

Minimization of SO₂ Emissions at ADGAS (Das Island, UAE): I- Current vs. Modified Schemes

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Abstract

The objective of this work is to explore feasible technologies to minimize SO₂ emissions from ADGAS and investigate the impact of implementing such modifications on the ambient air quality at the Island. In order to achieve this, two SO₂ emission minimization schemes have been proposed; a Fuel Gas Sweetening (FGS) scheme and a Flue Gas Desulfurization (FGD) scheme. The FGS scheme involves (1) directing most of the H₂S in the fuel gas to the Sulfur Recovery Units (SRUs) where it is converted to elemental sulfur. The unconverted H₂S will be routed to the SRUs' incinerators where it is oxidized to SO₂, and (2) revamping the UGAs internals (by replacing current packing with an approved higher efficiency packing) in order to produce a sweeter fuel gas. The FGD scheme involves installation of SW-FGD units downstream the SRUs' incinerators where the SO₂ in the flue gas is scrubbed by the plant spent seawater in a dedicated packed bed absorber and converting it to harmless sulfate ions (natural constituents of the seawater) that will be safely disposed to the sea.

The FGS scheme is expected to reduce the H₂S content in the fuel gas by 94% and result in decreasing the total SO₂ emissions due to fuel gas usage by 98%. The FGD scheme is expected to reduce the SO₂ emissions due to incomplete sulfur recovery in the SRUs by 99.5%. Implementation of both schemes is expected to reduce the total SO₂ emissions by 77%. Most of the remaining SO₂ emissions (23%) are due to the continuous flaring of the flash gas in the plant. In Part II of this work, the BREEZE AERMOD Pro Software is used to predict the SO₂ Ground Level Concentration (GLC) for the current and modified SO₂ minimization schemes. Compliance with the country limits and challenge with future regulatory standards have been examined.

Keywords: Minimization of SO₂ emissions; Fuel gas sweetening; Flue gas desulfurization; Seawater scrubbing

Nomenclature

Acronyms

ADGAS	Abu Dhabi Gas Liquefaction Company
ADMA-OPCO	Abu Dhabi Marine Operating Company
BFW	Boiler Feed Water
CEPCI	Chemical Engineering Plant Cost Index
DEA	Diethanol Amine
EAD	Environmental Agency – Abu Dhabi
FEA	Federal Environmental Agency
FGD	Flue Gas Desulfurization
GLC	Ground Level Concentration
HETP	Height of Equivalent Theoretical Plate
MW	Molecular weight
PGA	Process Gas Absorber
ppm	parts per million
SW	Seawater
SRU	Sulfur Recovery Unit
UAE	United Arab Emirates
UGA	Utility Gas Absorber
US EPA	United States Environmental Protection Agency
WHB	Waste Heat Boiler
WHO	World Health Organization

Symbols

<i>a</i>	Interfacial contact area
<i>D</i>	diameter of packed bed column
<i>F</i>	Packing factor
H _{OG}	Overall height of a gas transfer unit

K _{OG,a}	Overall mass transfer coefficient based on gas phase
N _{OG}	Number of overall mass transfer units
<i>P</i>	Pressure
<i>S</i>	Solubility of SO ₂
<i>T</i>	Temperature
<i>X</i>	Seawater initial salinity
<i>X</i>	Ion concentration in seawater
<i>y_{in}</i>	Mole fraction of pollutant at absorber inlet
<i>y_{out}</i>	Mole fraction of pollutant at absorber outlet
<i>Z</i>	Packed bed total height

Introduction

Das Island is a well offshore island of the UAE in the Arabian (Persian) Gulf, 160 km of mainland and inhabited by the personnel of oil and gas industries. The main function of the Abu Dhabi Gas Liquefaction Co. (ADGAS) plant at Das Island is to process sour natural gas supplied to produce Liquefied Natural Gas (LNG), Liquefied Petroleum Gas (LPG), paraffinic naphtha and sulfur. Some of the natural gas fed to ADGAS comes directly from gas fields as non-associated gas. The processing of

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sour gases from the LNG trains at ADGAS involves SO₂ emissions that mainly result from fuel gas usage to produce steam and electricity and incomplete recovery of sulfur from acid gases. The SO₂ emissions affect the ambient air quality in the Island and this has the potential to affect the health of its residents. Throughout its history, ADGAS suffers high rates of SO₂ emissions.

Impact of SO₂ Emissions

SO₂ is a non-combustible gas that is heavier than air and its emission is associated with a wide range of health (e.g., respiratory illness) and environmental (e.g., acid rain) impacts due to the way it reacts with other substances in the air [1-3]. SO₂ emissions are greatly controlled by international and national regulations that establish limits of discharging SO₂ to the atmosphere. The main objective of such regulations is to enhance ambient air quality and prevent the environmental effects of the SO₂ emissions.

SO₂ emissions have global, regional and local impacts. Globally, the SO₂ emitted has the potential to travel in any direction for hundreds of kilometers depending on climate conditions. SO₂ is relatively stable in the atmosphere and has the ability to travel as far as 1000 km [1]. Regionally, SO₂ emissions from ADGAS contribute to the high pollution levels in the western region of the Emirate of Abu Dhabi (UAE). The oil and gas industries are the main source of air pollution there and SO₂ air pollution is dominating in that Region [4]. Locally, SO₂ emissions from ADGAS affect the ambient air quality in Das Island. The high SO₂ emissions there usually lead to air quality deterioration which might have health effects on the residents of the island (including ADGAS employees).

The magnitude of the impact of SO₂ emissions necessitates the need to research all possible means to combat such impacts. This investigation should have a significant impact on the SO₂ emissions from the ADGAS Plant in Das Island. The ultimate goal is to minimize the impact of the SO₂ emissions at all levels and establish better environmental and occupational health for the residents of the Das Island.

Constraints on the Minimization of SO₂ Emissions at ADGAS

Several constraints have been considered in the process of identifying the most suitable technologies to minimize the SO₂ emissions from the ADGAS plant. These include:

- ADGAS plant is located in a remote area, nearly 180 km from the city of Abu Dhabi. Thus, any minimization technology must not depend on continuous resources supply from outside the Das Island.
- Any SO₂ minimization scheme must take into consideration the limited area of the ADGAS plant which is crowded with processing units and equipment; therefore, it is preferable that any proposed plant modification exposes the minimum footprint in the Island.

Sources of SO₂ emissions within the ADGAS plant

H₂S and CO₂ are the main contaminants in the natural gas feed to ADGAS Liquefied Natural Gas(LNG) trains. The natural gas feed as received contains (2.5-5.5) mol% H₂S and (4.5-6.5) mol% CO₂. H₂S is a very toxic and corrosive gas that freezes at very low temperatures. CO₂ is an impurity that also freezes at low temperatures and would, therefore, block the cryogenic sections of the plant where the natural gas feed is cooled to -160°C. Therefore, both gases must be removed

in order to meet the cryogenic liquefaction requirements and the end-products' specifications.

ADGAS operates three LNG processing trains. Trains 1 and 2 are identical, both process associated and non-associated gas and each is designed to produce 180 ton/h LNG. Train 3 processes non-associated gas direct from the gas field and is designed to produce 384 ton/h LNG.

