Rarely Acknowledged Energy Benefits of Sulphur Recovery

Angie Slavens, Saba Khan
UniverSUL Consulting

ABSTRACT

Energy conservation has always been important to the oil and gas industry but is becoming increasingly critical in the current low oil price climate. Sulphur recovery facilities are necessary to meet emissions regulations and are therefore often viewed as a cost of production. However, the sulphur plant is normally a net energy exporter, providing a frequently overlooked benefit to the energy balance of the processing complex. This is because the Claus reaction, which is employed to convert H_2S to elemental sulphur, is exothermic, and the waste heat from the process can be recovered as HP and LP steam.

This paper will explore the energy benefits of various sulphur recovery technologies. Benchmark key performance indicators (KPIs) will be provided as guidance to operators on how their sulphur recovery facilities should be performing, from an energy efficiency perspective. A Case Study for a sulphur recovery facility in a large sour gas plant will be presented and compared to the Benchmark Plant KPIs. Options for improving energy efficiency will be investigated.

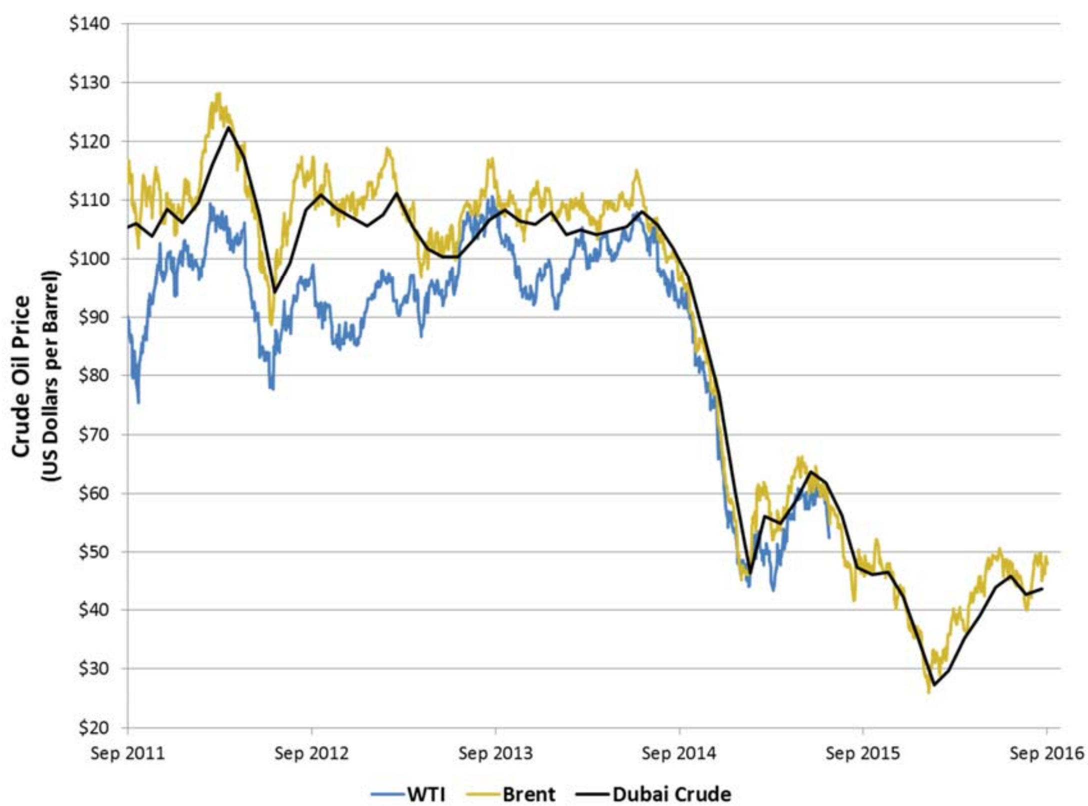
In 2016 the Middle East became the world's largest sulphur producing region. Current UAE sulphur production is approximately 19,000 MTPD (7 MMTPA), equivalent to roughly 11% of global production. The total power export from UAE sulphur recovery facilities should be in the range of 250 MW, which would be extremely beneficial for supplementing the power requirements of the sour gas processing facilities in which they are installed. However, certain factors may be eroding these energy benefits, some of which may be correctable via operational changes, while others may require hardware modifications. Within the next decade, UAE sulphur production could double, which presents the possibility for producing around 500 MW of power from all SRUs in the country. This figure could increase if technologies which maximize energy efficiency are employed and best operating practices are followed.

