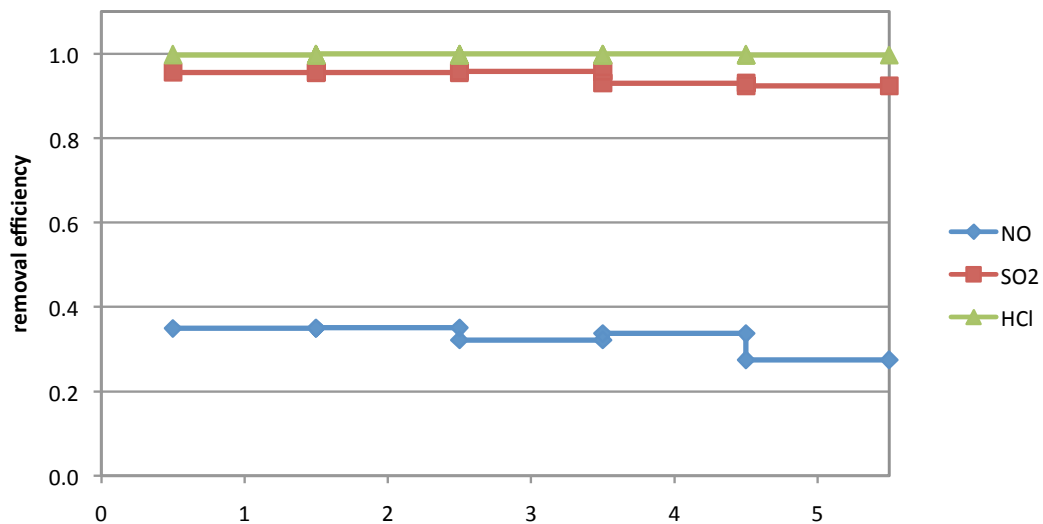
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.2%	14.4%	14.3%	14.5%	14.5%	vol%
H ₂ O	19.8%	19.7%	19.9%	19.1%	19.5%	vol%
NO	33.8	35.4	33.9	34.7	31.7	mg/Nm ³ (dry)
NO ₂	0.8	0.9	0.7	0.5	0.2	mg/Nm ³ (dry)
SO ₂	236.4	232.9	263.9	191.8	141.7	mg/Nm ³ (dry)
HCl	44.3	74.8	97.5	77.5	23.7	mg/Nm ³ (dry)
CO	4	4.2	4.2	4.6	4.3	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	15.2%	15.3%	15.2%	15.4%	15.4%	vol%
H ₂ O	6.1%	6.1%	6.1%	6.1%	6.1%	vol%
NO	22.0	23.0	23.0	23.0	23.0	mg/Nm ³ (dry)
NO ₂	0.0	0.0	0.0	0.0	0.0	mg/Nm ³ (dry)
SO ₂	10.4	10.4	11	13.4	10.8	mg/Nm ³ (dry)
HCl	0.066	0.071	0.07	0.074	0.074	mg/Nm ³ (dry)
CO	1.1	1.1	1.1	1.9	1	mg/Nm ³ (dry)

Removal efficiency:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.349	0.350	0.322	0.337	0.274	0.327
SO ₂	0.956	0.955	0.958	0.930	0.924	0.948
HCl	0.999	0.999	0.999	0.999	0.997	0.999

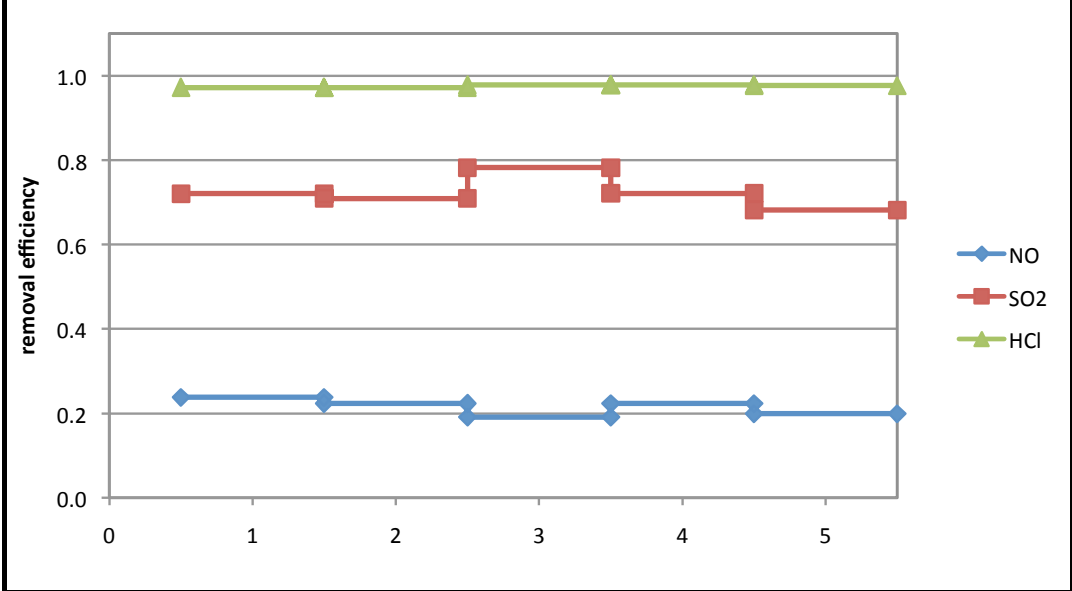


18.06.2013		Test run 01-B-F10		Page 1 of 2	
Operating conditions of the incineration plant:					
Primary fuel:	sewage sludge	feed rate:	595 kg/h		
Secondary fuel:	screenings	feed rate:	0 kg/h		
Bed material:	silica sand	amount:	2000 kg		
Start Test Run: 11:30		Stable:	12:00	End Test Run: 14:30	
Measurements flue gas:					
		Begin stable:	End stable:		
Flue gas volume flow	FI0A	7790	7750	Nm³/h	
Fluidized bed temp.	TC2	728	729	°C	
Free board temp.	TC3.1	832	852	°C	
Afterburner chamber temp.	TC4.1	907	923	°C	
Outlet TO-Heater temp.	TC7	278	277	°C	
Inlet filter temp.	TC9	176	181	°C	
Inlet quench cooler temp.	TI5805	147	150	°C	
Outlet quench cooler temp.	TI5810	60.4	60.4	°C	
Outlet spray scrubber temp.	TI5803	53.8	54.6	°C	
Stack temp.	TC0A	71	72	°C	
Measurements scrubbing liquid:					
		Begin stable:	End stable:		
Fresh liquid flow quench	FIS5803	0.0	0.0	m³/h	
Recirculation liquid flow quench	by calc.	4.5	4.5	m³/h	
Total liquid flow quench	FIS5806	4.5	4.5	m³/h	
Mix liquid temp. quench	TI5804	45.5	45.2	°C	
Fresh liquid flow spray scrubber	FI5801	10.1	10.1	m³/h	
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m³/h	
Total liquid flow spray scrubber	FI5802	10.1	10.1	m³/h	
Liquid to gas ratio	by calc.	1.61	1.62	kmol/kmol	
Recirculation ratio	by calc.	0.0	0.0	-	
Mix liquid temp. spray scrubber	TI5802	18.4	18.6	°C	
Fresh liquid temperature	TI5801	17.6	17.9	°C	
Used liquid temperature	TI5808	45.4	44.9	°C	

Flue gas analysis - Raw gas:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.6%	14.7%	14.5%	14.6%	14.6%	vol%
H ₂ O	18.5%	18.2%	18.7%	18.4%	18.4%	vol%
NO	31.5	32.2	30.9	30.9	31.2	mg/Nm ³ (dry)
NO ₂	0.1	0.0	0.0	0.0	0.0	mg/Nm ³ (dry)
SO ₂	51.5	46.1	73.4	50.2	39.6	mg/Nm ³ (dry)
HCl	2.5	2.4	2.7	2.8	2.8	mg/Nm ³ (dry)
CO	4.7	4.8	5	5	5	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	15.5%	15.6%	15.4%	15.5%	15.6%	vol%
H ₂ O	12.5%	12.5%	12.6%	12.5%	12.5%	vol%
NO	24.0	25.0	25.0	24.0	25.0	mg/Nm ³ (dry)
NO ₂	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
SO ₂	14.4	13.4	16	14	12.6	mg/Nm ³ (dry)
HCl	0.065	0.068	0.059	0.06	0.066	mg/Nm ³ (dry)
CO	4.3	4.5	5	4.9	4.7	mg/Nm ³ (dry)

Removal efficiency:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.238	0.224	0.191	0.223	0.199	0.215
SO ₂	0.720	0.709	0.782	0.721	0.682	0.730
HCl	0.974	0.972	0.978	0.979	0.976	0.976



18.06.2013		Test run 01-B-F15		Page 1 of 2	
Operating conditions of the incineration plant:					
Primary fuel:	sewage sludge	feed rate:	611 kg/h		
Secondary fuel:	screenings	feed rate:	0 kg/h		
Bed material:	silica sand	amount:	2000 kg		
Start Test Run: 08:30		Stable:	09:00	End Test Run: 11:30	
Measurements flue gas:					
		Begin stable:	End stable:		
Flue gas volume flow	FI0A	7820	7810	Nm³/h	
Fluidized bed temp.	TC2	728	732	°C	
Free board temp.	TC3.1	826	851	°C	
Afterburner chamber temp.	TC4.1	907	921	°C	
Outlet TO-Heater temp.	TC7	277	278	°C	
Inlet filter temp.	TC9	181	175	°C	
Inlet quench cooler temp.	TI5805	152	146	°C	
Outlet quench cooler temp.	TI5810	60.3	60.9	°C	
Outlet spray scrubber temp.	TI5803	43	44.6	°C	
Stack temp.	TC0A	61	62	°C	
Measurements scrubbing liquid:					
		Begin stable:	End stable:		
Fresh liquid flow quench	FIS5803	0.0	0.0	m³/h	
Recirculation liquid flow quench	by calc.	4.1	4.3	m³/h	
Total liquid flow quench	FIS5806	4.1	4.3	m³/h	
Mix liquid temp. quench	TI5804	42.7	43.2	°C	
Fresh liquid flow spray scrubber	FI5801	14.7	14.8	m³/h	
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m³/h	
Total liquid flow spray scrubber	FI5802	14.7	14.8	m³/h	
Liquid to gas ratio	by calc.	2.34	2.36	kmol/kmol	
Recirculation ratio	by calc.	0.0	0.0	-	
Mix liquid temp. spray scrubber	TI5802	17.8	18.1	°C	
Fresh liquid temperature	TI5801	17.4	17.5	°C	
Used liquid temperature	TI5808	42.9	42.7	°C	

Flue gas analysis - Raw gas:

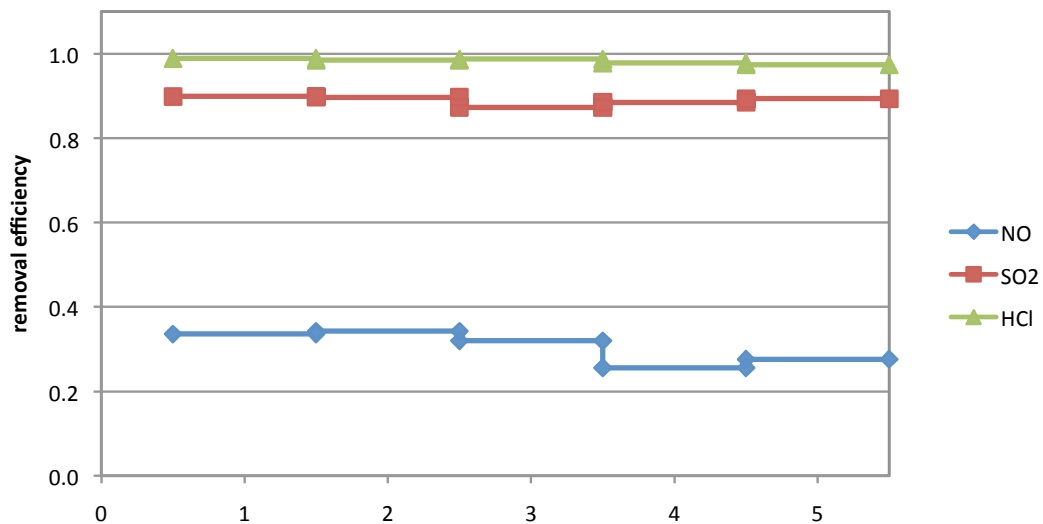
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.5%	14.4%	14.5%	14.6%	14.6%	vol%
H ₂ O	18.3%	18.7%	18.7%	18.4%	18.6%	vol%
NO	30.1	28.9	29.4	30.9	30.4	mg/Nm ³ (dry)
NO ₂	0.6	0.2	0.3	0.3	0.3	mg/Nm ³ (dry)
SO ₂	80.9	98.7	110.8	95.5	77.8	mg/Nm ³ (dry)
HCl	6	4.7	5	2.8	2.6	mg/Nm ³ (dry)
CO	5	5.2	5	4.7	4.6	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	15.4%	15.4%	15.4%	15.5%	15.5%	vol%
H ₂ O	9.6%	9.6%	9.8%	9.7%	9.6%	vol%
NO	20.0	19.0	20.0	23.0	22.0	mg/Nm ³ (dry)
NO ₂	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
SO ₂	8.2	10.2	14.1	11	8.3	mg/Nm ³ (dry)
HCl	0.066	0.071	0.065	0.061	0.068	mg/Nm ³ (dry)
CO	5	4.9	4.8	4.4	4.1	mg/Nm ³ (dry)

Removal efficiency:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.336	0.343	0.320	0.256	0.276	0.305
SO ₂	0.899	0.897	0.873	0.885	0.893	0.888
HCl	0.989	0.985	0.987	0.978	0.974	0.984

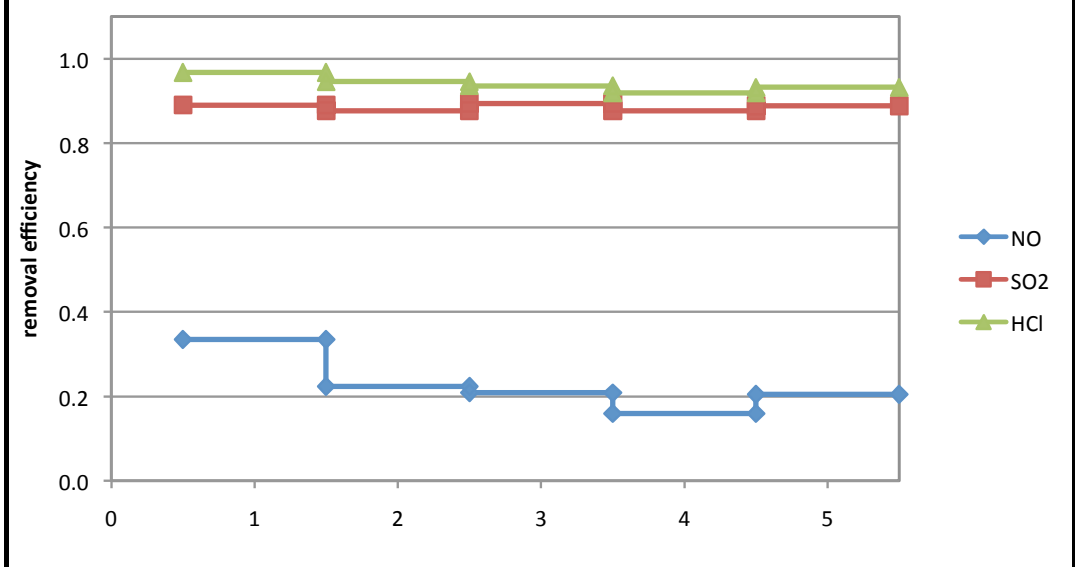


03.12.2013		Test run 01-B-F16		Page 1 of 2	
Operating conditions of the incineration plant:					
Primary fuel:	sewage sludge	feed rate:	787 kg/h		
Secondary fuel:	screenings	feed rate:	0 kg/h		
Bed material:	silica sand	amount:	2000 kg		
Start Test Run: 08:00		Stable:	08:30	End Test Run: 12:00	
Measurements flue gas:					
		Begin stable:	End stable:		
Flue gas volume flow	FI0A	8020	7910	Nm³/h	
Fluidized bed temp.	TC2	726	727	°C	
Free board temp.	TC3.1	790	803	°C	
Afterburner chamber temp.	TC4.1	886	901	°C	
Outlet TO-Heater temp.	TC7	297	298	°C	
Inlet filter temp.	TC9	165	153	°C	
Inlet quench cooler temp.	TI5805	144	139	°C	
Outlet quench cooler temp.	TI5810	60.1	61.8	°C	
Outlet spray scrubber temp.	TI5803	40.5	41.5	°C	
Stack temp.	TC0A	64	67	°C	
Measurements scrubbing liquid:					
		Begin stable:	End stable:		
Fresh liquid flow quench	FIS5803	0.0	0.0	m³/h	
Recirculation liquid flow quench	by calc.	5.9	5.9	m³/h	
Total liquid flow quench	FIS5806	5.9	5.9	m³/h	
Mix liquid temp. quench	TI5804	42.4	42.4	°C	
Fresh liquid flow spray scrubber	FI5801	15.3	15.2	m³/h	
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m³/h	
Total liquid flow spray scrubber	FI5802	15.3	15.2	m³/h	
Liquid to gas ratio	by calc.	2.38	2.39	kmol/kmol	
Recirculation ratio	by calc.	0.0	0.0	-	
Mix liquid temp. spray scrubber	TI5802	12.9	13.0	°C	
Fresh liquid temperature	TI5801	12.5	12.6	°C	
Used liquid temperature	TI5808	41.6	42.2	°C	

Flue gas analysis - Raw gas:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	15.0	13.6	13.6	13.7	13.5	vol%
H ₂ O	21.4	19.7	19.9	21.3	20.9	vol%
NO	48.1	42.5	51.8	45.2	41.5	mg/Nm ³ (dry)
NO ₂	13.1	8.5	8.9	9.0	9.1	mg/Nm ³ (dry)
SO ₂	91.9	90.7	128.5	81	87.8	mg/Nm ³ (dry)
HCl	2.5	1.4	0.9	1	1.3	mg/Nm ³ (dry)
CO	3.9	4	3.5	3.9	4.1	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.5	14.3	14.3	14.4	14.2	vol%
H ₂ O	8.5	8.9	9.6	9.6	9.3	vol%
NO	32.0	33.0	41.0	38.0	33.0	mg/Nm ³ (dry)
NO ₂	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
SO ₂	10.1	11.2	13.6	10	9.8	mg/Nm ³ (dry)
HCl	0.081	0.076	0.058	0.081	0.088	mg/Nm ³ (dry)
CO	0	0.1	0.2	0.4	0.2	mg/Nm ³ (dry)

Removal efficiency:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.335	0.224	0.208	0.159	0.205	0.227
SO ₂	0.890	0.877	0.894	0.877	0.888	0.886
HCl	0.968	0.946	0.936	0.919	0.932	0.946



Operating conditions of the incineration plant:

Primary fuel: sewage sludge	feed rate:	760 kg/h
Secondary fuel: screenings	feed rate:	0 kg/h
Bed material: silica sand	amount:	2000 kg

Start Test Run: 17:00	Stable: 17:30	End Test Run: 20:00
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Measurements flue gas:

Begin stable: End stable:

Flue gas volume flow	FI0A	8130	8140	Nm ³ /h
Fluidized bed temp.	TC2	730	733	°C
Free board temp.	TC3.1	784	805	°C
Afterburner chamber temp.	TC4.1	845	883	°C
Outlet TO-Heater temp.	TC7	299	305	°C
Inlet filter temp.	TC9	154	162	°C
Inlet quench cooler temp.	TI5805	134	138	°C
Outlet quench cooler temp.	TI5810	59.2	58.8	°C
Outlet spray scrubber temp.	TI5803	35.5	34.9	°C
Stack temp.	TC0A	58	58	°C

Measurements scrubbing liquid:

Begin stable: End stable:

Fresh liquid flow quench	FIS5803	0.0	0.0	m ³ /h
Recirculation liquid flow quench	by calc.	6.0	6.0	m ³ /h
Total liquid flow quench	FIS5806	6.0	6.0	m ³ /h
Mix liquid temp. quench	TI5804	39.5	39.5	°C
Fresh liquid flow spray scrubber	FI5801	18.6	18.3	m ³ /h
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m ³ /h
Total liquid flow spray scrubber	FI5802	18.6	18.3	m ³ /h
Liquid to gas ratio	by calc.	2.85	2.80	kmol/kmol
Recirculation ratio	by calc.	0.0	0.0	-
Mix liquid temp. spray scrubber	TI5802	13.0	13.0	°C
Fresh liquid temperature	TI5801	12.6	12.6	°C
Used liquid temperature	TI5808	39.9	38.9	°C

Flue gas analysis - Raw gas:

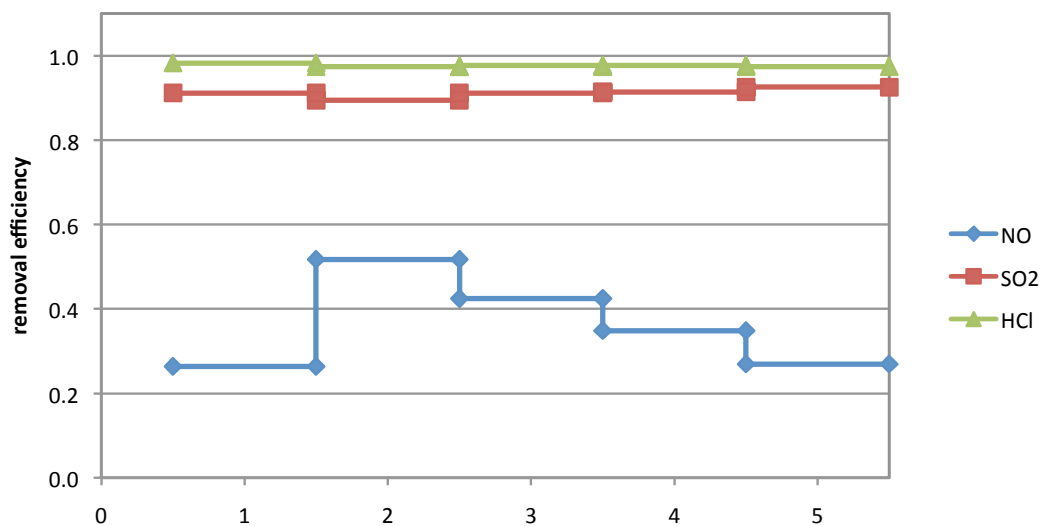
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	13.9	14.1	13.9	13.8	14.0	vol%
H ₂ O	18.5	18.2	18.7	18.9	18.6	vol%
NO	42.1	31.1	36.5	39.9	39.7	mg/Nm ³ (dry)
NO ₂	6.0	5.3	6.0	5.7	5.7	mg/Nm ³ (dry)
SO ₂	127.8	95.7	123.4	118.8	145.4	mg/Nm ³ (dry)
HCl	5.6	5.4	4.9	4.2	4	mg/Nm ³ (dry)
CO	4.9	4.8	4.5	4.2	3.8	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.6	14.9	14.7	14.6	14.7	vol%
H ₂ O	6.6	6.7	7.0	7.1	7.0	vol%
NO	31.0	15.0	21.0	26.0	29.0	mg/Nm ³ (dry)
NO ₂	0.1	0.0	0.0	0.1	0.0	mg/Nm ³ (dry)
SO ₂	11.3	10.1	10.9	10.2	10.8	mg/Nm ³ (dry)
HCl	0.096	0.14	0.115	0.095	0.098	mg/Nm ³ (dry)
CO	0.5	0.2	0.2	0.1	0.1	mg/Nm ³ (dry)

Removal efficiency:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.264	0.518	0.425	0.348	0.270	0.356
SO ₂	0.912	0.894	0.912	0.914	0.926	0.913
HCl	0.983	0.974	0.977	0.977	0.976	0.977

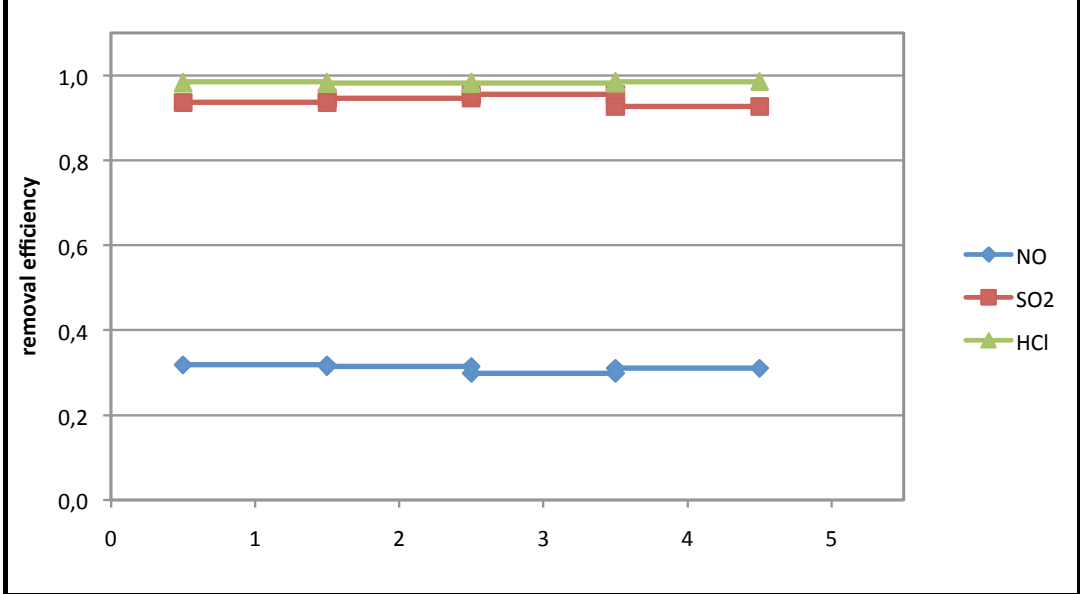


18.06.2013		Test run 01-B-F20		Page 1 of 2	
Operating conditions of the incineration plant:					
Primary fuel:	sewage sludge	feed rate:	626 kg/h		
Secondary fuel:	screenings	feed rate:	0 kg/h		
Bed material:	silica sand	amount:	2000 kg		
Start Test Run: 06:00		Stable:	06:30	End Test Run: 08:30	
Measurements flue gas:					
		Begin stable:	End stable:		
Flue gas volume flow	FI0A	7830	7730	Nm³/h	
Fluidized bed temp.	TC2	729	731	°C	
Free board temp.	TC3.1	840	833	°C	
Afterburner chamber temp.	TC4.1	945	906	°C	
Outlet TO-Heater temp.	TC7	284	277	°C	
Inlet filter temp.	TC9	175	180	°C	
Inlet quench cooler temp.	TI5805	145	145	°C	
Outlet quench cooler temp.	TI5810	61.2	62.2	°C	
Outlet spray scrubber temp.	TI5803	36.1	34.8	°C	
Stack temp.	TC0A	56	56	°C	
Measurements scrubbing liquid:					
		Begin stable:	End stable:		
Fresh liquid flow quench	FIS5803	0.0	0.0	m³/h	
Recirculation liquid flow quench	by calc.	4.7	3.8	m³/h	
Total liquid flow quench	FIS5806	4.7	3.8	m³/h	
Mix liquid temp. quench	TI5804	38.0	38.6	°C	
Fresh liquid flow spray scrubber	FI5801	20.0	20.0	m³/h	
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m³/h	
Total liquid flow spray scrubber	FI5802	20.0	20.0	m³/h	
Liquid to gas ratio	by calc.	3.18	3.22	kmol/kmol	
Recirculation ratio	by calc.	0.0	0.0	-	
Mix liquid temp. spray scrubber	TI5802	17.7	17.7	°C	
Fresh liquid temperature	TI5801	17.3	17.3	°C	
Used liquid temperature	TI5808	39.3	39.4	°C	

Flue gas analysis - Raw gas:					
	HMW 1	HMW 2	HMW 3	HMW 4	
O ₂	14.4%	14.4%	14.4%	14.3%	vol%
H ₂ O	18.7%	18.7%	18.6%	18.8%	vol%
NO	35.2	35.0	34.2	31.9	mg/Nm ³ (dry)
NO ₂	1.7	1.5	1.2	0.6	mg/Nm ³ (dry)
SO ₂	149	140.5	117.4	175.1	mg/Nm ³ (dry)
HCl	2.9	2.6	2.2	4.3	mg/Nm ³ (dry)
CO	12.3	9.6	4.7	4.4	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:					
	HMW 1	HMW 2	HMW 3	HMW 4	
O ₂	15.5%	15.5%	15.5%	15.3%	vol%
H ₂ O	7.2%	7.2%	7.0%	7.1%	vol%
NO	24.0	24.0	24.0	22.0	mg/Nm ³ (dry)
NO ₂	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
SO ₂	9.5	7.5	5.3	12.8	mg/Nm ³ (dry)
HCl	0.0	0.0	0.0	0.1	mg/Nm ³ (dry)
CO	8.6	2.7	2.1	4.3	mg/Nm ³ (dry)

Removal efficiency:					
	HMW 1	HMW 2	HMW 3	HMW 4	AV
NO	0.318	0.314	0.298	0.310	0.310
SO ₂	0.936	0.947	0.955	0.927	0.940
HCl	0.984	0.982	0.984	0.986	0.984



Operating conditions of the incineration plant:

Primary fuel: sewage sludge	feed rate:	591 kg/h
Secondary fuel: screenings	feed rate:	0 kg/h
Bed material: silica sand	amount:	2000 kg

Start Test Run: 00:00	Stable: 06:00	End Test Run: 08:45
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Measurements flue gas:

Begin stable: End stable:

Flue gas volume flow	FI0A	7480	7480	Nm ³ /h	
Fluidized bed temp.	TC2	731	734	°C	
Free board temp.	TC3.1	826	805	°C	
Afterburner chamber temp.	TC4.1	875	847	°C	
Outlet TO-Heater temp.	TC7	269	280	°C	
Inlet filter temp.	TC9	182	172	°C	
Inlet quench cooler temp.	TI5805	171	162	°C	
Outlet quench cooler temp.	TI5810	58.4	56	°C	
Outlet spray scrubber temp.	TI5803	53.2	49.9	°C	
Stack temp.	TC0A	78.3	75.3	°C	

Measurements scrubbing liquid:

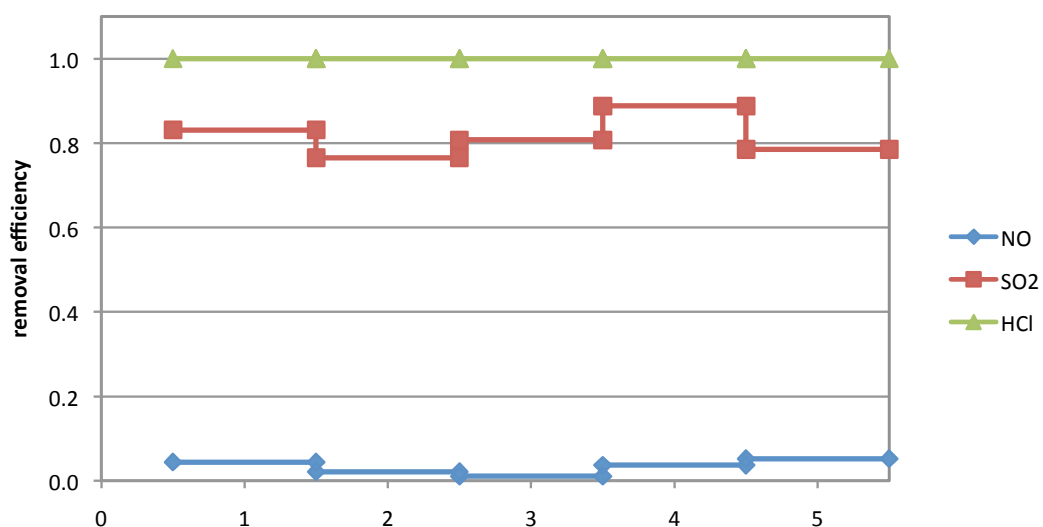
Begin stable: End stable:

Fresh liquid flow quench	FIS5803	0.0	0.0	m ³ /h	
Recirculation liquid flow quench	by calc.	10.0	10.1	m ³ /h	
Total liquid flow quench	FIS5806	10.0	10.1	m ³ /h	
Mix liquid temp. quench	TI5804	48.3	46.0	°C	
Fresh liquid flow spray scrubber	FI5801	10.7	10.7	m ³ /h	
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m ³ /h	
Total liquid flow spray scrubber	FI5802	10.7	10.7	m ³ /h	
Liquid to gas ratio	by calc.	1.78	1.78	kmol/kmol	
Recirculation ratio	by calc.	0.0	0.0	-	
Mix liquid temp. spray scrubber	TI5802	19.4	19.3	°C	
Fresh liquid temperature	TI5801	18.9	18.8	°C	
Used liquid temperature	TI5808	48.0	45.7	°C	