The processing of the sour gases at ADGAS involves Hi-Pure Benfield' units (certified by UOB) that utilize a two-stage sweetening process: The gas feed is treated in the first absorber with hot potassium carbonate solution which reduces the CO₂ and H₂S contents to 2,000 ppm and 800 ppm, respectively. In the second absorber it is treated with Diethanol Amine Solution (DEA) which reduces H₂S to <4 ppm and CO₂ to <50 ppm. The removed acid gases (H₂S and CO₂) are then sent to Sulfur Recovery Units (SRUs) where molten sulfur is recovered from the H₂S by the Super Claus' Process.

The main process sources of SO₂ emissions at ADGAS can be classified as flare and non-flare sources. Currently, these respectively constitute 1706 ton/yr (10.6%) from flare sources and 14,385 ton/yr (89.4%) from non-flare sources. The flare-SO₂ emission sources mainly include SO₂ emissions from pilot and purge usage and from flaring of sour gases in the case of emergency. The SO₂ emission sources from flares are considered intermittent except for the continuous flaring of flash gases from the sweetening plants. The non-flare SO₂ emission sources include SO₂ emissions from the fuel gas usage (e.g., in boilers, fired heaters, and incinerators) and from the incomplete sulfur recovery units.

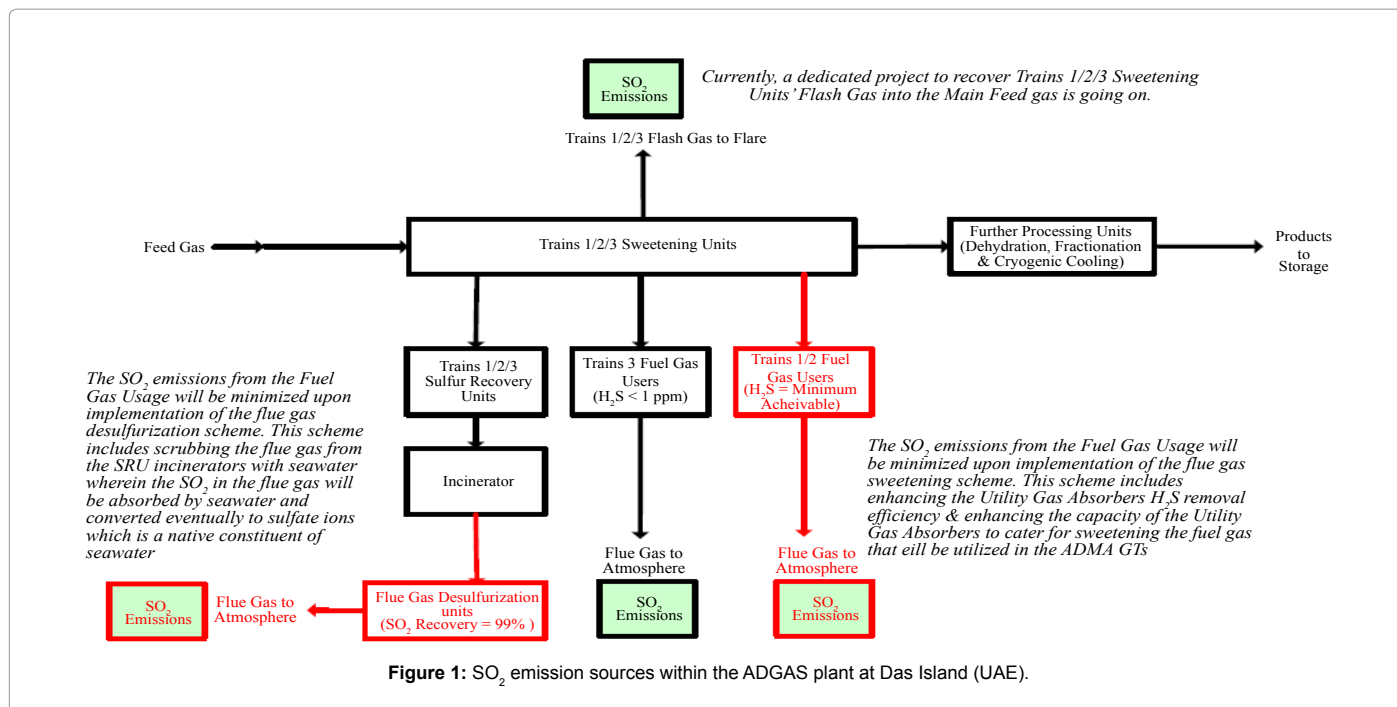
The fuel gas usage within the ADGAS plant contributes to about 20% of the plant total SO₂ emissions due to its H₂S containment. This is a concern only for Trains 1 and 2 which utilize fuel gas with 1200 ppm H₂S produced in the UGAs of Trains 1 and 2. It is also a concern for the ADMA-Gas Turbines (ADMA-GTs) which utilize untreated fuel gas with an H₂S content of 2.2 mol%. i.e., the fuel gas usage in the ADMA-GTs is one of the major sources of the SO₂ emissions within the ADGAS plant. There is no concern for Train 3 UGA since it utilizes very sweet gas with H₂S <5 ppm.

The tail gas off the SRUs is incinerated and the resultant flue gas is vented to the atmosphere through the incinerators stacks. The contribution of the vented flue gas accounts for about 70% of total SO₂ emissions from the ADGAS plant. Thus, the SRUs are considered the major contributors to SO₂ emissions within the company. The various SO₂ emission sources within the ADGAS plant are presented in Figure 1.

Previous investigations on SO₂ emission minimization at ADGAS

Atkins [5] investigated aerial emissions on Das Island to determine compliance of aerial emission sources with environmental regulations and assessed the impact of these emissions on ambient air quality and concluded that anyone spending long periods on Das Island could experience elevated health risks from exposure to SO₂. Atkins used US EPA ISCST3 for modeling the air dispersion. Atkins [5] recommended enhancing the SRUs from Claus to

Super Claus technology and reducing the sulfur content of the fuel gas. Al-Nuaimi [6] followed the Atkins recommendation to upgrade the SRUs from Claus to Super Claus process which will reduce the SO₂ ground-level concentration in Das Island by about 30%. In his work, Al Nuaimi used BREEZE ISC3 for air dispersion modeling.



Packing Type	Void Fraction, ϵ	HETP* (m)	ΔP (mm H ₂ O/m packing) **
1" HYPAK®	0.96	0.40	11.67
IMTP®#40	0.98	0.40	5.83

*HETP is for a standard test of iso-octane/toluene at 98.4 kPa in a packed tower (15.2 in diameter and 10 ft height), capacity factor = 0.20 ft/s.

** ΔP is for a standard test of Air/Water system (inlet gas rate = 1.10 kg/(m².s), liquid rate = 6.8 kg/(m².s) in packed bed (30 in diameter and 10 ft height).

Table 1: Random packing characteristics: IMTP® #40 [7] vs. 1" HYPAK® [8]. Further details about IMTP® #40 can be found elsewhere [7].