INTRODUCTION

In late 2014, oil prices plummeted dramatically in the space of just a few months. As shown in Figure 1, around mid-year, the price of oil started dropping rapidly and fell from about $100 per barrel to approximately half of this figure in less than six months. The price suffered additional losses in 2015, to a minimum around $25 per barrel, the lowest value in more than 13 years. The price has rebounded somewhat in 2016 and now sits at around $50 per barrel; however, this is still roughly half the figure enjoyed by the industry during the preceding 5 years.

Low oil price creates both challenges and opportunities for the industry. The challenges are obvious, with threatened economics for new projects and reduced margins on existing production. Those producers who stay focused on achieving success via efficient, cost-effective operations during these challenging times will prosper and thrive when oil price rebounds, allowing for stronger and more profitable operations in the future.

Figure 1. 5-Year Historical Crude Oil Price[1,2]



The sulphur recovery facility within a refinery or gas plant is required to meet SO_2 emissions regulations and is often viewed as a cost of production. However, waste heat from the exothermic Claus reaction is recovered as HP and LP steam, which usually makes the sulphur plant a net energy exporter, supplying needed steam and/or power to the processing complex. For extremely sour gas plants or refineries processing sour crude feedstock, the sulphur plant may be one of the areas of greatest interest for improving energy efficiency and strengthening the economics of production.

As such, this paper will:

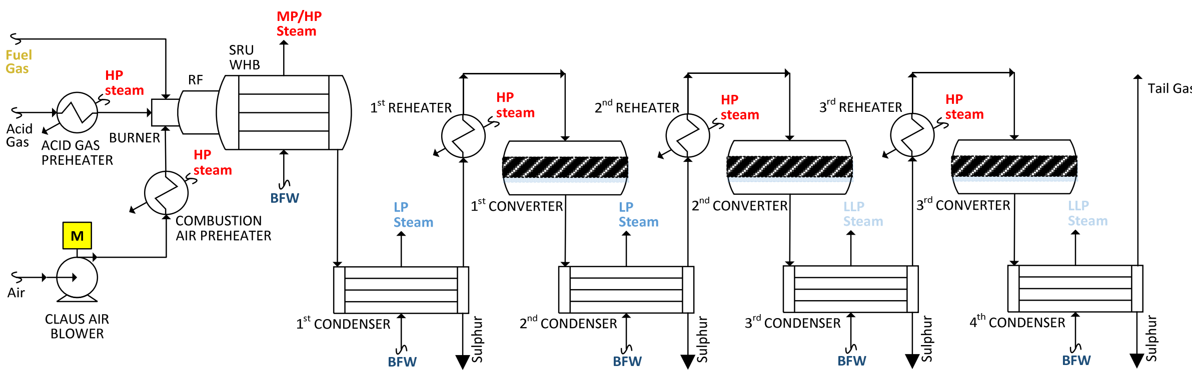
- Identify the top energy producers and consumers in a typical sulphur recovery facility
- Quantify the energy balance and provide benchmark energy performance indicators for a hypothetical Benchmark Plant, across a range of sulphur recovery technologies
- Compare KPIs for the hypothetical plant to a real world Case Study
- Provide suggestions for achieving superior energy performance in a sulphur recovery facility

This information will be used to examine the energy export potential from all SRUs in one of the world's largest sulphur-producing nations.

