

# START-UP AND OPERATION OF A NEW THIOPAQ O&G UNIT FOR FLARE GAS RECOVERY IN OMAN

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## ABSTRACT

In 2015, the World Bank launched the “Zero Routine Flaring by 2030” initiative to eliminate continuous flaring of associated gas by the end of this decade. The Sultanate of Oman and its main oil and gas company, Petroleum Development Oman (PDO), joined the initiative in 2017, reflecting the country’s commitment to decrease emissions from the production of oil and gas.

In Q1 2025, a new gas treatment plant for the recovery of sour flare gas from oil production has been taken into operation at an existing production station in Oman. To remove H<sub>2</sub>S from the sour flare gas, PDO has selected the Thiopaq O&G (TOG) process, supplied and licensed by SLB and Paqell. The TOG process is a biological process, applying naturally occurring bacteria to convert toxic H<sub>2</sub>S to reusable elemental biosulfur. It is an environmentally friendly and cost-effective alternative to conventional physico-chemical desulfurization processes, such as liquid-redox and scavengers. It is especially suitable for gas streams with an S-load of 1 – 50 ton-S/day. The Thiopaq O&G technology is owned by Paqell. For the current project, SLB is the Authorized Licensor.

The TOG unit is designed for a total gas flow rate of 365,000 Sm<sup>3</sup>/day of a combination of 3 flare gas streams: GRC, MCD-II and RTQ. It was started-up in February/March 2025. During start-up, it appeared that the CO<sub>2</sub> and mercaptan concentrations were significantly higher than anticipated at the design. The process margins with respect to alkalinity and lean solution flow rate utilized in the design allowed for the successful treatment despite the compositional variance of the sour flare gas.

One of the key considerations when selecting the TOG technology was the impact of Oman’s hot climate, as the temperature of the process solution needs to be maintained at 35 – 40 °C for optimal biological activity. By including several measures in the design, the solution temperature could very well be maintained <40 °C.

The TOG unit had to be shut down for repair of the inlet gas cooler and to resolve the carry-over of fine sulfur particles from the Outlet Gas Scrubber to the downstream Booster Compressor. Both issues were successfully resolved. 4 – 5 months after initial gas intake, the TOG unit achieved a gas flow rate of 355,000 Sm<sup>3</sup>/day, which is close to the design gas flow rate of 365,000 Sm<sup>3</sup>/day. The TOG unit appeared to operate very stable and robust without process upsets. The consumptions of nutrients, caustic and make-up water, and the production of the bleed stream are all well within design limits. Gas is being exported to the South Oman Gas Line. Additionally, the 72 hours Site Acceptance Test (SAT) for the facility was completed.

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## 1. Introduction

Routine gas flaring has been common practice in the oil and gas industry. During crude-oil production, natural gas is often released as a by-product, known as ‘associated gas’. When there is no viable infrastructure or market for its utilization, this gas is typically combusted in on-site flare stacks.

Globally, approximately 150 billion m<sup>3</sup> of gas is flared each year, resulting in the emission of about 300 million tons of CO<sub>2</sub> [1, 2]. In addition, around 55 billion m<sup>3</sup> of gas (mainly methane) is vented due to leaks or inefficient flare performance, responsible for roughly 1,180 million tons of CO<sub>2</sub>-equivalent [2]. Besides the negative impacts on climate change and local air quality, gas flaring represents a major economic loss, wasting a valuable energy resource estimated at over \$20 billion annually [3]. Despite global awareness and technological progress, the total volume of flared gas has remained largely unchanged over the past decades [1].

In 2015, the World Bank launched the “Zero Routine Flaring by 2030” initiative to eliminate continuous flaring of associated gas by the end of this decade [1, 4]. Currently, more than 35 countries and 60 petroleum companies have endorsed this initiative, committing to infrastructure development, gas recovery, and improved operational practices [5].

The Sultanate of Oman and its main oil and gas company, Petroleum Development Oman (PDO), joined the initiative in 2017, reflecting the country’s commitment to decrease emissions from the production of oil and gas [5, 6]. Although flaring intensity has decreased, about 2.6 billion m<sup>3</sup> of gas was flared in 2024 [1]. Achieving zero routine flaring by 2030 will require field-level action.

The topic of this paper is the implementation of the Thiopaq O&G (TOG) technology to recover flare gas at the Marmul Production Station in South-Oman. Most flare gas streams have a medium size S-load (i.e. between 0.2 – 50 ton-S/day), and the TOG technology is especially suitable for treatment of gas streams with an S-load in this range.

## 2. Project Background

### 2.1 Marmul

The Marmul field is located on the eastern side of the South Oman basin, approximately 200 km northeast of the city of Salalah. PDO acquired the concession license of the Marmul field in 1976. The field was brought on stream by PDO in 1980.

The Marmul Main Production Station (MMPS) receives gross fluids from all fields in the Marmul and Rahab Thuleilat Qaharir (RTQ) area. It also receives fluids from the Birba station, which are high in H<sub>2</sub>S. The fields fed into MMPS are divided based on their gas to oil ratio (GOR). Higher GOR production is fed to bulk separators with the separator liquid routed to the free water knock out (FWKO) tanks. The lower GOR production, the GRC gas, is fed directly to the FWKO tanks. The lower GOR production, the GRC gas, is fed directly to the FWKO tanks.

The gross production from Marmul B/C/D and G fields are received at the MMPS inlet separators which separate the production fluids into liquid and gas streams. The produced associated gas and gas received from the MM-A station, with contains approximately 700 ppmv H<sub>2</sub>S, is compressed