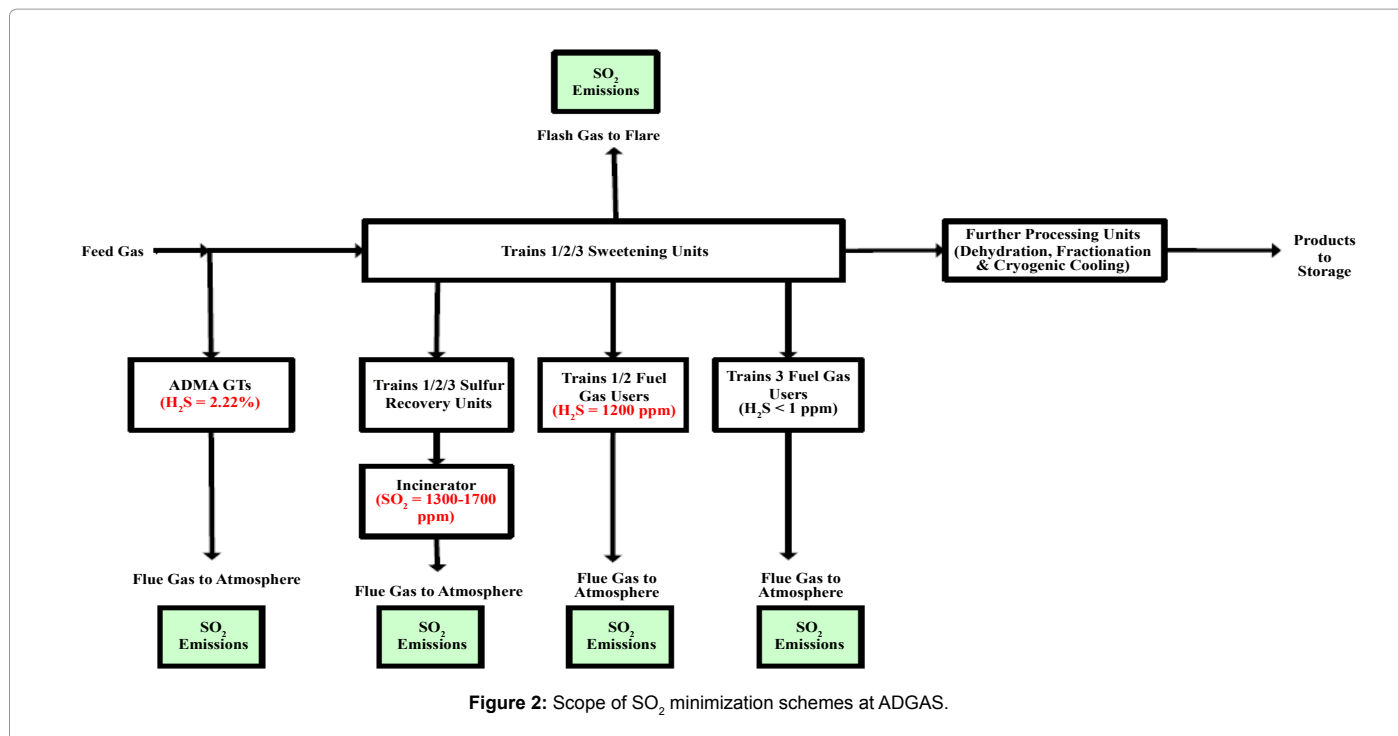
In the 2001/2002 revamp of Trains 1 and 2 Acid Gas Removal Units, ADGAS replaced the old HYPAK® random packing of the Hi-Pure Benfield' units with Koch-Glitsch IMTP' #40 random packing. The IMTP' packing was selected based on its improved hydraulics, mass transfer performance and its lower pressure drop [7] (Table 1). The performance of the Process Gas Absorber (PGA), which processes the main gas feed to Trains 1 and 2, has been enhanced in exactly the same manner as the UGAs. That is, the outlet H₂S was reduced from 1500 ppm to 450 ppm (70% reduction of the design value). Based on these results, it is proposed here to replace the current 1" HYPAK® random packing of the UGAs of Trains 1 and 2 with IMTP® #40 random packing [8]. In 2005, ADGAS commissioned Shell Global Solutions (SGS) to carry out a study with the objective of identifying measures to reduce emissions from the ADGAS LNG plant at Das Island. SGS pointed out that the poor ADGAS SO₂ emission performance is due to the unreliability of the SRUs. Since flared gases might contain as high as 15 mol% H₂S and in order to reduce the SO₂ emissions, SGS proposed the recovery of the flash gases from the sweetening plant and use them as fuel gas in the SRUs' incinerators [9]. In 2006, a process optimization at ADGAS has led to improving the sulfur recovery efficiency to > 99% by adding a Super Claus process in all the three LNG trains.

In 2007, ADGAS set itself an aggressive plan for accelerated implementation of flaring and emission reductions. ADGAS planning to reduce total gas flaring from 15 MMSCFD in 2007 to 1 MMSCFD by 2010 through four schemes: major flaring reductions, energy efficiency improvement, SRU reliability and integrity improvement, and other flaring reductions.

Proposed minimization schemes for the SO₂ emissions at ADGAS

The selection of optimum gas sweetening method is a hard task and depends on various factors [10] such as type and concentration of contaminants in the gas, desired degree of contaminant removal, required selectivity of the acid gas removal, temperature, pressure, volume and composition of the gas to be processed, CO₂/H₂S ratio in the gas, and the desirability of sulfur recovery due to process or environmental issues [11].

In order to minimize the resultant SO₂ emissions from fuel gas usage, it is also proposed here to improve the sweetening efficiencies of Trains 1 and 2 UGAs and sweeten the fuel gas supplied to the ADMA-GTs. This can be done through rejuvenation of the UGAs (i.e., the main fuel gas producers) through revamping the UGAs internals, and at the same time utilizing these absorbers to cater for the ADMA-GTs' fuel gas demand. This can be done by replacing the UGAs' current packing with a higher efficiency packing. This is expected to result in the enhancement of the mass transfer efficiency between the fuel gas and the absorbing agent in the UGAs. It is also proposed here to install Seawater-Flue Gas Desulfurization (SW-FGD) units downstream the SRUs' incinerators. In these units, the SO₂ in the flue gas will be scrubbed by the plant spent seawater in a dedicated packed bed absorber. In this case, the SO₂ in the flue gas will be converted to sulfate ions which are natural constituents of the seawater. The FGD system in this case must have the capability of recovering not less than 99.5% of the SO₂ emitted from the SRU incinerators. The recovered SO₂ must be disposed in the most environmentally-friendly manner. A schematic diagram of current and proposed SO₂ emission minimization schemes are shown in Figure 2.



Characteristic	Acid gas feed to UGAs	Sweet gas from UGAs	Lean carbonate solution	Rich carbonate solution
H ₂ S (mol %)	2.22	0.0072	-	0.42
CO ₂ (mol %)	3.67	0.0094	-	0.70
K ₂ CO ₃	-	-	5.41	5.35
Molar flow rate (kmol/h)	2,442.8	2,299.3	9,430.66	9,574.22
Mass flow rate (ton/h)	49.03	43.14	234.95	240.70
Pressure (barg)	15.45	15.5	24	15.9
T emp. (°C)	69	85	121	93

Table 2: Material Balance on the Acid Gas Feed Sweetening in the UGAs of Trains 1 & 2.

After choosing the minimization schemes, the following has been made

- Study the impact of the proposed SO₂ emission minimization schemes on the ADGAS plant operations. Such study will be exclusive of the associated utility requirements, real plant tie-ins, associated control system, and off-site (e.g., flares) connections
- Design of all process equipment for the proposed schemes to fit real plant conditions (e.g., feed composition, flow, temperature, pressure, etc.) for both fuel gas sweetening and flue gas desulfurization schemes.
- Conduct a cost-analysis study to estimate the cost associated with the selection of the minimization schemes of SO₂ emissions.
- Predict the impact of the proposed SO₂ emission minimization schemes on the Das Island air quality. The results of this part will be presented in Part II of this work.