ENERGY PRODUCTION & CONSUMPTION IN THE SULPHUR PLANT

The Modified Claus process is shown in Figure 2. In this well-known process, one third of the H_2S in the acid gas is burned to form SO_2, which then reacts with remaining H_2S to form elemental sulphur, via the exothermic Claus reaction. Key utilities produced/consumed in the process are steam (HP and LP), fuel gas and electrical power. Typically, some form of tail gas treating is required downstream of the SRU to satisfy SO_2 emissions regulations, the energy requirements of which can be substantial and will thus be discussed throughout this paper.

Figure 2. PFD of 3-stage Claus SRU with Key Utility Streams



Steam

The heat of reaction for the exothermic Claus reaction is as indicated in Equations 1-3. [3]

Thermal Section

$H_2S + 3/2O_2 \rightarrow SO_2 + H_2O$ $\Delta H@77^{\circ}F = -223{,}100$ BTU (1)

Thermal and Catalytic Reaction Section

$2H_2S + SO_2 \rightarrow 3/xS_x + 2H_2O$ $\Delta H@77^{\circ}F = -41{,}300$ BTU (2)

Overall Reaction

$3H_2S + 3/2O_2 \rightarrow 3/xS_x + 3H_2O$ $\Delta H@77^{\circ}F = -264{,}400$ BTU (3)

As shown in Figure 2, heat released in the process is recovered in the SRU waste heat boiler as HP or MP steam, and in the sulphur condensers as LP or LLP steam. In addition to heat released by the Claus reaction, the incinerator produces heat via combustion of fuel gas to achieve temperatures hot enough to ensure complete oxidation of H_2S in the tail gas stream. Incinerators are often equipped with waste heat boilers and/or HP steam superheaters to recover some of this heat and maximize process efficiency.

Steam consumers in the process include feed gas preheaters and process gas reheaters, all of which typically utilize HP saturated steam. The SRU and incinerator air blowers may also consume HP steam if steam turbine drives are employed. The only continuous LP steam consumer is the reboiler in an amine-based TGTU.

Overall, the sulphur recovery facility is a net HP steam exporter. It is typically also an LP steam exporter; however, this may not be the case for amine-based TGTUs with extremely high recovery efficiency requirements, which can consume all (or more) of the LP steam produced in the SRU. This is usually not the case unless sulphur recovery efficiency (SRE) significantly exceeds 99.9%.

Fuel Gas

The incinerator is a continuous fuel gas consumer. Fuel gas is burned with excess air and the combustion effluent is mixed with SRU tail gas to achieve a minimum temperature of 650 °C for nearly complete oxidation of H_2S to SO_2. Sometimes higher temperatures are required to achieve lower limits on CO and/or total reduced sulphur (TRS), up to a maximum of around 815 °C.

In some facilities which process lean acid gas, continuous fuel gas co-firing may be employed in the SRU burner to achieve temperatures high enough for BTEX destruction. Other methods for increasing furnace temperature such as acid gas enrichment or oxygen enrichment are preferred, as they reduce the risk of soot deposition and/or fire in the downstream catalyst beds, as well as minimizing the process gas flow through the facility, thereby minimizing the size of equipment and piping. Nevertheless, fuel co-firing is not an uncommon practice for increasing furnace temperature.

In older facilities, fuel gas is sometimes consumed in SRU fired reheaters; however, most modern SRUs utilize indirect HP steam reheaters to avoid the concerns mentioned above for fuel co-firing in the SRU burner. Most modern amine-based TGTUs employ preheating with HP saturated steam upstream of the hydrogenation reactor. However, for facilities that are not equipped with a hydrogen source, reducing gas generators (RGGs) are often installed. In an RGG, fuel is combusted substoichiometrically to produce reducing gas; the exhaust gas is then mixed with the SRU tail gas to achieve sufficient temperature for the hydrogenation and hydrolysis reactions to occur in the downstream reactor. RGGs result in increased energy consumption (vs. TGTU steam preheaters) due to fuel consumption in the burner and also result in increased process gas flow through all equipment downstream of the RGG.