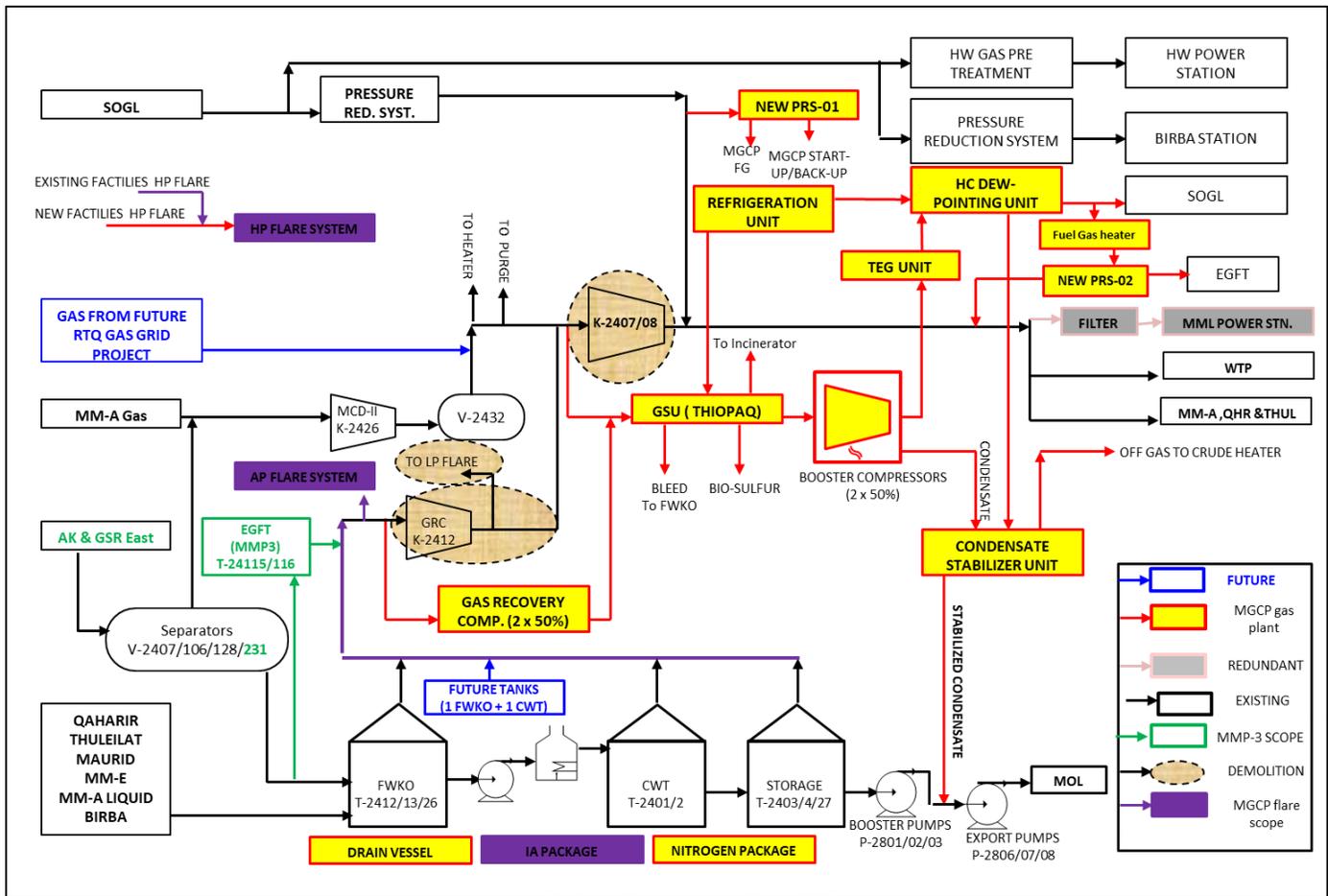


Figure 1 – MMPS Station Line-up with the New Facilities

in the MMPS old gas compression system (MCD-II) and was originally used as fuel gas for the MMPS internal consumers, with excess gas routed to the Marmul Power Station.

Production from Birba, Qaharir, and Thuleilat stations as well as other low GOR fields is received in FWKO tanks where it commingles with the separator liquids. Oil/water separation takes place in the FWKO tanks. The oil/water emulsion from the FWKO tanks goes through several processing stages before export. These include heating, concentric wash tanks (CWT) and storage tanks. The flash gas from the MMPS FWKO tanks, CWT and storage tanks are routed to a common header and compressed by the old Gas Recovery Compressor (GRC). This gas contains high levels of H<sub>2</sub>S (approximately 6000 ppmv H<sub>2</sub>S) because of Birba. In the original line-up, this high H<sub>2</sub>S flash gas was flared via the old GRC to avoid making the entire MMPS gas system to sour and due to the incompatibility of the power plant with high H<sub>2</sub>S gas.

With the old MMPS line-up being quite old and due for retirement, integrity issues with the old GRC and booster compression systems and PDO's aspiration to eliminate flaring across all its stations, the decision was made to install a new Marmul Gas Compression plant (MGCP). The GRC and Booster compressors were replaced and additional equipment for sweetening, dehydration, and dew pointing of the gas in order to route it to the South Oman Gas Line (SOGL) were installed. Figure 1 depicts the old lineup as well as the new lineup within MMPS station.

In the new MGCP lineup, the associated gas from the production separators (MCD-II) is mixed with high H<sub>2</sub>S flash gas recovered from various tanks and routed via the new GRC to the Gas Sweetening Unit (GSU) for removal of H<sub>2</sub>S, using the Thiopaq O&G (TOG) technology. Sweet gas leaving the GSU is compressed, and conditioned (dehydrated, compressed and dew pointed), before being sent to the SOGL.

## 2.2 PDO's Thiopaq O&G Journey

Due to the change in the flaring policy in Oman, it has been mandated to find alternatives for gas flaring for gas with lower sulfur loads. After a wide search for an economical option for the lower sulfur loads the Thiopaq O&G (TOG) technology was nominated as the best candidate.

Before the TOG technology was endorsed to be used in PDO, a deep investigation of the technology was conducted by PDO gas processing SMEs, the search included several visits to TOG plants and risks identification / mitigation processes. Several sessions were conducted between PDO SMEs and the TOG licensors (Paqell and SLB) to identify all risks and develop a mitigation plan. Considering novel technology for PDO, multidisciplinary risks items were identified, and mitigations were mandated before the technology was endorsed. These risks cover all aspects associated with the technology and interaction with PDO/Oman from the design stage to the operational phase. A major example of the risks identified is the capability of the bacteria to survive in Oman summer environmental conditions (+50 °C).

### 3. Thiopaq O&G Process

The Thiopaq O&G (TOG) process consists of 3 main integrated process steps: the Absorber, the Bioreactor, and the Sulfur Separation section, see Figure 2.