Impact of the proposed SO₂ emission minimization schemes on plant operations

The objective of proposed SO₂ emission minimization schemes

is to minimize ADGAS SO₂ emissions due to fuel gas usage. The implementation of the proposed fuel gas sweetening scheme requires increasing the entry temperature of the fuel gas feed to the UGAs, replacing of the UGAs' packings from HYPAK to IMTP, utilizing UGAs to sweeten the ADMA-GTs' fuel gas, and modifying the treatment capacity of the UGAs (i.e., circulation rates of fuel gas feed and lean carbonate solution). On the other hand, implementing the SW-FGD scheme will be associated with the determination of seawater and boiler feed water (BFW) requirements, estimation of the steam produced in the waste heat boilers and its utilization in the ADGAS plant, and determination of the neutralization and oxidation requirements of the spent seawater before being disposed to sea.

Heating fuel gas feed to trains 1 and 2 UGAs: One aspect of the fuel gas sweetening scheme is the way of establishing the same process conditions in Trains 1 and 2 UGAs (to be similar to those of Trains 1 and 2 PGAs) in order to promote the maximum H₂S removal efficiency. The fuel gas feed to the UGAs has to be heated to 69°C (= gas feed entry temperature to PGAs) using Low Pressure Steam (LS). The required steam can be supplied from the mean-steam distribution system of Trains 1 and 2, taking into account the steam generated in the waste heat boilers upon implementing the flue gas desulfurization scheme. Table 2 shows basic material balance on the acid gas feed sweetening process in the UGAs of Trains 1 and 2 using hot K₂CO₃ solution.

a. Process Gas Absorber ($T = 69\text{ }^{\circ}\text{C}$; $P = 53\text{ barg}$)

Parameter	Y _{in}	Y _{out}	NOG	Z	HOG	KOG.a	Carrying capacity
Unit	Mole fraction	Mole fraction	Transfer units	m	m	g-mol/h.m ² .Pa	kmol /kmol carbonate solution
H ₂ S	0.0316	0.000426	4.34	13.39	3.09	0.393	0.00748
CO ₂	0.0418	0.000448	4.53	13.39	2.95	0.410	0.00992

b. Utility Gas Absorber ($T = 69\text{ }^{\circ}\text{C}$; $P = 15.45\text{ barg}$)

Parameter	Y _{in}	Y _{out}	NOG	Z	HOG	KOG.a	Carrying capacity
Unit	Mole fraction	Mole fraction	Transfer units	m	m	g-mol/h.m ² .Pa	kmol /kmol carbonate solution
H ₂ S	0.0222	7.220E-05	5.75	19.48	3.39	0.393	0.00573*
CO ₂	0.0367	9.350E-05	6.01	19.48	3.24	0.410	0.00949**

* 130.42% excess capacity based on PGA removal efficiency; ** 104.57% excess capacity based on PGA removal efficiency.

Table 3: Mass Transfer Parameters and Carrying Capacity of PGA and UGA of Trains 1 and 2.

Packing replacements: The determination of the H₂S and CO₂ concentrations from the outlet of the UGAs of Trains 1 and 2 under the proposed conditions requires the determination of the overall mass transfer coefficient ($K_{OG.a}$) at similar conditions (i.e., the PGAs system of Trains 1 and 2). Table 3 shows the mass transfer parameters and carrying capacity of the PGA and the UGA of Trains 1 and 2.

The proposed fuel gas sweetening scheme requires replacing the current Trains 1 and 2 UGAs HYPAK[®] packing to the higher efficiency IMTP[®] packing, catering UGAs to supply the required fuel gas to ADMA-Gas Turbines, and heating the fuel gas feed to the UGAs to establish the same conditions of the PGAs of Trains 1 and 2.

IMTP[®] has less H_{OG} than HYPAK[®]. The fact that the IMTP[®] has higher mass transfer coefficient (K_{OG}) leads to decrease the height of the mass transfer unit (H_{OG}) and thus provide more transfer units (N_{OG}) for the same packing height (Z). The IMTP[®] #40 packing has less HOG than the HYPAK[®] packing. That is, the utilization of IMTP[®] packing will enhance the H₂S-removal efficiency of the carbonate solution as long as there is a room in the carbonate solution for H₂S absorption.

The packing replacement in this work has resulted in 48.87% reduction in H_{OG} (from 6.63 to 3.39 m) and consequently increased the number of available transfer units (N_{OG}) by 95.58% (from 2.64 to 5.75); i.e., almost doubled the carbonate solution removal capacity. Ultimately, the H₂S in the semi-sweet fuel gas treated in the UGAs was reduced by 93.98% (from 1200 to 72 ppm). Also, the proposed fuel gas sweetening scheme has resulted in 94.75% reduction in the H₂S content in the fuel gas feed to the boilers, 94% reduction H₂S content in fuel gas to all fuel gas users of Trains 1 & 2, and 99.67% reduction in the H₂S content in the fuel gas to ADMA-GTs.

Catering for the ADMA-GTs fuel gas demand and fuel gas and carbonate solution circulation rates: In order to cater for the ADMA-GTs fuel gas demand requires increasing Trains 1 and 2 UGAs treatment capacity by 18% (from 2,071 to 2,443 k mol/h). This will be accompanied with an increase in the lean carbonate solution circulation rate by about 20% (from 154 to 185 m³/h). The original Trains 1 and 2 design capacities for the fuel gas feed and carbonate solution circulation rates are 3,401 kmol/h and 303 m³/h, respectively. Thus, the rates of both fuel gas feed and carbonate solution are still within the design envelope of the UGAs. The current fuel gas supply to the ADMA-GTs is about 598 kmol/h. Therefore, upon utilization of the UGAs to supply the required fuel gas to the ADMA-GTs, the UGAs load will increase to about 2,669 kmol/h, which is still within the UGAs design capacity.

Calculations required for the proposed fuel gas sweetening scheme

As of the current conditions, the high-pressure gas feed at 53 barg is let down through a pressure control valve to 15.6 barg prior to being split into 2 main streams; one is directed to the UGAs of Trains 1 and 2 and the other to ADMA-GTs. The first is sweetened in the UGAs and fed to fuel gas system while the other is directly utilized in the ADMA-GTs as a fuel gas without any treatment.

The target of the modified fuel gas sweetening scheme is to achieve minimum H₂S concentration at the UGAs outlet. This requires routing all the partially depressurized gases to the UGAs of Trains 1 and 2 and maximizing the UGAs efficiency through utilization of the high performance IMTP[®] packing; this specification is limited due to the fact that all of the mass transfer parameters of the UGAs (including the packed bed diameter and height) are fixed. The implementation of the modified fuel gas sweetening scheme includes the determination of H₂S and CO₂ concentrations at the UGAs outlets as well as the required fuel gas treating capacities in these UGAs. The data required for the material balances around the UGAs and the PGAs include the gas feed and the carbonate solution properties. The gas feed data of PGAs and UGAs of Trains 1 and 2 were retrieved from real plant ADGAS Process History Data (PHD). The average data for the gas feed composition, flow, temperature and pressure for a test period of three months (Dec 2005-Feb 2006) were collected. The operation of the plant during this test period was steady as the plant did not experience any upsets or sudden process trips. The carbonate solution composition data were collected from ADGAS Laboratory Daily Logs representing the results of the routine sampling of the various solutions used in the trains' sweetening plants. The flow, pressure and temperature of the carbonate solution were taken from PHD for the same test period.