Overall, the sulphur recovery facility is a net fuel gas importer. All SRUs require continuous fuel firing in the incinerator; however, facilities which employ continuous fuel firing in the SRU burner, reheaters and/or TGTU RGG may require significantly more fuel consumption than units which do not.

Electric Power

The Claus and incinerator blowers are the primary electric power consumers in a sulphur recovery facility, when these machines are equipped with motor drivers. Other power consumers include air-cooled heat exchangers and pumps. In hot climates and/or when extremely high sulphur recovery efficiency is required, refrigeration may be required for solvent and quench water cooling in the TGTU.

Overall, the sulphur recovery facility is a net power importer. Facilities which employ amine-based tail gas treating may utilize significantly more power than those which do not, due to additional air cooled exchangers, pumps and possible refrigeration utilized in those facilities.

ENERGY PERFORMANCE COMPARISON OF SULPHUR RECOVERY TECHNOLOGIES

The overall impact of the various utility producers and consumers described above is that the sulphur recovery facility is typically a net energy exporter, although the quantity of energy exported can vary greatly depending on the type of tail gas treating technology employed. In some cases, the facility may actually need to import energy, when very high sulphur recovery efficiency is required, negating the energy benefits of the Claus process. To illustrate this, a hypothetical 1,000 MTPD sulphur recovery train has been considered, over a range of sulphur recovery efficiencies, which will be referred to as the Benchmark Plant. Considering that most refineries produce rich acid gas (H_2S > 85 mol%) and most gas plants produce relatively lean acid gas (40-50 mol% H_2S), a median concentration of 60 mol% is assumed. Feed gas flow and composition for the hypothetical plant are provided in Table 1.

Table 1. Feedstock for Hypothetical 1,000 MTPD Sulphur Recovery Train (Benchmark Plant)

	mol%	kmol/h
Component		
H_2S	60%	1,300
CO_2	30%	650
Hydrocarbon (as C_1)	1%	22
H_2O	9%	195
Total	100%	2,167
Temperature, °C	54	
Pressure, barg	0.69	

To compare relative energy balances for varying recovery efficiencies, simulations were generated, over a range of tail gas treating technologies and plant configurations. The following SRE cases were explored.

A. **97% SRE** – 97% recovery is based on a conventional 3-stage Claus unit.
B. **99.0% SRE** – 99.0% recovery is based on a sub-dewpoint process (2-stage Claus + 2 sub-dewpoint reactors), although it should be noted that a direct oxidation process would achieve similar SRE and energy balance.
C. **99.3% SRE** – 99.3% recovery is based on a 2-stage Claus unit + TGTU (MDEA). This SRE is just beyond the upper limit of an achievable guarantee value for sub-dewpoint and direct oxidation processes; therefore, it was investigated as the entry point for an amine-based TGTU.
D. **99.9% SRE** – 99.9% recovery is based on a 2-stage Claus unit + TGTU (MDEA).
E. **150 mg SO_2/Nm^3 (MDEA)** – The World Bank Standard case (99.98% SRE) is first investigated based on a 2-stage Claus unit + TGTU with generic solvent (MDEA).
F. **150 mg SO_2/Nm^3 (Proprietary Solvent)** – The World Bank Standard case (99.98% SRE) is investigated utilizing a more selective solvent in the TGTU and corresponding positive energy impact; thus, this case is based on a 2-stage Claus unit + TGTU with proprietary solvent.

Process flow diagrams for the six cases are provided in Figures 3-6.