Flue gas analysis - Raw gas:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	13.1	13.1	13.2	13.2	13.1	vol%
H ₂ O	20.7	20.9	20.3	20.4	20.5	vol%
NO	52.3	41.9	45.5	43.6	42.2	mg/Nm ³ (dry)
NO ₂	0.0	0.0	0.0	0.0	0.0	mg/Nm ³ (dry)
SO ₂	69.2	44.7	54	88.2	130.3	mg/Nm ³ (dry)
HCl	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
CO	2.5	2.6	2.5	2.5	2.5	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	13.2	13.5	13.5	13.6	13.9	vol%
H ₂ O	19.7	19.4	19.3	18.7	18.1	vol%
NO	50.0	41.0	45.0	42.0	40.0	mg/Nm ³ (dry)
NO ₂	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
SO ₂	11.7	10.5	10.4	9.9	28	mg/Nm ³ (dry)
HCl	0.0	0.0	0.0	0.0	0.0	mg/Nm ³ (dry)
CO	2.8	2.6	2.7	2.3	2.5	mg/Nm ³ (dry)

Removal efficiency:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.044	0.021	0.011	0.037	0.052	0.033
SO ₂	0.831	0.765	0.807	0.888	0.785	0.818
HCl	1.000	1.000	1.000	1.000	1.000	1.000



Operating conditions of the incineration plant:

Primary fuel: sewage sludge	feed rate:	599 kg/h
Secondary fuel: screenings	feed rate:	0 kg/h
Bed material: silica sand	amount:	2000 kg

Start Test Run: 10:00	Stable: 12:30	End Test Run: 15:00
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Measurements flue gas:

Begin stable: End stable:

Flue gas volume flow	FI0A	7830	7810	Nm ³ /h
Fluidized bed temp.	TC2	723	738	°C
Free board temp.	TC3.1	850	823	°C
Afterburner chamber temp.	TC4.1	883	876	°C
Outlet TO-Heater temp.	TC7	290	289	°C
Inlet filter temp.	TC9	180	184	°C
Inlet quench cooler temp.	TI5805	170	170	°C
Outlet quench cooler temp.	TI5810	60.6	60.2	°C
Outlet spray scrubber temp.	TI5803	48.6	48	°C
Stack temp.	TC0A	75.7	74.7	°C

Measurements scrubbing liquid:

Begin stable: End stable:

Fresh liquid flow quench	FIS5803	0.0	0.0	m ³ /h
Recirculation liquid flow quench	by calc.	10.1	10.2	m ³ /h
Total liquid flow quench	FIS5806	10.1	10.2	m ³ /h
Mix liquid temp. quench	TI5804	53.3	52.3	°C
Fresh liquid flow spray scrubber	FI5801	10.4	10.6	m ³ /h
Recirculation liquid flow spray sc.	by calc.	5.6	5.5	m ³ /h
Total liquid flow spray scrubber	FI5802	16.0	16.1	m ³ /h
Liquid to gas ratio	by calc.	2.54	2.57	kmol/kmol
Recirculation ratio	by calc.	0.5	0.5	-
Mix liquid temp. spray scrubber	TI5802	35.3	32.2	°C
Fresh liquid temperature	TI5801	19.2	19.5	°C
Used liquid temperature	TI5808	53.2	51.9	°C

Flue gas analysis - Raw gas:

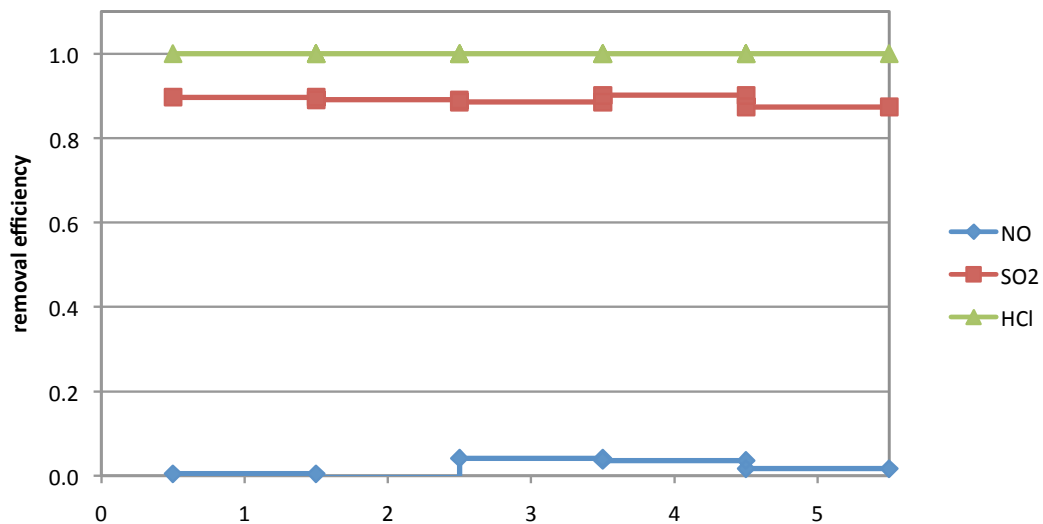
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	13.6	13.7	13.4	13.5	13.7	vol%
H ₂ O	19.0	19.0	20.0	19.4	19.2	vol%
NO	44.2	46.8	40.7	41.5	41.7	mg/Nm ³ (dry)
NO ₂	0.0	0.0	0.0	0.0	0.0	mg/Nm ³ (dry)
SO ₂	114.4	120	106.3	104.5	81.5	mg/Nm ³ (dry)
HCl	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
CO	0.2	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.4	14.5	14.2	14.4	14.4	vol%
H ₂ O	12.6	12.5	12.5	12.9	12.7	vol%
NO	44.0	47.0	39.0	40.0	41.0	mg/Nm ³ (dry)
NO ₂	0.2	0.2	0.2	0.2	0.2	mg/Nm ³ (dry)
SO ₂	11.8	13.1	12.2	10.3	10.3	mg/Nm ³ (dry)
HCl	0.0	0.0	0.0	0.0	0.0	mg/Nm ³ (dry)
CO	1	0.7	0.7	1.1	0.8	mg/Nm ³ (dry)

Removal efficiency:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.005	-0.004	0.042	0.036	0.017	0.018
SO ₂	0.897	0.891	0.885	0.901	0.874	0.890
HCl	1.000	1.000	1.000	1.000	1.000	1.000



Operating conditions of the incineration plant:

Primary fuel: sewage sludge	feed rate:	601 kg/h
Secondary fuel: screenings	feed rate:	0 kg/h
Bed material: silica sand	amount:	2000 kg

Start Test Run: 00:00	Stable: 05:30	End Test Run: 08:00
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Measurements flue gas:

Begin stable: End stable:

Flue gas volume flow	FI0A	7880	7900	Nm ³ /h
Fluidized bed temp.	TC2	729	734	°C
Free board temp.	TC3.1	828	839	°C
Afterburner chamber temp.	TC4.1	870	879	°C
Outlet TO-Heater temp.	TC7	285	289	°C
Inlet filter temp.	TC9	186	188	°C
Inlet quench cooler temp.	TI5805	164	162	°C
Outlet quench cooler temp.	TI5810	59.6	59.7	°C
Outlet spray scrubber temp.	TI5803	47.9	48	°C
Stack temp.	TC0A	72	71.2	°C

Measurements scrubbing liquid:

Begin stable: End stable:

Fresh liquid flow quench	FIS5803	0.0	0.0	m ³ /h
Recirculation liquid flow quench	by calc.	9.9	9.9	m ³ /h
Total liquid flow quench	FIS5806	9.9	9.9	m ³ /h
Mix liquid temp. quench	TI5804	52.5	52.0	°C
Fresh liquid flow spray scrubber	FI5801	10.7	10.8	m ³ /h
Recirculation liquid flow spray sc.	by calc.	10.3	10.4	m ³ /h
Total liquid flow spray scrubber	FI5802	21.0	21.2	m ³ /h
Liquid to gas ratio	by calc.	3.32	3.34	kmol/kmol
Recirculation ratio	by calc.	1.0	1.0	-
Mix liquid temp. spray scrubber	TI5802	40.5	38.6	°C
Fresh liquid temperature	TI5801	19.4	19.3	°C
Used liquid temperature	TI5808	52.1	52.0	°C

Flue gas analysis - Raw gas:

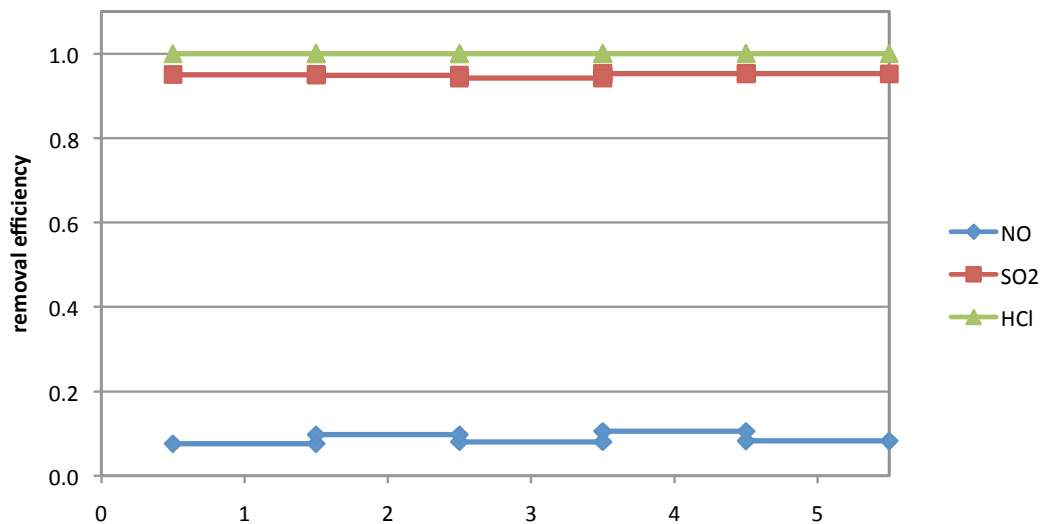
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.1	14.1	14.0	14.1	14.1	vol%
H ₂ O	18.5	18.5	17.8	18.2	17.5	vol%
NO	42.2	43.2	44.6	44.7	42.5	mg/Nm ³ (dry)
NO ₂	0.0	0.0	0.0	0.0	0.0	mg/Nm ³ (dry)
SO ₂	134.8	143.3	138.7	158.4	153.4	mg/Nm ³ (dry)
HCl	0.1	0.1	0.1	0.1	0.4	mg/Nm ³ (dry)
CO	0.1	0.1	0.7	0.2	0.2	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	15.0	15.0	15.2	15.0	15.1	vol%
H ₂ O	11.9	11.9	11.7	11.9	11.4	vol%
NO	39.0	39.0	41.0	40.0	39.0	mg/Nm ³ (dry)
NO ₂	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
SO ₂	6.7	7.3	8	7.4	7.3	mg/Nm ³ (dry)
HCl	0.0	0.0	0.0	0.0	0.0	mg/Nm ³ (dry)
CO	1.9	2	2.5	1.1	1.5	mg/Nm ³ (dry)

Removal efficiency:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.076	0.097	0.081	0.105	0.082	0.088
SO ₂	0.950	0.949	0.942	0.953	0.952	0.950
HCl	1.000	1.000	1.000	1.000	1.000	1.000