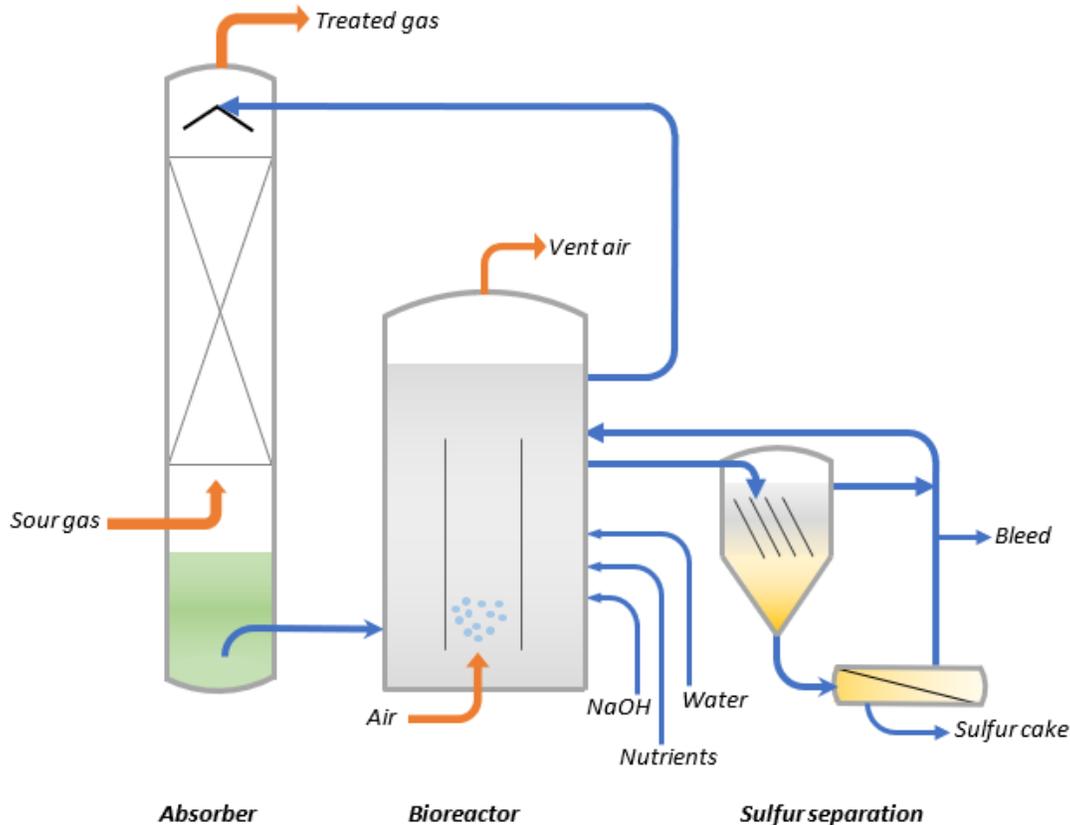


Figure 2 – Traditional Thiopaq O&G Process Line-up [7]

#### Absorber

The first step is the Absorber column, where sour gas is counter currently contacted with Thiopaq process solution coming from the Bioreactor. For low pressure gas streams (i.e. up to 3 bar(g)), a treated gas  $\text{H}_2\text{S}$  concentration of < 25 ppmv is achieved, whereas for high pressure applications (i.e. > 3 bar(g)),  $\text{H}_2\text{S}$  is removed to < 4 ppmv. The process can be applied for gas streams up to 80 bar(g) and is not limited to a certain  $\text{H}_2\text{S}$  concentration range. The treated gas leaves the Absorber from the top for further processing. For certain applications, the feed gas is pre-conditioned entering into the Absorber column.

Three reactions occur in the Absorber. Due to the alkalinity of the process solution,  $\text{H}_2\text{S}$  is converted into bisulfide ( $\text{HS}^-$ ), see reaction 1 in Table 1. Co-absorption of  $\text{CO}_2$  per reaction 2, also takes place. Both reactions consume alkalinity and lead to a slight decrease in the pH of the solution. Due to the presence of bio-sulfur particles in the process solution, part of the sulfide reacts with  $\text{S}^0$  to form polysulfide ( $\text{S}_x^{2-}$ ) per reaction 3, also occurs in the Absorber. The sulfide-rich solution is collected in the Absorber sump and routed to the Bioreactor, which is the next step of the process.

**Table 1** – Main Reactions in the Thiopaq O&G Process [7]

1	$\text{H}_2\text{S (aq)} \rightleftharpoons \text{HS}^- + \text{H}^+$	H <sub>2</sub> S deprotonation (Absorber, chemical)
2	$\text{CO}_2 + \text{H}_2\text{O} \rightleftharpoons \text{HCO}_3^- + \text{H}^+$	CO <sub>2</sub> dissociation (Absorber, chemical)
3	$\text{HS}^- + \frac{x-1}{8} \text{S}_8 \text{ (s)} \rightleftharpoons \text{S}_x^{2-} + \text{H}^+$	Polysulfide formation (Absorber, chemical)
4	$\text{HS}^- + \frac{1}{2} \text{O}_2 \rightarrow \frac{1}{8} \text{S}_8 \text{ (s)} + \text{OH}^-$	Sulfur formation (Bioreactor, biological)
5	$\text{HS}^- + 2 \text{O}_2 \rightarrow \text{SO}_4^{2-} + 2 \text{H}^+$	Sulfate formation (Bioreactor, biological)
6	$\text{HS}^- + \text{O}_2 \rightarrow \frac{1}{2} \text{S}_2\text{O}_3^{2-} + \frac{1}{2} \text{H}_2\text{O}$	Thiosulfate formation (Bioreactor, chemical)
7	$\text{HCO}_3^- + \text{H}^+ \rightleftharpoons \text{CO}_2 \uparrow + \text{H}_2\text{O}$	CO <sub>2</sub> stripping (Bioreactor, chemical)

### Bioreactor

In the Bioreactor, multiple reactions occur but there are four (4) biological/chemical reactions that are of primary interest. Air is injected into the bottom of the Bioreactor via an air distribution grid for the oxygen requirement and efficient mixing of the Bioreactor inventory. The air flow rate to the Bioreactor is controlled via the online measurement of the oxidation / reduction potential (ORP or redox) in the lean Thiopaq solution exiting the Bioreactor.

The dissolved sulfide is oxidized by a mixed culture of Sulfide Oxidizing Bacteria (SOB) to elemental sulfur per reaction 4. This regenerates the process solution, which means that the alkalinity that is consumed in the Absorber for H<sub>2</sub>S absorption, is restored. Thus, the Thiopaq solution that is recycled back to the Absorber is called lean solution as it is free of residual HS<sup>-</sup>.

A small fraction of the incoming sulfide is oxidized to sulfate (SO<sub>4</sub><sup>2-</sup>) and thiosulfate (S<sub>2</sub>O<sub>3</sub><sup>2-</sup>). The formation of sulfate (reaction 5) occurs biologically, whereas thiosulfate is formed by a chemical process (reaction 6). The formation of these unwanted by-products does not regenerate the process solution and therefore requires the addition of caustic soda (NaOH) and make-up water. Since sulfate and thiosulfate are dissolved sodium salts, a waste stream (bleed) is produced to remove these by-products. Since no toxic chemicals are used in the process, the bleed is non-hazardous and normally can be discharged to a conventional waste-water treatment plant. The air strips CO<sub>2</sub>, resulting in a slight increase in the pH of the process solution (see reaction 7). The exhaust air from the Bioreactor (vent air), which is oxygen-depleted air, exists the top of the Bioreactor.

The Thiopaq O&G process solution is an aqueous sodium (bi)carbonate solution, containing sulfate- and thiosulfate ions, SOB, and particles of bio-sulfur. The pH can range between 8.0 – 9.5. The actual pH is dependent on the plant-specific process conditions, mainly the CO<sub>2</sub>/H<sub>2</sub>S ratio of the feed gas. The (bi)carbonate acts as a buffer to minimize pH fluctuations and ensures efficient H<sub>2</sub>S absorption in the Absorber column. The concentration of (bi)carbonate is also referred to as ‘alkalinity’, which is a measure of the buffer capacity of the solution.