Figure 3. PFD for Benchmark Plant Case A

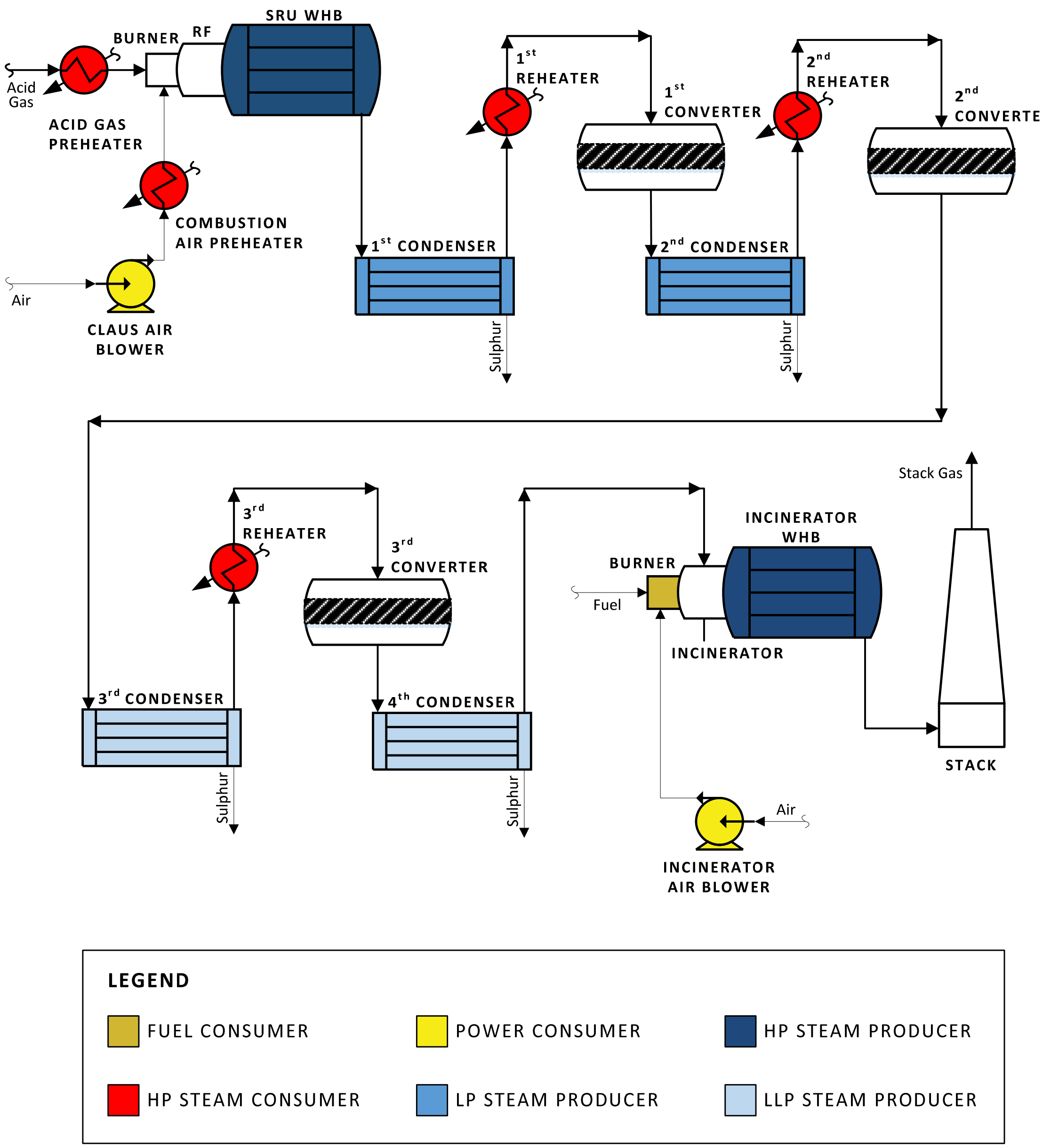


Figure 4. PFD for Benchmark