The characteristics of the fuel gas hot K₂CO₃ solution before and after the UGAs and PGAs of Trains 1 and 2 along with fuel gas supply and demand of Trains 1 and 2 under current conditions are available elsewhere.

The heat input for each fuel gas user (e.g., boilers, fired heaters, and incinerators) has been calculated. The heat input for each train is then used as a basis for the determination of the required fuel gas flow (that will deliver the same heat input to that train) once the fuel gas composition is changed post implementing the proposed scheme.

The H₂S content in the sour gas feed has been reduced from 2.22 to 0.12 mol% in the UGAs and from 3.16 to 0.0426 mol% in the PGAs of Trains 1 & 2 using the hot K₂CO₃ solution. The main results are shown in Table 4.

Characteristic	UGAs	PGAs
Sour gas feed rate, kmol/h	2,071	11,719
H ₂ S in sour gas feed (mol%)	2.22	3.16
(K ₂ CO ₃ /Sour gas feed) ratio	5.04	4.10
Total H ₂ S removed (kmol/h)	43.48	365.33

Table 4: Sour Gas Treatment Results.

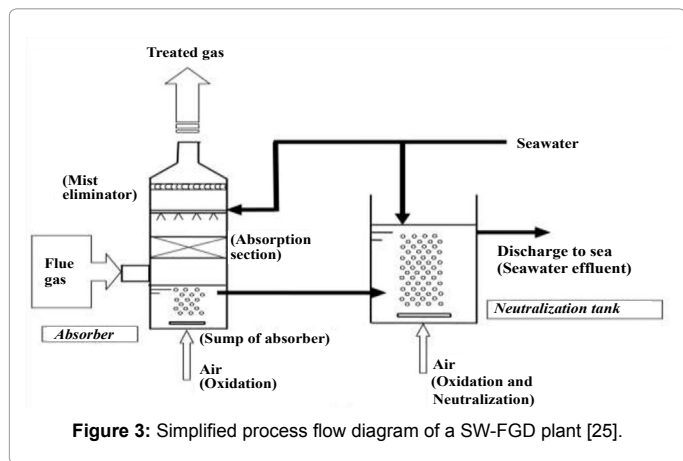


Figure 3: Simplified process flow diagram of a SW-FGD plant [25].

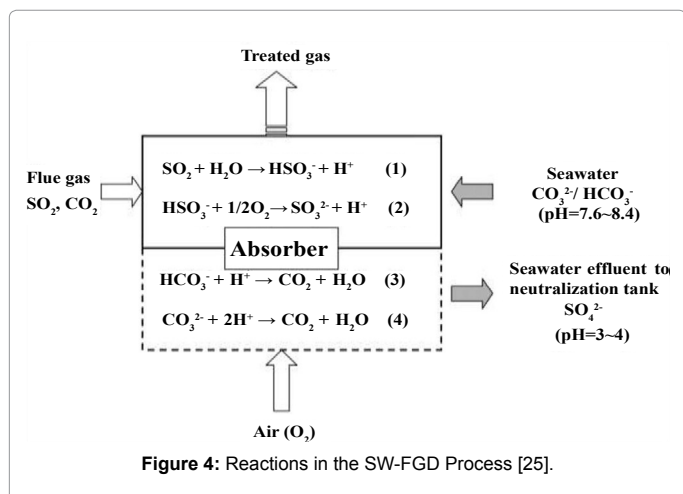


Figure 4: Reactions in the SW-FGD Process [25].

The Seawater-Flue Gas Desulfurization Scheme

The SW-FGD process is developing rapidly in fired power plants. It is a viable desulfurization option that can also be adopted in seashore-located process plants [12]. It utilizes the natural alkalinity of the process plant spent seawater to absorb the SO₂ from the flue gases. Seawater has a high capacity to absorb SO₂ due to its carbonate and bicarbonate ions [13,14]. It is a convenient absorption medium for flue gases emitted from many industrial sites compared to all other alternatives proposed in the literature [15]. The SW-FGD process is considered as a promising technology environmentally sound from an engineering point of view [16]. The seawater scrubbing is also found to be a promising technology for reducing SO₂ emissions from ships. Karle and Turner [17] modelled and analyzed the marine chemical aspects of shipboard seawater scrubbing based on a 12 MW engine burning fuel with a 3% sulphur content. Andreasen and Mayer [18] studied the mechanism of SO₂ absorption in seawater with emphasis on scrubbing of marine engine exhaust gas containing SO₂. They formulated a model and used it to predict the influence of various parameters on SO₂ absorption efficiency

such as seawater temperature, partial pressure of SO₂, seawater salinity, and seawater alkalinity.

The SW-FGD process imposes relatively low impact on the environment as the main product of this process is sulfate ions (SO₄²⁻) which is a natural constituent of seawater. The contribution of 10 this product to total sulfur in seawater is negligible. Sulfur in oceans is approximately occupying a 1.5 m layer, and if all sulfur in coal, oil and acid gas is added to the oceans, the thickness of this layer will increase by the thickness of a sheet of paper [10,19]. The increase in the total sulfate content in oceans seawater, if all the sulfur in fossil fuels is deposited in seawater, is approximated to be 905.075 g compared to a present value of 905.000 g of sulfate per ton of seawater [13].

Reliability and efficiency of SW-FGD processes

Several studies in the literature proved the reliability and the high removal efficiency of the SW-FGD systems:

- Utilization of free seawater and air along with their availability results in a significant reduction in the SW-FGD plant operating costs [20,21].
- Waste disposal is not a concern in the SW-FGD process as it does not generate any waste. This eliminates any cost to be spent on waste disposal, and leads to higher reliability [20,22].
- It has been reported that SW-FGD plants with 98.8% SO₂ removal efficiency and 98.8% availability throughout the year [10]. This indicates that SW-FGD can achieve very high removal efficiencies at higher plant availability and reliability rates.

A simplified process flow diagram for the SW-FGD system is shown in Figure 3. In the SW-FGD process, the absorption process usually takes place in a counter-current packed bed absorber in which the flue gas entering the bottom of the absorber flows upward and come into contact with the falling seawater introduced at the top of the absorber over a randomly packed bed.

The chemistry of the SW-FGD process is described through the instantaneous reactions presented in Figure 4. The SO₂ in the upcoming flue gas dissolves in the falling seawater and reacts with it to produce bisulfite HSO₃⁻ (Reaction 1) which is rapidly oxidized to sulfate (SO₄²⁻) (Reaction 2) by the oxygen contained either in the flue gas or in the seawater itself. The formed H⁺ in Reactions 1 and 2 acidifies the seawater; therefore, the seawater should be neutralized prior its discharge to the sea. In this case, the seawater neutralizes itself as it utilizes its natural alkalinity capacity (i.e., the bicarbonate, HCO₃⁻, and the carbonate, CO₃²⁻, it contains) through Reactions 3 and 4.