The SOB derives their metabolic energy for reproduction by the oxidation of the sulfide. Nutrients are supplied to the Bioreactor, which provides the SOB with the essential elements for growth. As such, the SOB are self-sustaining and after their addition to the system at the initial start-up, no addition of SOB is required. Furthermore, the SOB are robust and adapt to changing and fluctuating process conditions.

The elemental sulfur, also referred to as bio-sulfur, that is produced by the bacteria is hydrophilic and is suspended in the Thiopaq process solution. The hydrophilic nature of the bio-sulfur prevents plugging.

### Sulfur Separation Section

The removal of the bio-sulfur from the Thiopaq process solution occurs in the sulfur separation section, which is the final step of the process. The sulfur is removed from the process solution by a decanter-centrifuge, which produces a sulfur cake with a dry weight of approximately 60 wt.% solids. For units with a higher S-load, a gravity settler is included upstream the decanter-centrifuge. The bio-sulfur has various applications, such as fungicide or fertilizer production. The centrate recovered from the centrifuge is usually recycled back to the Bioreactor but when required due to level control of the overall system, it is routed to disposal as an effluent bleed stream.

## 4. Thiopaq O&G Design for Marmul

### 4.1 Design Basis

The TOG unit at Marmul is designed to treat the combination of 3 gas streams: GRC, MCD-II and RTQ. The flow rates and compositions of the gas streams can change overtime, and in some occasions, only one of the gas streams is available. Therefore, multiple cases were defined, of which the most relevant cases are summarized in Table 2.

**Table 2** – Gas Flow Rates and Composition of the 3 Main Design Cases. In the Max Case it is Considered That The GRC Gas Stream Contains 100 ppm Mercaptans.

	GRC + MCD-II + RTQ (Max case)	GRC Only	MCD-II only (Turndown case)
Flow (Sm <sup>3</sup> /day)	365,000	146,805	166,085
H <sub>2</sub> S (ppmv)	4000	5400	600
CO <sub>2</sub> (vol%)	1.18	1.24	0.96
C1 (vol%)	68.5	54.0	88.5
C2 (vol%)	4.5	5.6	3.5
C3+ (vol%)	22.3	37.8	5.4
S-load (kg-S/day)	1980	1055	143
Total Mercaptans (ppmv)	0	100	0
Total BTEX (ppmv)	1100	1400	600

In the ‘Max case’ all 3 gas streams are operated, resulting in a total flow rate of 365,000 Sm<sup>3</sup>/day. The H<sub>2</sub>S and CO<sub>2</sub> concentration are 4000 ppmv and 1.18 vol% respectively. Based on this gas flow rate and H<sub>2</sub>S concentration, the Sulfur load will be 1980 kg-S/day.

As can be seen in Table 2, most of the Sulfur load comes from the GRC gas stream, which contains the highest H<sub>2</sub>S concentration. In case the unit is operated on GRC gas only, the Sulfur load will be 1055 kg-S/day, and in case the unit is operated on MCD-II only (i.e. the Turndown case), the S-load can be as low as 143 kg-S/day, which is 7% of the design S-load. An “RTQ only” case has

not been defined, since this gas stream is always operated in combination with GRC and/or MCD-II.

All gas streams have a pressure of about 4 - 5 bar(g), and the design pressure at the inlet of the TOG Absorber is 3.75 bar(g).

## 4.2 TOG Design

The TOG unit at Marmul consists of a Feed Gas pre-conditioning section, an Absorber section, a Bioreactor section, a Sulfur Separation section, and a Chemical Dosing section.

### Feed Gas Pre-Conditioning Section

In the Feed Gas pre-conditioning section, the incoming feed gas is first cooled with the Inlet Gas Cooler. The cooled gas subsequently passes through an Inlet Gas Scrubber to knock-out condensed liquids. The high-efficiency internals in the Inlet Gas Scrubber ensure minimal carry-over of condensed liquids to downstream equipment. The gas is then reheated by the Inlet Gas Heater, before it enters the TOG Absorber column.

### Absorber Section

The Absorber column ('Contactor') contains 2 beds of random packing where gas and liquid are counter-currently contacted. Pre-conditioned feed gas enters the Absorber below the bottom bed. Lean process solution is supplied to the top of the upper bed. The rich solution collected in the Absorber sump. The gas exiting the Absorber passes through an Outlet Gas Scrubber. The high-efficiency internals in the Outlet Gas Scrubber minimize carry-over of particles and liquid droplets to downstream equipment.

### Bioreactor Section

Rich solution from the Absorber is routed to the Bioreactor via a Rich Solution Cooler to control the solution temperature. The Bioreactor is an atmospheric tank. In the bottom of the Bioreactor, air is injected via an air distribution system. The air is supplied by an air blower, with an installed capacity of 2x100%. A Trim Cooler and Air Cooler cool the air before entering the Bioreactor. Vent air leaves the Bioreactor at the top and is sent to a Thermal Oxidizer. Rich solution is injected via a rich solution distribution system, which is located just above the air distribution system. The 'lean' process solution, which is free in H<sub>2</sub>S leaves the Bioreactor then flows on gravity to the Lean Solution Tank, which is also an atmospheric tank. From the bottom of the Lean Solution Tank, the solution flows to the main Circulation Pump (installed capacity 2x100%), which pumps the lean solution to the top of the Absorber, and to the Sulfur Separation section.

### Sulfur Separation Section

The Sulfur Separation Section has 2 parallel Decanter-Centrifuges, with a capacity of 100% each. The sulfur cake is collected in 1000 kg big bags, and the centrate overflows into a Filtrate Tank. The majority of the centrate is sent back to the lean solution tank, and a small fraction is discharged to the bleed facility.

### Chemical Dosing Section

Caustic (NaOH), Nutrients, and Make-up water are supplied from their respective dosing tanks with dedicated dosing pumps into the Bioreactor. Since water is used as seal for the Circulation Pumps, a minimum amount of Make-up water is continuously supplied to the Pump seal system. Furthermore, Antifoam can be dosed on an as-need basis from the Antifoam dosing tank.

### 4.3 Temperature Control Design Considerations

Maintaining the TOG solution temperature between 35 – 40 °C is essential as the bacteria can be irreversibly inhibited when exposed to temperatures >45 °C for prolonged time. This was one of PDO's main concerns when selecting the TOG technology. Since ambient temperatures in Oman can easily reach 50 °C, extra attention was paid in the design to control temperature.

The oxidation of H<sub>2</sub>S to elemental sulfur is an exothermal reaction that releases heat. As such, any TOG unit needs a solution cooler to remove the reaction heat. For the Marmul design, the design margin on Solution Cooler was extended and a larger frame was selected to allow for additional plates. The air from the Air Blower into the Bioreactor always requires cooling to keep the air temperature entering the Bioreactor < 50 °C. In Marmul, a trim cooler has been added to assist the primary air-cooled exchanger.