30.08.2013		Test run 02-R-150		Page 1 of 2	
Operating conditions of the incineration plant:					
Primary fuel:	sewage sludge	feed rate:	613 kg/h		
Secondary fuel:	screenings	feed rate:	0 kg/h		
Bed material:	silica sand	amount:	2000 kg		
Start Test Run: 08:00		Stable:	10:00	End Test Run: 12:30	
Measurements flue gas:					
		Begin stable:	End stable:		
Flue gas volume flow	FI0A	8080	8090	Nm³/h	
Fluidized bed temp.	TC2	726	722	°C	
Free board temp.	TC3.1	820	804	°C	
Afterburner chamber temp.	TC4.1	855	853	°C	
Outlet TO-Heater temp.	TC7	283	284	°C	
Inlet filter temp.	TC9	177	183	°C	
Inlet quench cooler temp.	TI5805	157	158	°C	
Outlet quench cooler temp.	TI5810	59.9	59.9	°C	
Outlet spray scrubber temp.	TI5803	48.7	48.8	°C	
Stack temp.	TC0A	72.7	72.6	°C	
Measurements scrubbing liquid:					
		Begin stable:	End stable:		
Fresh liquid flow quench	FIS5803	0.0	0.0	m³/h	
Recirculation liquid flow quench	by calc.	10.0	9.9	m³/h	
Total liquid flow quench	FIS5806	10.0	9.9	m³/h	
Mix liquid temp. quench	TI5804	54.0	54.5	°C	
Fresh liquid flow spray scrubber	FI5801	10.3	10.4	m³/h	
Recirculation liquid flow spray sc.	by calc.	15.4	15.3	m³/h	
Total liquid flow spray scrubber	FI5802	25.7	25.7	m³/h	
Liquid to gas ratio	by calc.	3.96	3.96	kmol/kmol	
Recirculation ratio	by calc.	1.5	1.5	-	
Mix liquid temp. spray scrubber	TI5802	42.4	42.5	°C	
Fresh liquid temperature	TI5801	19.4	19.6	°C	
Used liquid temperature	TI5808	54.2	54.7	°C	

Flue gas analysis - Raw gas:

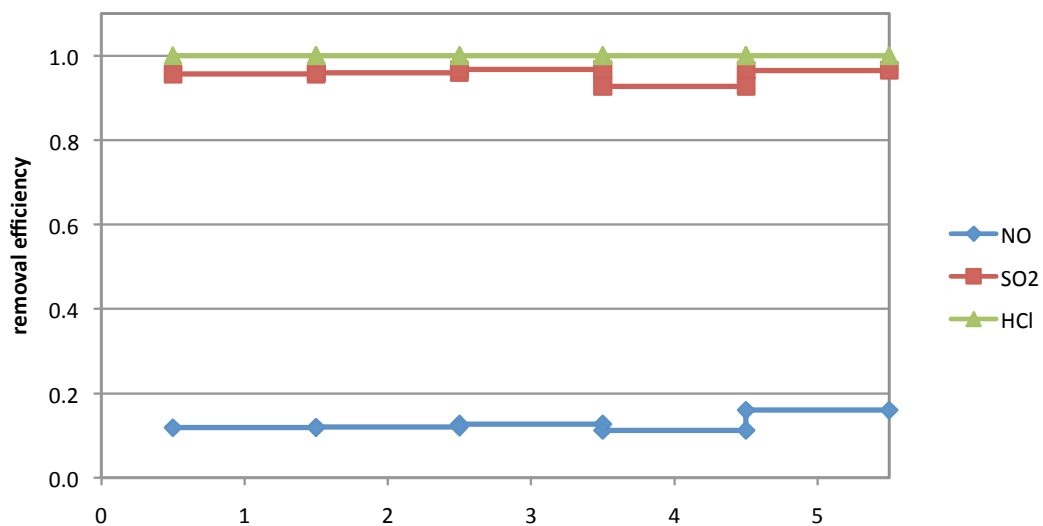
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.4	14.4	14.4	14.4	14.2	vol%
H ₂ O	18.1	18.5	18.1	18.6	18.9	vol%
NO	43.1	40.9	42.4	42.8	39.3	mg/Nm ³ (dry)
NO ₂	0.9	0.0	0.0	0.0	0.0	mg/Nm ³ (dry)
SO ₂	64.1	52.2	63.6	34.4	82.7	mg/Nm ³ (dry)
HCl	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
CO	1	1.4	1.6	2.3	2.3	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	15.3	15.2	15.2	15.2	15.0	vol%
H ₂ O	12.5	12.7	12.6	12.8	12.8	vol%
NO	38.0	36.0	37.0	38.0	33.0	mg/Nm ³ (dry)
NO ₂	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
SO ₂	2.8	2.1	2.1	2.5	2.9	mg/Nm ³ (dry)
HCl	0.0	0.0	0.0	0.0	0.0	mg/Nm ³ (dry)
CO	3.4	3.7	3.5	3.8	3.7	mg/Nm ³ (dry)

Removal efficiency:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.118	0.120	0.127	0.112	0.160	0.127
SO ₂	0.956	0.960	0.967	0.927	0.965	0.958
HCl	1.000	1.000	1.000	1.000	1.000	1.000



19.06.2013		Test run 03-A-T19		Page 1 of 2	
Operating conditions of the incineration plant:					
Primary fuel:	sewage sludge	feed rate:	574 kg/h		
Secondary fuel:	screenings	feed rate:	0 kg/h		
Bed material:	silica sand	amount:	2000 kg		
Start Test Run: 12:10		Stable:	12:35	End Test Run: 15:00	
Measurements flue gas:					
		Begin stable:	End stable:		
Flue gas volume flow	FI0A	7810	7830	Nm³/h	
Fluidized bed temp.	TC2	729	727	°C	
Free board temp.	TC3.1	845	842	°C	
Afterburner chamber temp.	TC4.1	915	915	°C	
Outlet TO-Heater temp.	TC7	276	278	°C	
Inlet filter temp.	TC9	154	162	°C	
Inlet quench cooler temp.	TI5805	137	141	°C	
Outlet quench cooler temp.	TI5810	53.7	54	°C	
Outlet spray scrubber temp.	TI5803	39.2	39.6	°C	
Stack temp.	TC0A	58.5	59.7	°C	
Measurements scrubbing liquid:					
		Begin stable:	End stable:		
Fresh liquid flow quench	FIS5803	0.0	0.0	m³/h	
Recirculation liquid flow quench	by calc.	20.2	20.7	m³/h	
Total liquid flow quench	FIS5806	20.2	20.7	m³/h	
Mix liquid temp. quench	TI5804	45.7	46.8	°C	
Fresh liquid flow spray scrubber	FI5801	14.8	14.9	m³/h	
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m³/h	
Total liquid flow spray scrubber	FI5802	14.8	14.9	m³/h	
Liquid to gas ratio	by calc.	2.36	2.37	kmol/kmol	
Recirculation ratio	by calc.	0.0	0.0	-	
Mix liquid temp. spray scrubber	TI5802	18.8	19.0	°C	
Fresh liquid temperature	TI5801	18.1	18.4	°C	
Used liquid temperature	TI5808	45.8	47.1	°C	

Flue gas analysis - Raw gas:

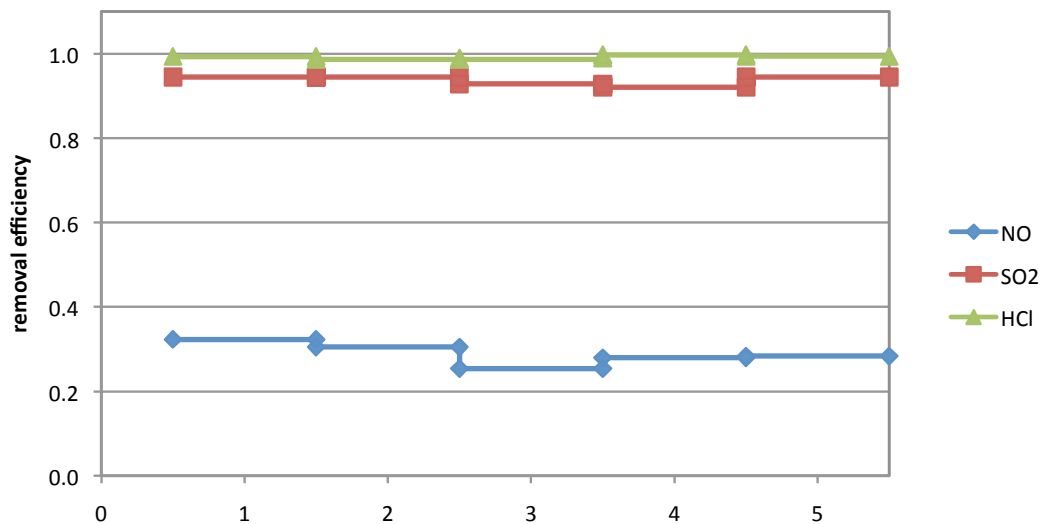
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.5%	14.5%	14.3%	14.4%	14.5%	vol%
H ₂ O	19.2%	19.6%	20.1%	19.7%	19.4%	vol%
NO	32.5	33.1	33.5	33.3	32.1	mg/Nm ³ (dry)
NO ₂	0.1	0.0	0.3	0.2	0.0	mg/Nm ³ (dry)
SO ₂	99.4	97.7	195.3	185.1	101.1	mg/Nm ³ (dry)
HCl	15.1	7.4	9.2	36.1	16.3	mg/Nm ³ (dry)
CO	4.2	4.4	4.5	4.5	4.4	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	15.5%	15.5%	15.3%	15.3%	15.5%	vol%
H ₂ O	8.5%	8.7%	8.6%	8.5%	8.7%	vol%
NO	22.0	23.0	25.0	24.0	23.0	mg/Nm ³ (dry)
NO ₂	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
SO ₂	5.5	5.4	14	14.7	5.6	mg/Nm ³ (dry)
HCl	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
CO	1.3	2.1	2.4	1.8	1.9	mg/Nm ³ (dry)

Removal efficiency:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.323	0.305	0.254	0.279	0.283	0.289
SO ₂	0.945	0.945	0.928	0.921	0.945	0.933
HCl	0.993	0.986	0.989	0.997	0.994	0.994