On the other hand, the formed HSO₃⁻ in Reaction 1 might be converted to sulfite ions, SO₃²⁻, a chemical oxygen demanding element; thus, it is essential to ensure complete oxidation of all formed HSO₃⁻ to SO₄²⁻. In general, the oxygen in the flue gas and seawater is not sufficient to completely oxidize all the formed bisulfite ions (HSO₃⁻). Therefore, forced oxidation using compressed air is employed in most SW-FGD plants [10,13,23].

Design basis of the feeds and products of the SW-FGD scheme

The basis of selecting the SW-FGD technology its application at ADGAS includes the availability of seawater. The main users of seawater within the ADGAS plant are the surface condensers of the main compressors' steam turbines. The total amount of seawater needed for cooling purposes in Trains 1 and 2 is 85,000 m³/h. The spent seawater

Characteristic	Flue Gas to Trains 1 and 2 (each)	Flue Gas to Train 3
SO ₂ (mol %)	0.17	0.13
CO ₂ (mol %)	16.71	32.8
O ₂ (mol %)	1	1
N ₂ (mol %)	57.24	45.38
Ar (mol %)	0.59	0.5
H ₂ O (mol %)	24.28	20.19
Pressure (barg)	0.01	0.01
Temperature (°C)	650	650
MW (kg/kmol)	28.43	31.39
Density (kg/m ³)	0.374	0.413
Viscosity (cP)	0.037	0.037

Table 5: Characteristics of Flue Gas Feeds to Trains 1, 2, and 3 of the SW-FGD Plant.

Composition (mg/L)	Value
Cl ⁻	25,150
SO ₄ ⁻²	3,325
HCO ₃ ⁻	92
CO ₃ ⁻²	30
Ca ⁺²	548
Mg ⁺²	1,738
Na ⁺	13,374
SiO ₂	5
Temperature (°C)	39
pH	8.4
Total dissolved solids (mg/L)	43,235
Specific Gravity @ 20 °C	1.031
Resistivity @ 20 °C (Ohms)	0.1755
Equivalent salt content ((wt/wt) %)	4.2
Available pressure (barg)	0.39
Available flow rate (m ³ /h)	42,537

Table 6: Characteristics of Seawater to the SO₂ Absorbers in the SW-FGD Unit.

Characteristic	Air	BFW
H ₂ O (mol %)	0	100
O ₂ (mol %)	21	0
N ₂ (mol %)	79	0
Pressure (barg)	1.01	4.51
Temperature (°C)	30	30
MW (kg/kmol)	28.81	18.02
Density (kg/m ³)	1.16	1004
Viscosity (cP)	0.019	0.797

Table 7: Characteristics of Oxidation Air and Boiler Feed Water (BFW).

from these two trains joins together in a common line which eventually discharges the spent cooling water into the sea.

The feeds to the proposed SW-FGD plant include the flue gas leaving the SRU plant and the seawater coming from the common seawater return header of Trains 1 and 2. The set specifications of the clean gas product from the SW-FGD plant are 8.5 ppm for Trains 1 and 2 and 6.5 ppm Train 3.

The collected data and design basis of the seawater-flue gas desulfurization scheme include the characteristics of flue gas feeds and sweater feed sweater feed to the SO₂ absorbers (Tables 5 and 6), and oxidation air and boiler feed water characteristics (Table 7).

A schematic diagram of the proposed (modified) SW-FGD plant for Trains 1, 2 and 3 is shown in Figure 5. In that diagram the flue gas from the SRUs' incinerators is cooled in a waste heat boiler (WHB) from 650°C to 130°C then compressed through a gas blower from 1

bar to 2 bars. Further cooling of the flue gas is made before entering the SO₂ absorber against the clean flue gas leaving that absorber in a gas-gas heat exchanger. The flue gas in this case is desulfurised using the ADGAS plant spent seawater where its SO₂ content is reduced from 0.17 to 0.00085 mol% for Trains 1 and 2 and from 0.13 to 0.00065 mol% for Train 3.

The flue gas, sweater, and air flowrates, steam produced and power requirements for Trains 1, 2, and 3 are listed in Table 8. Note that the seawater is currently pumped to the ADGAS plant at a rate of 42,537 ton/h which is almost double the amount needed for the three SO₂ absorbers altogether (23,256 ton/h).

Implementation of the SW-FGD scheme depends on the feeds conditions as well as on the required desulfurization efficiency. The feeds include the flue gas from the SRU incinerator and the scrubbing medium (seawater). The flue gas being at low pressure and high temperature is required to be pressurized and cooled before being introduced to the SW-FGD absorber. In addition, the seawater is required to be pressurized to the absorber pressure. The sizing of the SW-FGD absorber includes determination of the diameter and height of the packed bed. Finally, the aeration and neutralization requirement for the seawater effluent from the absorption process has to be established. The completion of these steps will result in a complete SW-FGD process package that could be implemented at the ADGAS plant. Hysys simulation has been used to build a complete material balance for the proposed SW-FGD plant but not to size the SW-FGD absorbers.

In order to reduce the load on the flue gas blower, the flue gas from the SRU incinerator has to be pre-cooled in a waste heat boiler where it loses some of its heat to produce low pressure steam. The pre-cooled flue gas is then pressurized to the absorption pressure, which usually results in a temperature increase and hence further cooling is required, e.g. in a waste heat boiler where heat is recovered by hot water to produce steam that can be used in other areas within the plant. The flue gas may be further cooled prior to being introduced to the absorber by the cold clean flue gas leaving that absorber.

The source of seawater supply to the SW-FGD plant is a slip stream

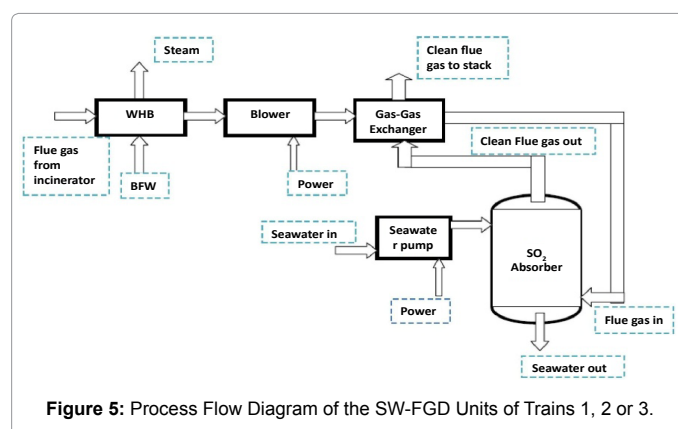


Figure 5: Process Flow Diagram of the SW-FGD Units of Trains 1, 2 or 3.

Train	Flue gas rate (ton/h)	Seawater rate (ton/h)	Oxidation air rate (kg/h)*	Steam produced (ton/h)	Flue gas blower power (kW)	Seawater pump power (kW)
1 & 2 (each)	134.62	8,957.90	606.83	51.72	10,830	205
3	97.85	5,339.81	364.01	36.39	7,089	188.4

Table 8: Flue Gas, Seawater and Oxidation Air Rates, Steam Produced and Power Requirements for Trains 1, 2, and 3 for the SW-FGD Units.

Parameter	Train 1	Train 2	Train 3
Absorber diameter (m)	5.19	5.19	4.27
Absorber height (m)	5.28	5.28	5.67

Table 9: Sizing of the packed bed SO₂ Absorbers.