All main process tanks were insulated to minimize the impact of ambient temperature, which also lowers the cooling duty of the solution cooler. Utility tanks, for example the Nutrient dosing tank, was kept smaller to lower the residence time of chemicals in their respective dosing tanks. A temperature controlled warehouse near the plant ensures a regular shelf life of Nutrients and Seed Sludge.

The building for the Decanter-Centrifuges was open sided to provide natural air circulation which reduces sulfur dusting. Additionally, the collecting of the sulfur cake in big bags further minimizes sulfur dusting.

## **5. Project Execution**

After selecting Thiopaq O&G as the technology of choice, PDO decided that the way forward was to develop a basic design package (BDP) for the technology before the final decision is given on the project. As owner of the Thiopaq O&G technology, Paqell made a process design using its proprietary design tool. In the process design, sizing of equipment, solution flow rates and utility and chemical consumptions are determined.

For project execution, one of Paqell's Authorized Licensors had to be selected, which was SLB. Based on the process design of Paqell, SLB developed a BDP, which was essentially a more detailed FEED package for the technology deployment in MGCP. The BDP ensured that all the risks identified during the technology evaluation stage were properly mitigated with acceptable measures and the residual risks are tolerable. Following the PDO approval of BDP, the project was endorsed to be executed.

The detailed engineering and equipment delivery for the TOG unit was done by SLB. The skids were built in Oman, utilizing the experience of Omani contractors, which ensured easy logistics of the equipment to site. The installation of the TOG unit was performed by the main EPC contractor selected by PDO. The EPC contractor was also responsible for the installation of all other process units at the MGCP site, such as the compressors, dehydration, and dew-pointing units. A picture of the TOG unit is shown in Figure 3.



**Figure 3** – Picture of the TOG Unit at Marmul. 1: Feed Gas Inlet and Pre-Conditioning; 2: Absorber; 3: Bioreactor, 4: Lean Solution Tank, 5: Decanter-Centrifuges and Sulfur Bagging, 6: Dosing Tanks of Make-up Water, Caustic, Nutrients and Antifoam Agent; 7: Thermal Oxidizer.

## 6. Commissioning And Start-up

### 6.1 Initial Start-Up of the TOG (February – April 2025)

Both SLB and Paqell were present to support and guide PDO and the Main EPC contractor during the initial start-up of the TOG unit.

To prepare for start-up of the TOG unit, the entire unit was cleaned and filled with make-up water. The complete liquid circulation circuit was started and tested during a water run, in which the DCS loops, flows, and setpoints were checked and optimized for start-up. When the unit was found to be completely functional, sodium bicarbonate salt ( $\text{NaHCO}_3$ ), commonly known as baking soda, was loaded into the Bioreactor to establish the required alkalinity and conductivity. Subsequently, bacteria were added to the process solution by loading ‘Seed Sludge’ into the Bioreactor.

The subsequent day, the feed gas line was purged using fuel gas. Then sour feed gas addition to the TOG unit was initiated. At the time of initial start-up, the RTQ gas stream was not yet available. It was decided to start-up with the GRC gas stream because it has the highest sulfur load. At initial stages of the start-up, the gas flow rate is introduced at a limited flow rate and then gradually increased over the next few days. At Marmul, this was achieved by partially opening the feed gas valve (FCV-10001).

A few hours after start-up it was noticed that both the vent air from the Bioreactor as well as the sample had a strong smell of mercaptans. This was unexpected since the GRC gas was expected to contain <100 ppm mercaptans. It is known that mercaptans are partly absorbed by the TOG process solution, and leave the system via the vent air, biosulfur and bleed. Excessively high mercaptan concentration inhibits the bacteria, which limits the sulfide oxidizing capacity. In the next couple of days, the operation was challenging and the increase in biological activity was very slow, of which the presence of the mercaptans is believed to have been the major cause.

Consequently, the TOG feed gas source was changed from the GRC to the MCD-II gas. This immediately resulted in a considerable improvement in the operation. The gas flow rate could be increased to about 180,000 Sm<sup>3</sup>/day (50% of design) within 4 days, which at that time was the maximum available amount of MCD-II gas. Two days after starting MCD-II gas, the Decanter-Centrifuge was started-up, which produced a dry sulfur cake.

Besides the mercaptan concentration, also the CO<sub>2</sub> concentration appeared to be much higher than anticipated in the design phase. In general, a higher CO<sub>2</sub> concentration leads to a lower efficiency for H<sub>2</sub>S absorption, resulting in a higher H<sub>2</sub>S concentration in the treated gas. To compensate for the higher CO<sub>2</sub> concentration in the gas stream, the TOG was operated at the maximum possible alkalinity and maximum possible lean solution flow rate. Under these conditions an H<sub>2</sub>S concentration of 5 – 10 ppm could be achieved, which was acceptable to start-up the downstream units.

In the next days and weeks, the chemical dosing rates, redox setpoint, operation of the decanter-centrifuges and other process parameters were optimized. Furthermore, more MCD-II gas became available, and the gas flow rate was further increased to 330,000 Sm<sup>3</sup>/day (90% of the design).

## 6.2 TOG Shutdown for Inlet Gas Cooler Repair (April – May 2025)

Just as it appeared that the performance of the TOG unit was lining out, the chilled water unit which supplies the cooling medium for the TOG solution cooler and inlet cooler began to have some issues. The unit tripped on low level in the chilled water expansion tank in late April. After inspection and investigation of the entire chilled water system across all the process units, it was determined that the TOG Inlet Gas Cooler, which is part of the pretreatment section of the process on the inlet to the Absorber, was leaking. The TOG Inlet Gas Cooler is floating head, shell & tube exchanger (TEMA type AES) with the gas on the tube side and the chilled water on the shell side. Pictures of the TOG Inlet Gas Cooler are shown in Figure 4.

The floating head end shell has a 2” vent valve. After isolation of the chilled water supply and return lines, H<sub>2</sub>S was detected from the shell-side vent valve. This indicated a possible leak in the cooler with the H<sub>2</sub>S coming from the tube side of the exchanger since it is the sour flare gas. The isolation of the chilled water side and venting of the shell was repeated numerous times with the same results. It was suspected that there was a tube leak which would be unusual based on the MOC of 316SS and the short length of operation.



**Figure 4** – Inlet Gas Cooler Floating Head (left) and Floating Head Tube Sheet (right)

The unit was shut down late April to inspect the inlet cooler, verify the suspected tube leak and make any required repairs. The shutdown duration was expected to be perhaps 1 week. The initial hydrotesting from the shell side failed to detect a tube leak and the pressure held. The floating head cover was removed to visually inspect the tubes and complete a penetration test, which detected no issue. The floating head cover was boxed up but the gasket failed at 19.5 bar(g) pressure. The floating head gasket was replaced with the last spare gasket but did not hold at 6.5 bar(g).