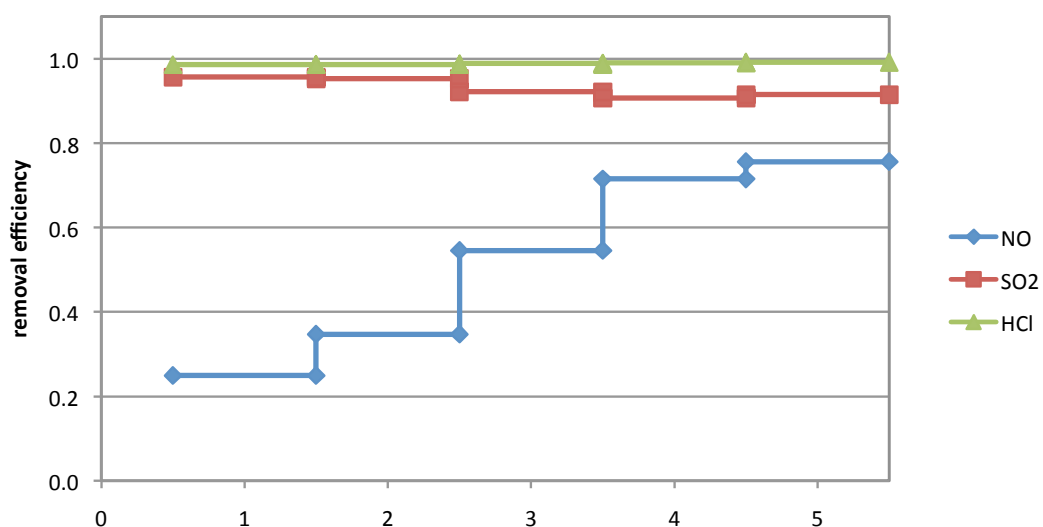


29.08.2013		Test run 03-A-T24		Page 1 of 2	
Operating conditions of the incineration plant:					
Primary fuel:	sewage sludge	feed rate:	706 kg/h		
Secondary fuel:	screenings	feed rate:	0 kg/h		
Bed material:	silica sand	amount:	2000 kg		
Start Test Run:	08:15	Stable:	09:00	End Test Run: 11:30	
Measurements flue gas:					
		Begin stable:	End stable:		
Flue gas volume flow	FI0A	8070	8050	Nm³/h	
Fluidized bed temp.	TC2	776	732	°C	
Free board temp.	TC3.1	812	823	°C	
Afterburner chamber temp.	TC4.1	867	884	°C	
Outlet TO-Heater temp.	TC7	295	295	°C	
Inlet filter temp.	TC9	171	179	°C	
Inlet quench cooler temp.	TI5805	153	156	°C	
Outlet quench cooler temp.	TI5810	54.8	56.5	°C	
Outlet spray scrubber temp.	TI5803	42.9	43.9	°C	
Stack temp.	TC0A	69.3	70	°C	
Measurements scrubbing liquid:					
		Begin stable:	End stable:		
Fresh liquid flow quench	FIS5803	0.0	0.0	m³/h	
Recirculation liquid flow quench	by calc.	24.5	24.6	m³/h	
Total liquid flow quench	FIS5806	24.5	24.6	m³/h	
Mix liquid temp. quench	TI5804	47.5	48.0	°C	
Fresh liquid flow spray scrubber	FI5801	15.5	15.4	m³/h	
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m³/h	
Total liquid flow spray scrubber	FI5802	15.5	15.4	m³/h	
Liquid to gas ratio	by calc.	2.39	2.38	kmol/kmol	
Recirculation ratio	by calc.	0.0	0.0	-	
Mix liquid temp. spray scrubber	TI5802	24.2	24.4	°C	
Fresh liquid temperature	TI5801	19.0	19.1	°C	
Used liquid temperature	TI5808	51.1	51.5	°C	

Flue gas analysis - Raw gas:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	13.9	13.7	13.7	13.9	13.7	vol%
H ₂ O	20.5	21.1	21.2	21.0	20.9	vol%
NO	53.3	42.9	30.8	28.1	28.7	mg/Nm ³ (dry)
NO ₂	0.0	0.0	0.0	0.0	0.0	mg/Nm ³ (dry)
SO ₂	94.6	114.2	84.3	73	89.4	mg/Nm ³ (dry)
HCl	7.2	7.3	8.6	9.7	11.8	mg/Nm ³ (dry)
CO	5.3	5.1	4.6	4.8	5.1	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.7	14.4	14.5	14.5	14.4	vol%
H ₂ O	10.0	10.2	10.6	10.4	10.3	vol%
NO	40.0	28.0	14.0	8.0	7.0	mg/Nm ³ (dry)
NO ₂	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
SO ₂	4.1	5.4	6.6	6.8	7.6	mg/Nm ³ (dry)
HCl	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
CO	4.1	3.7	1.6	1.4	1.5	mg/Nm ³ (dry)

Removal efficiency:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.250	0.347	0.545	0.715	0.756	0.472
SO ₂	0.957	0.953	0.922	0.907	0.915	0.933
HCl	0.986	0.986	0.988	0.990	0.992	0.989



Operating conditions of the incineration plant:

Primary fuel: sewage sludge	feed rate:	628 kg/h
Secondary fuel: screenings	feed rate:	0 kg/h
Bed material: silica sand	amount:	2000 kg

Start Test Run: 12:30	Stable: 13:30	End Test Run: 16:00
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Measurements flue gas:

Begin stable: End stable:

Flue gas volume flow	FI0A	8050	8080	Nm ³ /h
Fluidized bed temp.	TC2	734	740	°C
Free board temp.	TC3.1	821	826	°C
Afterburner chamber temp.	TC4.1	876	870	°C
Outlet TO-Heater temp.	TC7	291	289	°C
Inlet filter temp.	TC9	173	181	°C
Inlet quench cooler temp.	TI5805	150	157	°C
Outlet quench cooler temp.	TI5810	53	53.2	°C
Outlet spray scrubber temp.	TI5803	45.5	47.1	°C
Stack temp.	TC0A	70.1	70.1	°C

Measurements scrubbing liquid:

Begin stable: End stable:

Fresh liquid flow quench	FIS5803	0.0	0.0	m ³ /h
Recirculation liquid flow quench	by calc.	22.5	22.4	m ³ /h
Total liquid flow quench	FIS5806	22.5	22.4	m ³ /h
Mix liquid temp. quench	TI5804	41.5	41.9	°C
Fresh liquid flow spray scrubber	FI5801	15.5	15.4	m ³ /h
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m ³ /h
Total liquid flow spray scrubber	FI5802	15.5	15.4	m ³ /h
Liquid to gas ratio	by calc.	2.40	2.37	kmol/kmol
Recirculation ratio	by calc.	0.0	0.0	-
Mix liquid temp. spray scrubber	TI5802	29.9	30.1	°C
Fresh liquid temperature	TI5801	19.2	19.2	°C
Used liquid temperature	TI5808	49.8	49.5	°C

Flue gas analysis - Raw gas:

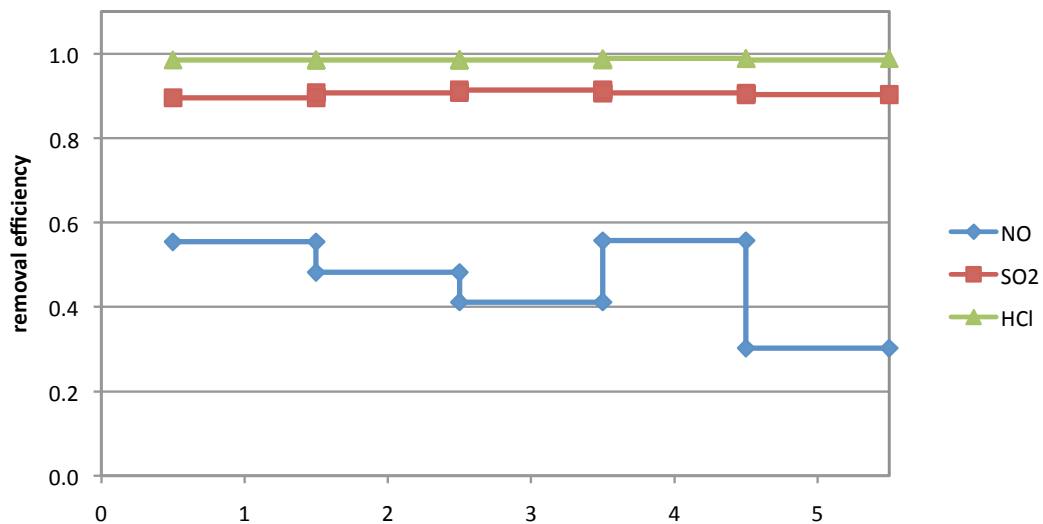
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	13.8	13.9	13.9	13.8	14.1	vol%
H ₂ O	20.9	20.2	20.2	20.8	19.7	vol%
NO	31.4	32.8	34.0	31.6	38.7	mg/Nm ³ (dry)
NO ₂	2.3	0.8	0.0	0.1	0.0	mg/Nm ³ (dry)
SO ₂	100.4	102	112.5	108.1	89.9	mg/Nm ³ (dry)
HCl	6.7	7	7.7	8.7	8.5	mg/Nm ³ (dry)
CO	11.7	5.6	7.2	4.9	4.6	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.5	14.7	14.7	14.5	14.9	vol%
H ₂ O	11.4	11.2	11.0	11.2	11.5	vol%
NO	14.0	17.0	20.0	14.0	27.0	mg/Nm ³ (dry)
NO ₂	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
SO ₂	10.5	9.4	9.7	10.1	8.7	mg/Nm ³ (dry)
HCl	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
CO	8.1	1.8	3.4	1.8	2	mg/Nm ³ (dry)

Removal efficiency:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.554	0.482	0.412	0.557	0.302	0.454
SO ₂	0.895	0.908	0.914	0.907	0.903	0.906
HCl	0.985	0.986	0.987	0.989	0.988	0.987



29.08.2013		Test run 03-A-T37		Page 1 of 2	
Operating conditions of the incineration plant:					
Primary fuel:	sewage sludge	feed rate:	716 kg/h		
Secondary fuel:	screenings	feed rate:	0 kg/h		
Bed material:	silica sand	amount:	2000 kg		
Start Test Run:	00:00	Stable:	05:30	End Test Run:	08:15
Measurements flue gas:					
		Begin stable:	End stable:		
Flue gas volume flow	FI0A	8050	8050	Nm³/h	
Fluidized bed temp.	TC2	731	740	°C	
Free board temp.	TC3.1	802	815	°C	
Afterburner chamber temp.	TC4.1	867	877	°C	
Outlet TO-Heater temp.	TC7	294	293	°C	
Inlet filter temp.	TC9	170	176	°C	
Inlet quench cooler temp.	TI5805	150	152	°C	
Outlet quench cooler temp.	TI5810	49.3	49.8	°C	
Outlet spray scrubber temp.	TI5803	48.4	49.4	°C	
Stack temp.	TC0A	72.7	73.1	°C	
Measurements scrubbing liquid:					
		Begin stable:	End stable:		
Fresh liquid flow quench	FIS5803	0.0	0.0	m³/h	
Recirculation liquid flow quench	by calc.	19.6	19.5	m³/h	
Total liquid flow quench	FIS5806	19.6	19.5	m³/h	
Mix liquid temp. quench	TI5804	34.8	35.4	°C	
Fresh liquid flow spray scrubber	FI5801	15.3	15.0	m³/h	
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m³/h	
Total liquid flow spray scrubber	FI5802	15.5	15.0	m³/h	
Liquid to gas ratio	by calc.	2.40	2.32	kmol/kmol	
Recirculation ratio	by calc.	0.0	0.0	-	
Mix liquid temp. spray scrubber	TI5802	37.0	36.9	°C	
Fresh liquid temperature	TI5801	19.1	19.0	°C	
Used liquid temperature	TI5808	47.9	47.9	°C	

Flue gas analysis - Raw gas:

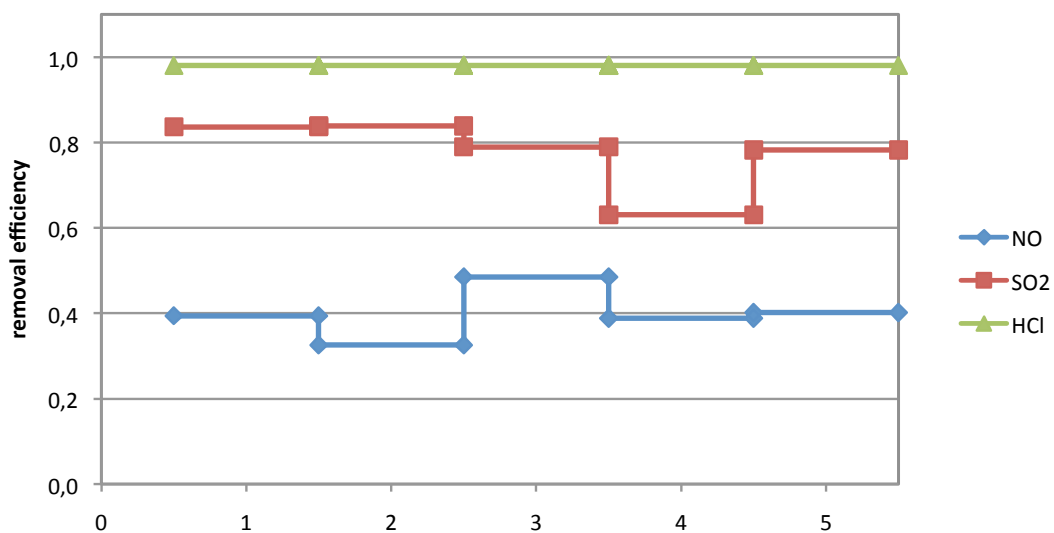
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	13.6%	13.8%	13.7%	13.7%	13.6%	vol%
H ₂ O	21.5%	21.6%	21.7%	21.2%	21.6%	vol%
NO	33.0	35.6	31.1	32.7	35.1	mg/Nm ³ (dry)
NO ₂	0.0	0.0	0.0	0.0	0.0	mg/Nm ³ (dry)
SO ₂	45.9	51.5	38.5	21.4	27.6	mg/Nm ³ (dry)
HCl	5.1	5.2	5.4	5.1	5.1	mg/Nm ³ (dry)
CO	5	5.2	4.9	5	5.2	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.5%	14.6%	14.5%	14.6%	14.5%	vol%
H ₂ O	12.7%	13.0%	12.9%	12.9%	12.9%	vol%
NO	20.0	24.0	16.0	20.0	21.0	mg/Nm ³ (dry)
NO ₂	0.2	0.2	0.1	0.1	0.2	mg/Nm ³ (dry)
SO ₂	7.5	8.3	8.1	7.9	6	mg/Nm ³ (dry)
HCl	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
CO	3	3.5	2.5	2.7	3.8	mg/Nm ³ (dry)

Removal efficiency:

	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.394	0.326	0.486	0.388	0.402	0.397
SO ₂	0.837	0.839	0.790	0.631	0.783	0.796
HCl	0.980	0.981	0.981	0.980	0.980	0.981

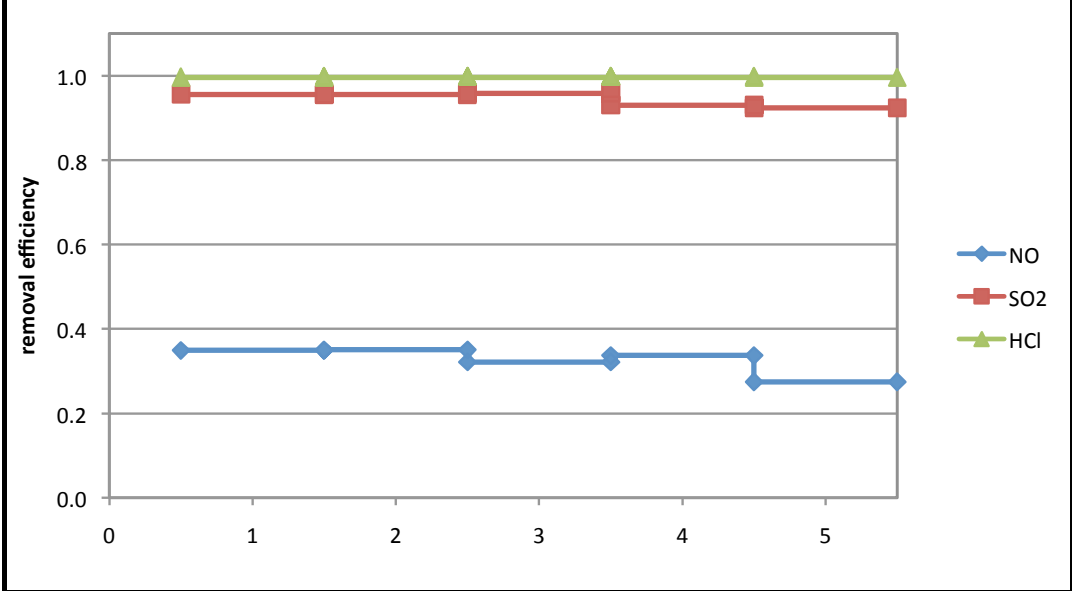


19.06.2013		Test run 03-B-T15		Page 1 of 2	
Operating conditions of the incineration plant:					
Primary fuel:	sewage sludge	feed rate:	577 kg/h		
Secondary fuel:	screenings	feed rate:	0 kg/h		
Bed material:	silica sand	amount:	2000 kg		
Start Test Run: 08:30		Stable:	09:30	End Test Run: 12:00	
Measurements flue gas:					
		Begin stable:	End stable:		
Flue gas volume flow	FI0A	7780	7810	Nm³/h	
Fluidized bed temp.	TC2	727	730	°C	
Free board temp.	TC3.1	835	843	°C	
Afterburner chamber temp.	TC4.1	907	915	°C	
Outlet TO-Heater temp.	TC7	278	275	°C	
Inlet filter temp.	TC9	145	157	°C	
Inlet quench cooler temp.	TI5805	134	141	°C	
Outlet quench cooler temp.	TI5810	52	51.1	°C	
Outlet spray scrubber temp.	TI5803	32	31.6	°C	
Stack temp.	TC0A	53.3	53.6	°C	
Measurements scrubbing liquid:					
		Begin stable:	End stable:		
Fresh liquid flow quench	FIS5803	0.0	0.0	m³/h	
Recirculation liquid flow quench	by calc.	20.2	20.2	m³/h	
Total liquid flow quench	FIS5806	20.2	20.2	m³/h	
Mix liquid temp. quench	TI5804	42.5	41.5	°C	
Fresh liquid flow spray scrubber	FI5801	20.0	20.0	m³/h	
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m³/h	
Total liquid flow spray scrubber	FI5802	20.0	20.0	m³/h	
Liquid to gas ratio	by calc.	3.20	3.19	kmol/kmol	
Recirculation ratio	by calc.	0.0	0.0	-	
Mix liquid temp. spray scrubber	TI5802	18.4	18.6	°C	
Fresh liquid temperature	TI5801	17.9	18.0	°C	
Used liquid temperature	TI5808	42.8	41.6	°C	

Flue gas analysis - Raw gas:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.2%	14.4%	14.3%	14.5%	14.5%	vol%
H ₂ O	19.8%	19.7%	19.9%	19.1%	19.5%	vol%
NO	33.8	35.4	33.9	34.7	31.7	mg/Nm ³ (dry)
NO ₂	0.8	0.9	0.7	0.5	0.2	mg/Nm ³ (dry)
SO ₂	236.4	232.9	263.9	191.8	141.7	mg/Nm ³ (dry)
HCl	44.3	74.8	97.5	77.5	23.7	mg/Nm ³ (dry)
CO	4	4.2	4.2	4.6	4.3	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	15.2%	15.3%	15.2%	15.4%	15.4%	vol%
H ₂ O	6.1%	6.1%	6.1%	6.1%	6.1%	vol%
NO	22.0	23.0	23.0	23.0	23.0	mg/Nm ³ (dry)
NO ₂	0.0	0.0	0.0	0.0	0.0	mg/Nm ³ (dry)
SO ₂	10.4	10.4	11	13.4	10.8	mg/Nm ³ (dry)
HCl	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
CO	1.1	1.1	1.1	1.9	1	mg/Nm ³ (dry)

Removal efficiency:						
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.349	0.350	0.322	0.337	0.274	0.327
SO ₂	0.956	0.955	0.958	0.930	0.924	0.948
HCl	0.998	0.999	0.999	0.999	0.996	0.998



Operating conditions of the incineration plant:

Primary fuel: sewage sludge	feed rate:	626 kg/h
Secondary fuel: screenings	feed rate:	0 kg/h
Bed material: silica sand	amount:	2000 kg

Start Test Run: 08:30	Stable: 10:00	End Test Run: 12:30
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Measurements flue gas:

Begin stable: End stable:

Flue gas volume flow	FI0A	8080	8050	Nm ³ /h
Fluidized bed temp.	TC2	731	731	°C
Free board temp.	TC3.1	816	814	°C
Afterburner chamber temp.	TC4.1	859	870	°C
Outlet TO-Heater temp.	TC7	283	289	°C
Inlet filter temp.	TC9	179	174	°C
Inlet quench cooler temp.	TI5805	150	154	°C
Outlet quench cooler temp.	TI5810	50.5	50.6	°C
Outlet spray scrubber temp.	TI5803	42.3	41.9	°C
Stack temp.	TC0A	66.5	67.5	°C

Measurements scrubbing liquid:

Begin stable: End stable:

Fresh liquid flow quench	FIS5803	0.0	0.0	m ³ /h
Recirculation liquid flow quench	by calc.	16.4	16.4	m ³ /h
Total liquid flow quench	FIS5806	16.4	16.4	m ³ /h
Mix liquid temp. quench	TI5804	33.1	33.1	°C
Fresh liquid flow spray scrubber	FI5801	19.6	20.1	m ³ /h
Recirculation liquid flow spray sc.	by calc.	0.0	0.0	m ³ /h
Total liquid flow spray scrubber	FI5802	19.6	20.1	m ³ /h
Liquid to gas ratio	by calc.	3.02	3.11	kmol/kmol
Recirculation ratio	by calc.	0.0	0.0	-
Mix liquid temp. spray scrubber	TI5802	30.1	30.0	°C
Fresh liquid temperature	TI5801	19.2	19.2	°C
Used liquid temperature	TI5808	46.1	46.0	°C

Flue gas analysis - Raw gas:

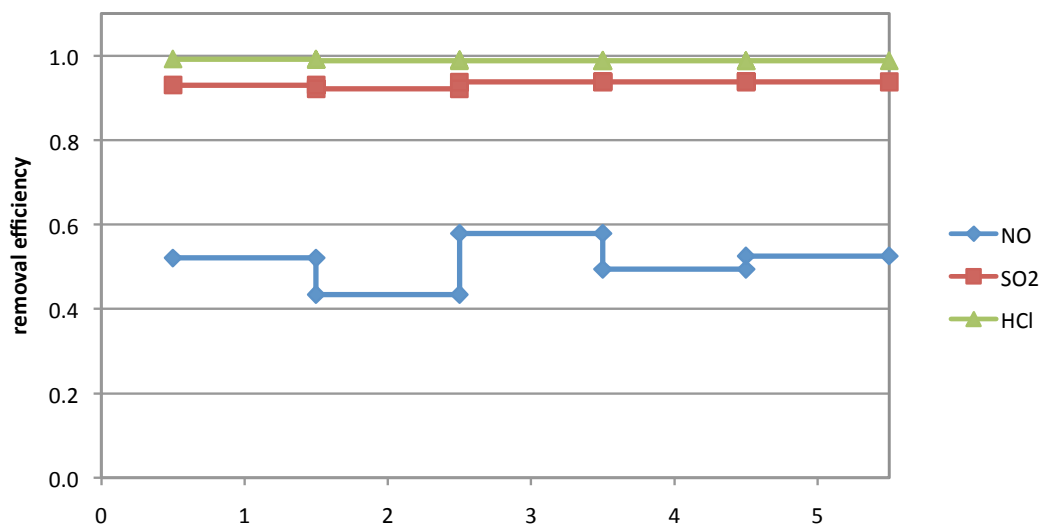
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.0	13.9	13.8	13.7	13.8	vol%
H ₂ O	19.9	20.0	20.3	20.8	20.8	vol%
NO	33.4	31.8	30.9	31.6	29.5	mg/Nm ³ (dry)
NO ₂	1.9	2.2	2.0	1.9	2.5	mg/Nm ³ (dry)
SO ₂	120.8	127.5	143.2	138.3	138.9	mg/Nm ³ (dry)
HCl	12.5	10.9	9.9	8.9	8.4	mg/Nm ³ (dry)
CO	5.4	11	5.4	4.7	4.6	mg/Nm ³ (dry)

Flue gas analysis - Clean gas:

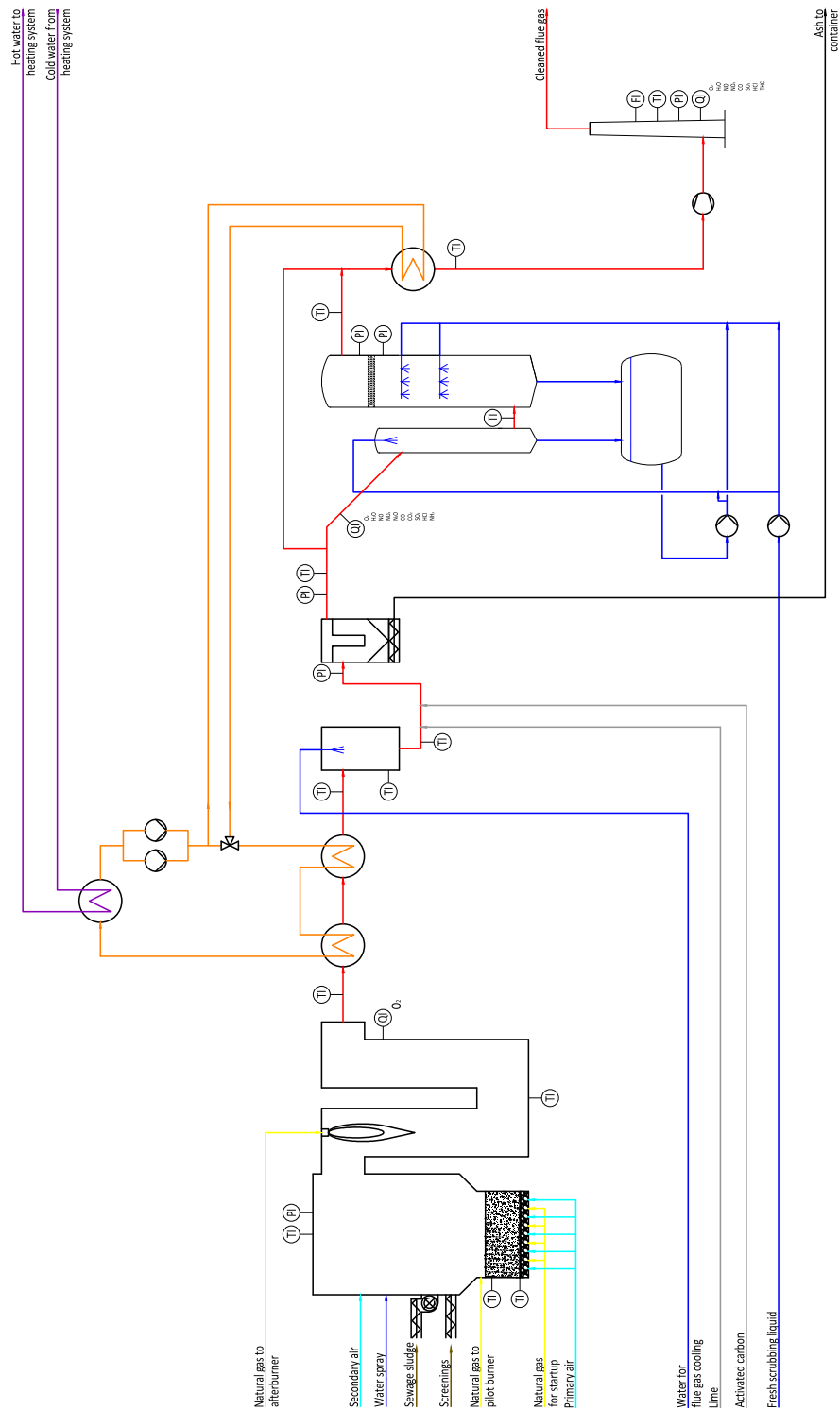
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	
O ₂	14.7	14.7	14.6	14.4	14.5	vol%
H ₂ O	9.3	9.7	9.8	10.1	9.8	vol%
NO	16.0	18.0	13.0	16.0	14.0	mg/Nm ³ (dry)
NO ₂	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
SO ₂	8.4	10	8.9	8.5	8.6	mg/Nm ³ (dry)
HCl	0.1	0.1	0.1	0.1	0.1	mg/Nm ³ (dry)
CO	4	9.2	3.4	2.8	2.4	mg/Nm ³ (dry)

Removal efficiency:

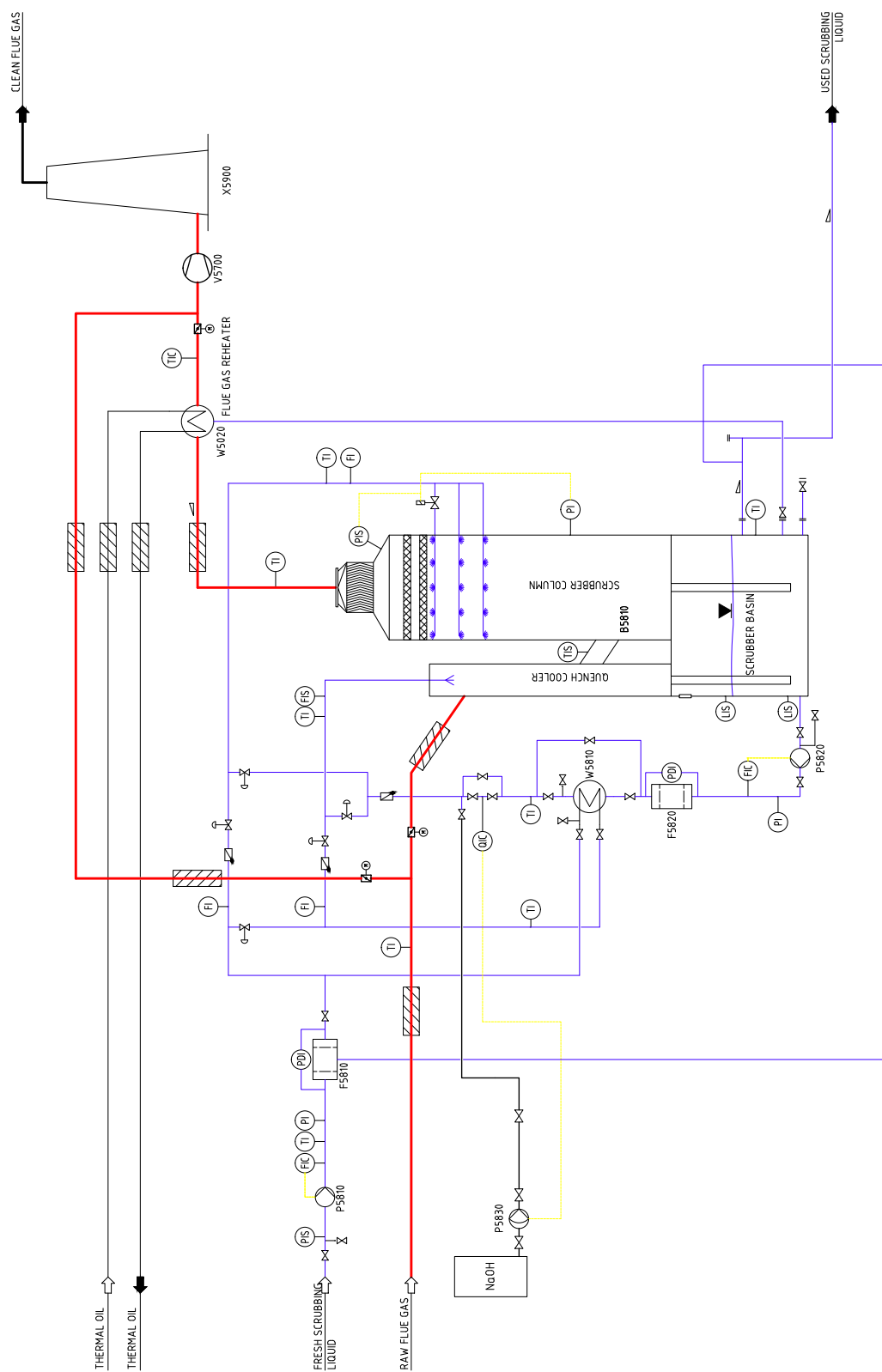
	HMW 1	HMW 2	HMW 3	HMW 4	HMW 5	AV
NO	0.521	0.434	0.579	0.494	0.525	0.510
SO ₂	0.930	0.922	0.938	0.939	0.938	0.934
HCl	0.992	0.991	0.990	0.989	0.988	0.990



Appendix C. Flowsheet of the incineration plant and the scrubber system



Appendix D. P&ID of the scrubber system



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Education

- 2010 - current **Vienna University of Technology**
Doctoral Program for Chemical and Process Engineering
PhD Thesis: *"Development of a flue gas scrubber system for sewage sludge incineration with biological scrubbing liquid regeneration"*
- 2005 - 2010 **Vienna University of Technology**
Diploma Program for Chemical and Process Engineering
Focus: Apparatus, Plant and Process Engineering
Diploma Thesis: *"Co-gasification of sewage sludge and wood in a dual fluidized bed gasifier"*
- 1998 - 2003 **Technical College for Mechanical Engineering and Automation Wiener Neustadt**
Graduation Project: *"Packaging of medical swaps"*

Work experience

- 2010 - current **Kalogeo Anlagenbau GmbH, Leobersdorf, Austria**
Process engineer for design of plants for thermal utilization of sewage sludge
- 2007 - 2009 **Vienna University of Technology, Faculty of Mathematics and Geoinformation**
Tutor for mathematics
- 2004 - 2005 **Lohmann & Rauscher GmbH, Schönau/Tr., Austria**
Technician and Project Manager for machine design and development