Parameter	Trains 1 and 2 (each)	Train 3
Total HSO ₃ ⁻ formed (mol)	8,051	4,052
O ₂ required for aeration (mol)	4,025	2,026
O ₂ available from flue gas (mol)	47,359	31,174

Table 10: Aeration Requirements for the SW-FGD Plant.

Parameter	Train 1	Train 2	Train 3
Total H ⁺ formed (mol)	15,941.04	15,941.04	10,546.16
Seawater neutralization capacity:			
HCO ₃ ⁻ available (mol)	13,929.13	13,929.13	9,168.83
CO ₃ ²⁻ available (mol)	4,617.81	4,617.81	3,039.67
Total (mol)	18,546.94	18,546.94	12,208.50

Table 11: Neutralization Requirement for the SW-FGD Plant.

from Trains 1 and 2 common seawater return line header. A centrifugal pump is used to bring the pressure of the seawater stream to the absorber.

Relevant calculations for the SW-FGD scheme

The FGD-Absorber is the main equipment in the SW-FGD scheme where the mass transfer between the flue gas and seawater will take place. The sizing of this absorber requires establishing the seawater flow requirements and finding the absorber height and diameter. Other relevant calculations include aeration requirements for the formed bisulfite ions (HSO₃⁻) in the absorber and the neutralization requirements for the seawater used in the absorber before being disposed to the sea.

Establishing seawater flow requirement: The minimum amount of seawater required to achieve 99.5% desulfurization efficiency is established from the solubility data of SO₂ in seawater. The general practice is to operate the absorption tower at 25-50% of the minimum required absorbing solution rate [24]. Rodri'guez-Sevilla et al. [25] studied the absorption equilibria of dilute SO₂ in seawater. The correlation of Al-Enezi et al. [26] has been adopted in this work for the solubility of SO₂ in the Arabian Gulf seawater as a function of seawater initial salinity and temperature (Eq. 1). This correlation is valid for a temperature range of 10 to 40°C, a SO₂ partial pressure of 0.0224 kPa, pH = 2.5 to 2.6; and a salinity range of 0 to 40,000 ppm.

$$S = a_1 + a_2T + a_3X + a_4T^2 + a_5X^2 \quad (1)$$

Where S is the solubility of SO₂ in seawater (mol/L), T is seawater inlet temperature (°C), X is seawater initial salinity (ppm), a₁ = 1.017702×10⁻³, a₂ = -2.71138×10⁻⁵, a₃ = 1.03347×10⁻⁸, a₄ = 2.43198×10⁻⁷, and a₅ = 1.24418×10⁻¹³.

The required seawater flow rate to the prospective SW-FGD units is presented in Table 8. This flow rate is based on a calculated SO₂ solubility of 0.984 moles of SO₂ per m³ of seawater, and a required 99.5% SO₂ removal efficiency.

Diameter of the SW-FGD packed bed absorber: The procedure for sizing packed columns is extensively discussed in literature. The procedure adopted in this work is that of Manyele [24]. A 3-in Super Intalox[®] ceramic packing was chosen here based on the fact that this type and size of packing has a low packing factor and hence is expected to result in an optimum packed bed diameter. On the other hand, the

determination of the overall height of a gas transfer unit (H_{OG}) and the number of overall mass transfer units (N_{OG}) can be achieved either experimentally or calculated from a variety of methods in the literature.

Baaliña [27] developed a laboratory plant where flue gas is scrubbed by seawater pumped directly from the sea. The scrubbing process took place in a counter-current 2×0.25 m cylindrical stainless steel scrubber packed with 3/4 inch Intalox[®] ceramic saddles to a height of 0.92 m. The experimentally determined H_{OG} is equal to 0.15 m at the following conditions (y_{in} = 1.7, L/G = 69.7, desulfurization efficiency = 99.8%). Since the conditions of this work are very close to those of Baaliña et al. (i.e., y_{in} = 1.7, L/G = 66.5, desulfurization efficiency = 99.5%), the Baaliña [27] H_{OG} value was used here to determine the mass transfer coefficient of the flue gas-seawater system. Details of these calculations and input data are presented elsewhere [28]. Table 9 shows the calculated diameter and height of the packed bed SO₂ absorber to achieve the projected SO₂ removal efficiency for Trains 1, 2 and 3.

Aeration requirements: The required O₂ to achieve complete oxidation of the formed bisulfite ions in the absorption process is calculated for the Reactions 1 and 2 as follows (Figure 4):

According to Reaction 1,

$$HSO_3^- \text{ formed} \left[\frac{\text{mol}}{\text{h}} \right] = y_{SO_2} \times \text{Total Flue Gas Flowrate} \left[\frac{\text{mol}}{\text{h}} \right]$$

According to Reaction 2, O₂ required [mol] = 0.5 X HSO₃⁻ formed [mol]

In fact, the amount of O₂ in the flue gas is sufficient to completely oxidize all the formed bisulfite ions (HSO₃⁻). Thus, it is not required to have forced oxidation for the proposed SW-FGD plants. Anyway, the O₂ required for aeration in the SW-FGD units of Trains 1, 2 and 3 is presented in Table 10. The air required for oxidation in the SW-FGD units is presented in Table 11.

Neutralization requirements: The formation of H⁺ ions in Reactions 1 and 2 (Figure 4) necessitates the need to neutralize the seawater effluent from the absorber. Usually, the seawater has the capability to neutralize itself depending on its bicarbonate and carbonate ions containment. Karle and Turner [17] made the calculations for the volumes of water required for (i) uptake of Sox (the scrubbing process), (ii) dilution of the scrubbing water to achieve a pH of 6.5, (iii) further dilution to achieve a pH within 0.2 units of that in the ambient water, and no more than a 1% reduction in the dissolved oxygen concentration.

The produced H⁺ ions due to scrubbing reactions (Reactions 1 and 2) are calculated as follows:

$$H^+ \text{ produced in Reaction 1} \left[\frac{\text{mol}}{\text{h}} \right] = \text{Total SO}_2 \text{ Absorbed} = y_{SO_2} \times \text{Total Flue Gas Flow} \left[\frac{\text{mol}}{\text{h}} \right] \quad (2)$$

$$H^+ \text{ produced in Reaction 2} \left[\frac{\text{mol}}{\text{h}} \right] = \text{HSO}_3^- \text{ formed} \left[\frac{\text{mol}}{\text{h}} \right] = y_{SO_2} \times \text{Total Flue Gas Flow} \left[\frac{\text{mol}}{\text{h}} \right] \quad (3)$$

Total H⁺ produced = H⁺ produced in Reaction 1 + H⁺ produced in Reaction 2 (4)

Where y_{SO₂} is the mole fraction of SO₂ in the flue gas.