New gaskets and bolts were put on order. Additionally, a serration machine to repair/resurface the floating head cover flange was mobilized to site around this same time. In parallel with the flange serration repair, alternative gasket options were explored. In the meantime, the floating head cover was removed, and the heat exchanger tubes were cleaned. Debris found on the shell side was sent to the lab and determined to be iron sulfide. A small scratch was found on the floating head flange serrations that was thought to be causing the gasket failures. Over torquing was suspected to be a contributing factor. About a week to 10 days into shutdown, an old gasket for the floating head was installed with 3 new bolts. After several hydrotests at both the shell and tube side, and a borescope inspection of all tubes which confirmed the tubes were not leaking, the remaining root cause seemed to be a leak of the floating head due to either a gasket failure (misalignment or service compatibility) or damage to the floating head flange.

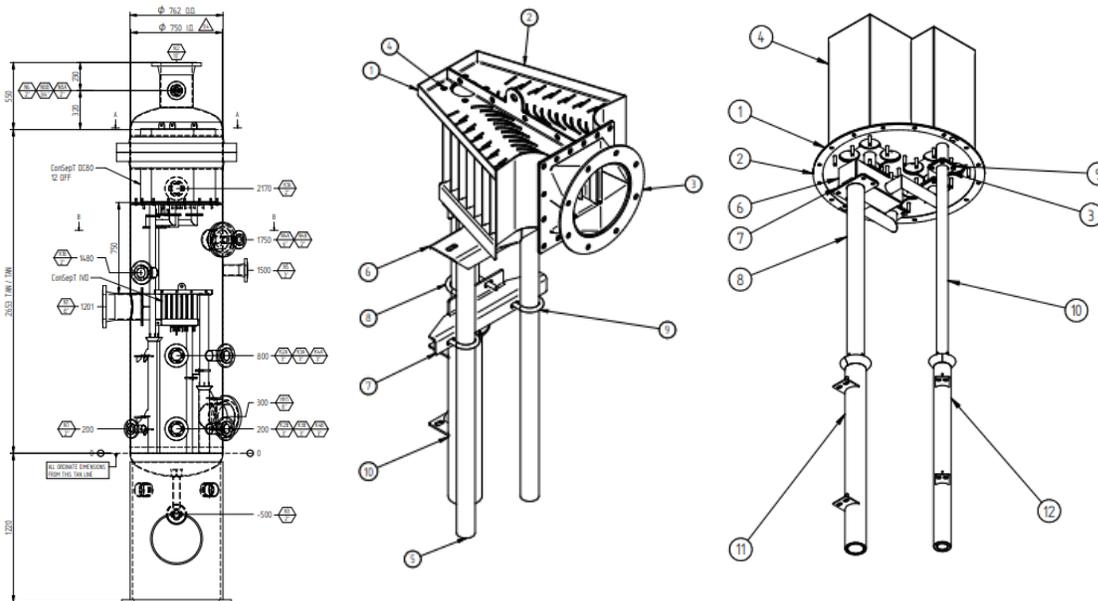
The inlet and outlet chilled water circulation was reestablished to observe the expansion tank level stability and kept on through the night. The expansion tank level was stable. New gaskets, bolts and nuts were received and used to reinstall the floating head cover. A tube side hydrotest failed due to a leak observed from the bottom side of the floating head flanges at 1.8 bar(g). At this point it was decided to use the serration machine to repair the floating head cap flange. If the field serration repair failed, a shop serration repair would be required which would be a significant issue. The field tube sheet flange serration repair was also completed. A tube side hydro test to 2 bar(g) failed. The site was pretty much out of options but a final attempt with an alternate gasket, a Kamm Profile gasket or the floating head flange proved successful. A hydrotest of the tube side at 19.5 bar(g) was successful. Additional testing and inspection was completed but ultimately maintenance approved a return to service order for the cooler, and the TOG unit was restarted at

the end of May. After the restart, the chilled water expansion tank level has held steady and no H<sub>2</sub>S has been detected on the shell side of the inlet gas cooler.

### 6.3 Troubleshooting of the TOG Outlet Gas Scrubber (June 2025)

In the 1<sup>st</sup> week of June, after a successful re-start and ramp-up, successful gas exports to the South Oman Gas Line (SOGL) was achieved. This was a significant milestone. However, shortly thereafter, operations reported experiencing high dP across suction strainers on the Gas Booster Compressor (GBC), which is immediately downstream of the TOG unit.

Inspection of the GBC suction strainers detected deposits of a gray/yellow solid particles that was suspected to be sulfur from the TOG unit. The TOG unit has an Outlet Gas Scrubber on the TOG Absorber outlet gas, which serves as a carryover scrubber to collect any moisture, solution carryover and entrained sulfur in the treated gas prior to going to the GBC. The Outlet Gas Scrubber is fitted with specialty internals that help remove up to 99% of the entrained droplets and solids in the treated gas. A ConSept IVD (Internal Vane Diffuser) assembly is fitted to the gas inlet and a ConSept DC80 (demisting cyclone) apparatus is on the gas outlet. Figures of the Outlet Gas Scrubber lay-out and internals are provided in Figure 5.



**Figure 5** – TOG Outlet Gas Scrubber. Left: General Arrangement Drawing; Middle: Inlet ConSept IVD; Right: Outlet ConSept DC80 (demisting cyclone).

The first step of troubleshooting revolved around collecting information around the Outlet Gas Scrubber which included the level trend of the scrubber. The scrubber level is on gap level control, and if performing correctly, the level trend should appear very much like a saw blade, with the level trending up left to right over time until the LCV opens which drives the level down to the top the close setpoint. Review of the level in the scrubber sump of the first two weeks of June, revealed that there was no level in the outlet scrubber. As pointed out by SLB's subject matter expert

sufficient level is required for the proper function of the Outlet Scrubber. This level (i.e. seal) is required from the start; it is not being build up by itself during operation.

If the sump does not have enough level to submerge the drain tube from the IVD and the demisting cyclones to provide a liquid seal, a portion of the inlet gas exiting the IVD can divert up through the drain tube to the demisting cyclones which impacts the cyclone efficiency. The lack of level in the scrubber sump was determined to be the primary cause for the sulfur migration to the GBC. The lack of level resulted from draining the Outlet Gas Scrubber as part of the shutdown for Gas Cooler repair. After providing a liquid seal using make-up water the unit was restarted, and operation was improved. The necessity of providing a correct liquid seal by using make-up water at startup and that this liquid seal needs to be maintained during operations was included in the startup and restarts procedures.

#### 6.4 Managing Solution Temperature During Long Shutdown (April – May 2025)

During the extended shutdown, the daytime temperatures at Marmul began approaching typical summertime values and PDO had instituted their standard HSE practice of limiting operating personnel exposure to the hot climate during the peak heat in the afternoon. The other challenge involved with the high summer temperature was to maintain the solution temperature < 40 °C to keep the bacteria in a healthy condition. During normal operation, the solution is cooled by the continuous solution circulation over the Rich Solution Cooler. The solution of circulation would also prevent excessive heating of solution in piping. However, the continuous operation of the Circulation Pumps during a long shutdown would result into a considerably addition of Make-Up water via the seal system would cause excessive dilution of the TOG process solution, which is unwanted. Therefore, a specific protocol was developed to maintain the solution temperature < 40 °C throughout the entire system while minimizing the time that solution was circulated.

A lean solution bypass from the discharge of the Circulation Pump to the inlet of the Rich Solution Cooler was rigged up. The Absorber, with a relatively small liquid volume, was depressurized and drained. Furthermore, the Filtrate Tank and small diameter lines in the Sulfur Separation Section and for the tank sprays, were drained. Since the solution in piping would heat up quicker than in the main tanks, a heat gun was used to monitor the temperature in these stagnant lines. When the solution temperature in these lines was approaching 38 °C, the TOG solution was circulated, and the Rich Solution Cooler was operated at maximum duty to remove heat. This was typically done for about 4 hours per day, from 12:00h to 16:00h. At the start of the circulation the solution pump tank temperature might be 38 – 40 °C. After the 4-hour circulation, the cooling provided by the lean solution cooler would drop the temperature to 33 °C which would be cool enough so that the heat up in the early morning hours before noon would not be excessive.

Samples were taken twice per week to monitor the health of the bacteria and TOG solution quality. Over the course of the shutdown, the biomass/bacteria count declined some but stayed within the acceptable limits. The shutdown took exactly 1 month. Prior to restart, the small diameter lines were flushed prior to being put in service. Once the plant was restarted, the bacteria responded well to the reintroduction of the H<sub>2</sub>S. The unit could be restarted within a day, and near design gas rates could be achieved in 5 days after restart.

## 6.5 Site Acceptance Test

Due to the difference between the design and actual gas flow rates and compositions, the Site Acceptance Test (SAT) was performed with MCD-II gas only. The 72-hour SAT was completed by the end of July 2025 (i.e. in the same period as discussed in section 5.2).

Although GRC and RTQ gas were not part of the performance test, the stated streams will be gradually introduced at a later time to determine the actual limit of the TOG unit in its current configuration (in terms of mercaptans and CO<sub>2</sub>) and determine any necessary modification in the MGCP TOG to handle the full gas flow of all streams.

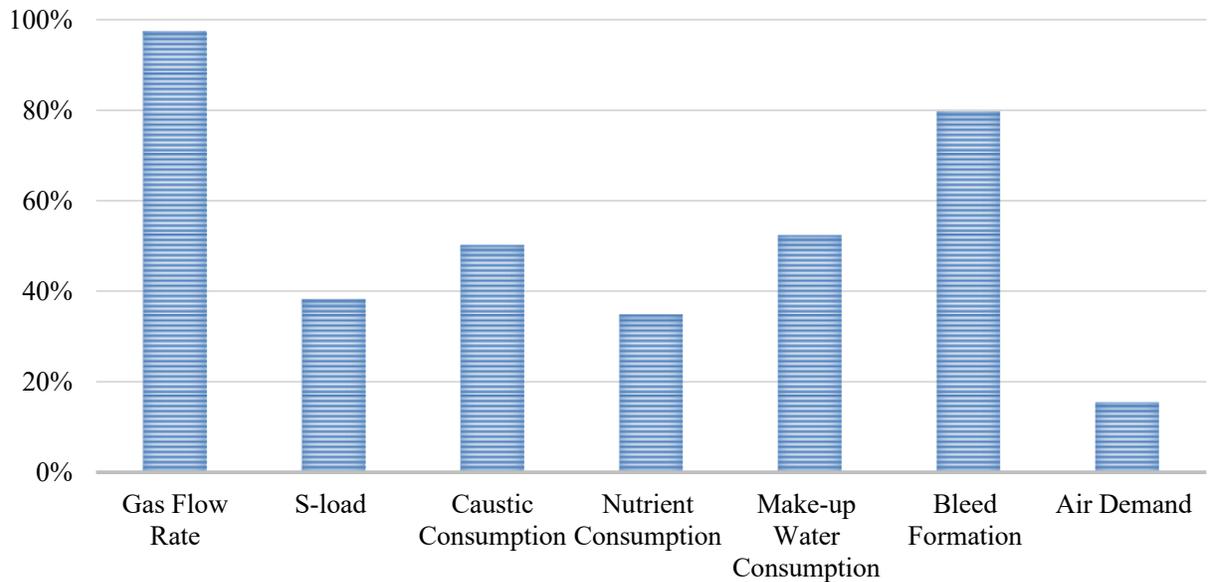
## **7. Process Performance Review**

Due to the high mercaptan levels of the GRC gas, this gas stream is currently not being processed, and the unit operates MCD-II gas only.

Since the repair of the Outlet Gas Scrubber, for which the unit was down for about 3 days, the TOG unit has been in continuous operation. In a period of 4 months (July – October 2025), only 1 automated shutdown was reported, which was caused by a false reading of a level transmitter. The unit could be restarted in 1 hour. Furthermore, no upsets or foaming incidents occurred, and no antifoam agent is being dosed. The solution temperature can be easily maintained < 40 °C with the solution cooler. Due to the low relative humidity, the air injected into the Bioreactor likely has a cooling effect on the solution by evaporation.

The process performance of the TOG unit has been assessed based on operation of MCD-II gas over a period of 2 weeks in July 2025. During this 2-week period, the average gas flow rate was 356,000 Sm<sup>3</sup>/day (design 365,000 Sm<sup>3</sup>/day). The actual operating pressure at the inlet of the Absorber was slightly higher than design (3.83 vs 3.74 bar(g)). The H<sub>2</sub>S and CO<sub>2</sub> concentrations of MCD-II gas are measured several times per week. The H<sub>2</sub>S concentration of MCD-II is roughly double the value that was expected during the design phase (600 – 800 ppm). Still, the H<sub>2</sub>S concentration is lower than the overall H<sub>2</sub>S concentration because the GRC gas stream, which has an H<sub>2</sub>S concentration of 6000 ppmv, is not being processed. The CO<sub>2</sub> concentration is more than twice as high as the design value. The Sulfur load to the unit (resulting from the gas flow rate and H<sub>2</sub>S concentration) is on average about 40% of the design value. This is well above the turndown S-load, which is 10% of the design.

The average H<sub>2</sub>S concentration in the treated gas was 7 ppm, which is below the specification of 10 ppmv. Although the TOG is operated at the maximum possible alkalinity and maximum possible lean solution flow rate, the H<sub>2</sub>S concentration is higher than preferred. This is a result of the higher CO<sub>2</sub> concentration in the gas stream. A higher CO<sub>2</sub> concentration leads to a less efficient absorption of H<sub>2</sub>S, because H<sub>2</sub>S and CO<sub>2</sub> ‘compete’ for the alkalinity of the process solution. The consumptions of Caustic, Nutrients, and Make-up water and the bleed water formation are all significantly lower than the design values (i.e. all values are <100%). This is mainly the result of the lower actual S-load, see Figure 6.