The seawater neutralization capacities in terms of bicarbonate (HCO₃⁻) and carbonate (CO₃²⁻) ions (Reactions 3 and 4) are calculated as follows:

$$HCO_3^- \left[\frac{\text{g-mol}}{\text{h}} \right] = x_{HCO_3} \left(\frac{\text{mg}}{\text{l}} \right) \left(\frac{1\text{g}}{1000\text{mg}} \right) \left(\frac{1}{MW_{HCO_3}} \right) \left(\frac{1000\text{l}}{1\text{m}^3} \right) \times \text{Seawater flow} \left(\frac{\text{m}^3}{\text{h}} \right) \quad (5)$$

Equipment	Quantity	Purchased Cost (\$)	Bare Module Cost (\$)	Total Module Cost (\$)
Train 1 and 2 Flue Gas Blowers	2	2,700,000	7,380,000	11,180,000
Train 3 Flue Gas Blower	1	1,020,000	2,790,000	4,230,000
Train 1 and 2 Flue Gas Blower Drivers	2	850,000	2,980,000	4,500,000
Train 3 Flue Gas Blower Driver	1	398,000	1,390,000	2,110,000
Train 1, 2 and 3 Waste Heat Boilers	3	81,300	274,200	323,400
Train 1, 2 and 3 Gas-Gas Heaters	3	81,300	267,900	316,200
Train 1 and 2 Seawater Pumps	2	111,000	442,000	522,000
Train 3 Seawater Pump	1	51,900	207,000	244,000
Train 1 and 2 FGD Absorbers	2	486,000	3,300,000	3,900,000
Train 3 FGD Absorber	1	170,000	1,071,000	1,264,000
Train 1 and 2 UGA Feed Heaters	2	54,400	180,400	274,000
Total		6,003,900	20,282,500	28,863,600

Table 12: Estimated capital cost for the equipment needed for the proposed SW-FGD scheme using CAPCOST [29]. CEPCI = 509.1 [31].

Equipment	Quantity	Utility used	Annual Utility Cost (\$)
Train 1 and 2 Flue Gas Blower Drivers	2	Medium pressure steam	3,214,000
Train 3 Flue Gas Blower Driver	1	Medium pressure steam	1,066,000
Train 1 and 2 Seawater Pumps	2	Electricity	242,000
Train 3 Seawater Pump	1	Electricity	111,000
Train 1 and 2 UGA Feed Heaters	2	Low pressure steam	6320
		Total	4,639,320

Table 13: Annual utilities cost for the proposed SW-FGD Plant using CAPCOST.

$$CO_3^{-2} \left[\frac{g-mol}{h} \right] = x_{CO_2} \left(\frac{mg}{l} \right) \left(\frac{1g}{1000mg} \right) \left(\frac{1}{MW_{CO_2}} \right) \left(\frac{1000l}{1m^3} \right) \times Seawater\ flow \left(\frac{m^3}{h} \right) \quad (6)$$

Where x is ion concentration in seawater (mg/l) and MW is ion molecular weight (g/g-mol).

Further neutralization of the spent seawater from the SW-FGD plant is achieved through returning the spent seawater to common seawater return header of Trains 1 and 2. Upon such, the spent seawater will be mixed with fresh seawater before being discharged into the sea. The mixing of the spent cooling water with fresh seawater prior discharge will enhance the properties of the spent seawater. The neutralization requirements for Trains 1, 2 and 3 FGDs are presented in Table 11

Cost estimation of SO₂ minimization schemes

The drive of this work is purely environmental; reduce the global, regional and local impacts of SO₂ emissions from the ADGAS plant, and ensure compliance of all SO₂ emission sources within the ADGAS plant to the standards of the United Arab Emirates Federal Environmental Agency (UAE-FEA). The cost associated with the selected SO₂ minimization schemes includes the capital cost for the proposed SW-FGD units (Table 12), its associated annual utility and operating cost (Table 13). The cost of operating labor is \$750,000 per year as estimated by CAPCOST [29]. The SW-FGD units will produce 140 ton/h medium pressure steam (T = 220°C; P = 9.41 barg). The price of medium pressure steam is \$ 3.34/ton; this price is based on actual steam turbine for fixed process steam loads [30]. The price of the produced steam will be \$2,973,531. The production of medium pressure steam will result in reducing the demand of medium and low pressure steams within ADGAS. The direct result is eliminating the need to produce 140 ton/h of high pressure steam in ADGAS boilers. For a high pressure steam price of \$5.78/ton [31], the saving of the produced medium pressure steam from the proposed flue gas desulfurization plants is \$5,145,811 per year. The total income and savings will be \$8,119,343 per year. As a result, this will cover the associated annual utilities and operating cost of \$5,389,320 per year. The net yearly income is \$2,730,022. The payback period of this will be 10.57 years.

In addition, the SO₂ minimization schemes will result in a positive impact and savings that will be gained from the achievement of an improved air quality, and consequently a better environment, which will lead to a better occupational health impact and less risk to the health of the workers. The later will lead to an increase in the productivity of the workers and enhancing the performance of the company they work in.

Conclusions and Recommendations

This work focuses on quantifying SO₂ emissions from ADGAS following an approach centered on equipping the SRUs with FGD units, and maximizing H₂S diversion from fuel gas to SRUs. Upon such an approach, most of the H₂S in the fuel gas will be directed to the SRUs, and will be converted to sulfur product. Unconverted H₂S will be routed to the SRUs incinerator where it is oxidized to SO₂ and then directed to the FGD units where it is scrubbed by the desulfurization solution and converted into the harmless products that are disposed safely and economically. As a result, the SO₂ emissions will be largely reduced. This approach guarantees optimum SO₂ minimization; thus, minimizing the global, regional and local impacts of ADGAS SO₂ emissions.

Two minimization schemes have been proposed in this work; fuel gas sweetening scheme and flue gas desulfurization scheme. (1) In the fuel gas sweetening schemes, Trains 1 and 2 UGA H₂S removal efficiency will be maximized through replacement of the packing with a higher efficiency packing and the catering of the required fuel gas by ADMA-GTs which at the current conditions utilize untreated gas. (2) In Flue Gas Desulfurization (FGD) scheme, the SO₂ in the flue gas will be scrubbed by the readily available seawater as an absorbing medium. The seawater will absorb SO₂ and convert it to harmless SO₄ ions; a natural constituent of seawater. The seawater will be eventually disposed into the open sea.

- Implementation of the proposed fuel gas sweetening and flue gas desulfurization schemes exposes the minimum foot-print within ADGAS plant in Das Island compared to any other technologies. Also, the proposed schemes do not depend on continuous resource supply from outside the Das Island.

- Implementation of the proposed SO₂ minimization schemes is expected to reduce the total SO₂ emissions by 77% (from 27,532 to 6364 ton/yr): The fuel gas sweetening scheme is expected to reduce H₂S content in the fuel gas by 94% (from 1200 to 72 ppm) and result in decreasing the total SO₂ emissions due to fuel gas usage by 98% (from 10,092 to 168 ton/yr). The flue gas desulfurization scheme is expected to reduce the SO₂ emissions due to incomplete sulfur recovery in the SRUs by 99.5% (from 11,299 to 57 ton/yr).
- The total module cost of implementing the suggested SO₂ minimization schemes is \$28,863,600. The annual cost of the associated utilities and operating cost is \$4,639,300. The income and savings that will be gained from the produced steam is \$8,119,300 per year. Thus, implementing the SO₂ minimization schemes will cover all operating and utilities cost and will yield a net annual income of \$2,730,000. The payback period on the original investment is 10.6 years.

The implementation of the proposed SO₂ minimization schemes is expected to have all SO₂ emission sources (except the sour flares due to flash gas flaring) comply with UAE-FEA limits (500 mg/Nm³). Implementation of the flash gas recovery system will result in the compliance of all trains sour flare with the UAE-FEA limits. Also the SO₂ emissions will have the potential to challenge any future stringent UAE-FEA limits with high level of confidence as the emission rates at the proposed conditions are reduced to about 5% of the standard (i.e., 25 mg/Nm³). This will be presented in Part II of this work.

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