**Figure 6** – Relative Values for the Consumption of Caustic, Nutrients, Make-up Water and Air, and the Production of Bleed. 100% Indicates the Design Value.

The caustic consumption is 52% of the design value, whereas the S-load is 38% of the design. Hence, the specific caustic consumption (kg-NaOH / kg-S) is higher than design. In general, the caustic consumption is determined by: (i) the formation of the unwanted by-products sulfate and thiosulfate, and (ii) the actual alkalinity. A higher formation of by-products, which is expressed as a lower selectivity for  $S^0$  formation results in higher caustic consumption. The selectivity for sulfur formation is better than design. Furthermore, a higher alkalinity (i.e. sodium bi-carbonate concentration) leads to a higher caustic consumption, because more sodium bi-carbonate is lost via the bleed. Since the unit is operated at the maximum possible alkalinity, the specific caustic consumption is higher.

The Nutrient consumption is 35% of the design value, which is explained by the lower S-load. The specific nutrient consumption is practically the same as the design value.

The make-up water consumption and bleed water formation are 52% and 80% of the design values, respectively. Typically, the make-up water consumption and bleed water formation are approximately equal. In case of relatively low humidity, a significant amount of water is evaporated by the air in the Bioreactor. In that case, the bleed flow will be lower than the make-up water flow due to the loss of water via the Bioreactor vent air. Since the ambient humidity is indeed relatively low in Oman, in the design, the bleed flow rate was about 65% of the make-up water flow. However, in practice it appears that make-up water flow rate and bleed flow rate are practically equal. Apparently, not much water is lost with the Bioreactor vent air, which could be due to the relatively low air demand, as further discussed below. Although for process reasons less water could be dosed, this is not possible due to a minimum amount required for the pump seal systems. This is the reason that make-up water and bleed flow are relatively high compared to the S-load.

Furthermore, it appears that air demand is much lower than the design value. This is also the result of the lower S-load (with less H<sub>2</sub>S, less air is needed), but also the ‘specific’ air demand (i.e. the Nm<sup>3</sup>/h air per kg-S/day) is also much lower than the design/expected value. This is considered an indication of very active bacteria.

The biosulfur cake produced by the Decanter-Centrifuges is very dry, see Figure 8. In most TOG installations, the sulfur cake is still slightly wet, but in Marmul the cake is in powder form most of the time.



**Figure 8** – Produced Biosulfur Cake at the Decanter-Centrifuge Outlet

## **8. Conclusions and Outlook**

### **8.1 Conclusion**

To meet the commitment of zero flaring in 2030, PDO recently has installed a new gas treatment plant for the recovery of flare gas from the Marmul Main Production Station in South-Oman. For the removal of H<sub>2</sub>S, a Thiopaq O&G (TOG) unit was successfully commissioned and started-up. The TOG process is a biological process, using naturally occurring bacteria to convert H<sub>2</sub>S to elemental bio-sulfur.

During start-up, it appeared that the CO<sub>2</sub> and mercaptan concentrations were significantly higher than anticipated at the design. The process margins with respect to alkalinity and lean solution flow rate utilized in the design allowed for the successful treatment despite the compositional variance of the sour flare gas.

One of the key considerations when selecting the TOG technology was the impact of Oman's hot climate, as the temperature of the process solution needs to be maintained at 35 – 40 °C for optimal biological activity. By including several measures in the design, such as extra capacity of the solution cooler and insulation of main tanks, the solution temperature could very well be maintained <40 °C.

The TOG unit had to be shut down for repair of the inlet gas cooler and to resolve the carry-over of fine sulfur particles from the Outlet Gas Scrubber to the downstream Booster Compressor. Both issues were successfully resolved. The shutdown for repair of the gas cooler took 1 month, while ambient temperatures were high. By draining various lines and tanks, and temporary operation of the solution cooler, the temperature could be maintained <40 °C during this long shutdown, preventing excessive decay of bacteria. The unit could be restarted within a day, and near design gas rates could be achieved in 5 days after restart. After the shutdown for the Outlet Gas Scrubber, which took 4 days, near-design gas rates were achieved within a day after restart.

In mid-July, the TOG unit operated at an average gas flow rate of 355,000 Sm<sup>3</sup>/day, which is close to the design gas flow rate of 365,000 Sm<sup>3</sup>/day. Gas is being exported to the South Oman Gas Line. Additionally, the 72 hours Site Acceptance Test (SAT) for the facility was completed. The TOG unit appeared to operate very stable and robust. In a period of 4 months (July – October 2025), only 1 automated shutdown was reported, which was caused by a false reading of a level transmitter. The unit could be restarted in 1 hour. Furthermore, no upsets or foaming incidents occurred, and no antifoam agent is being dosed.

The consumptions of nutrients, caustic and make-up water, and the production of the bleed stream are all well within design limits. Relative to the S-load, the consumption of caustic and water, and the production of bleed is increased, which is related to the operation at increased alkalinity, and the minimum water required for the seal systems of the circulation pumps. The air demand of the bioreactor is very low, which is considered an indication of very active bacteria. Furthermore, the quality and dryness of the produced bio-sulfur cake is very good.

Although the H<sub>2</sub>S in the treated gas is on specification, it is higher than expected and a lower H<sub>2</sub>S concentration in the treated gas would be preferred. Several options are proposed to improve the H<sub>2</sub>S absorption efficiency.

## 8.2 Outlook

PDO is eventually aiming to completely stop flaring at the Marmul Main Production Station. The total gas flow rate is 485,000 – 550,000 Sm<sup>3</sup>/day, which is beyond the design capacity of the current unit. Furthermore, the increased concentrations of CO<sub>2</sub> and mercaptans (for the GRC stream), make it even more challenging to increase gas rates whilst ensuring a low H<sub>2</sub>S concentration in the treated gas and keeping the toxicity within limits.

Together with Paqell and SLB, PDO is currently investigating various options to expand the current capacity of the unit. It seems feasible to increase both the gas capacity and the treating

efficiency of the Absorber by optimization of the Absorber internals. For example, the spray nozzle in the top of the Absorber which distributes the solution to the Absorber packing, could be replaced by a spray nozzle with a higher effective area. This allows for a higher packing height, ensuring that H<sub>2</sub>S is maintained <10 ppm when gas rates are increased. Furthermore, increasing the trim sizes of the flow control valves around the Absorber allow for a higher flow rate, which also improves H<sub>2</sub>S removal.

Based on process simulations by Paqell, it appears feasible to introduce a significant amount of the GRC gas together with MCD-II without toxification of the bacteria in the Bioreactor. Based on these simulations, the operation on GRC only will lead to a too high toxicity (as was also experienced during initial start-up). However, when combining GRC with MCD-II (or RTQ), the mercaptan concentration decreases to acceptable levels.

It's worth noting that are several projects within region may utilize TOG as a potential technology to be deployed to eliminate flaring of gas having a medium sulfur load.

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