Plant Case B

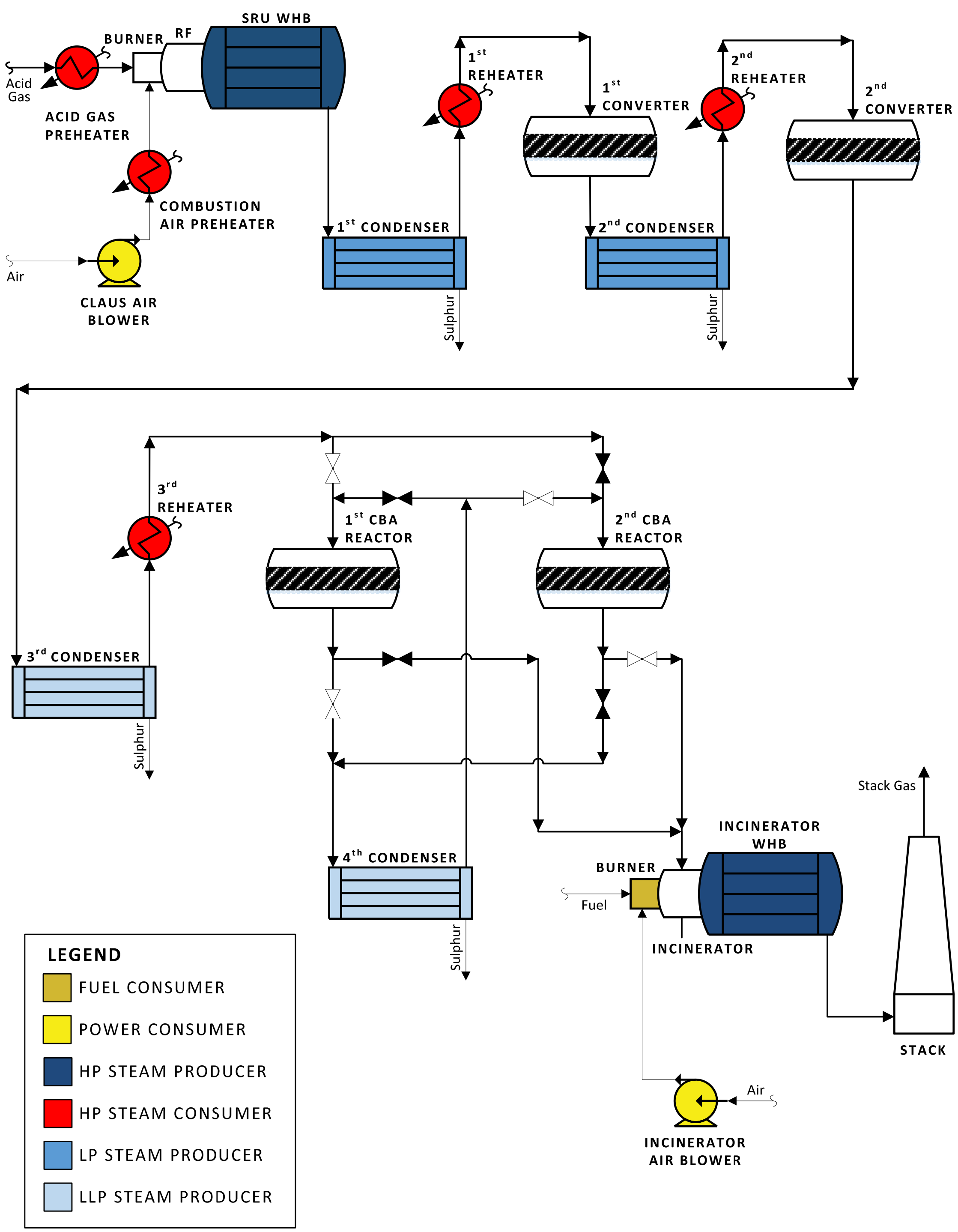


Figure 5. PFD for Benchmark Plant Cases C and D

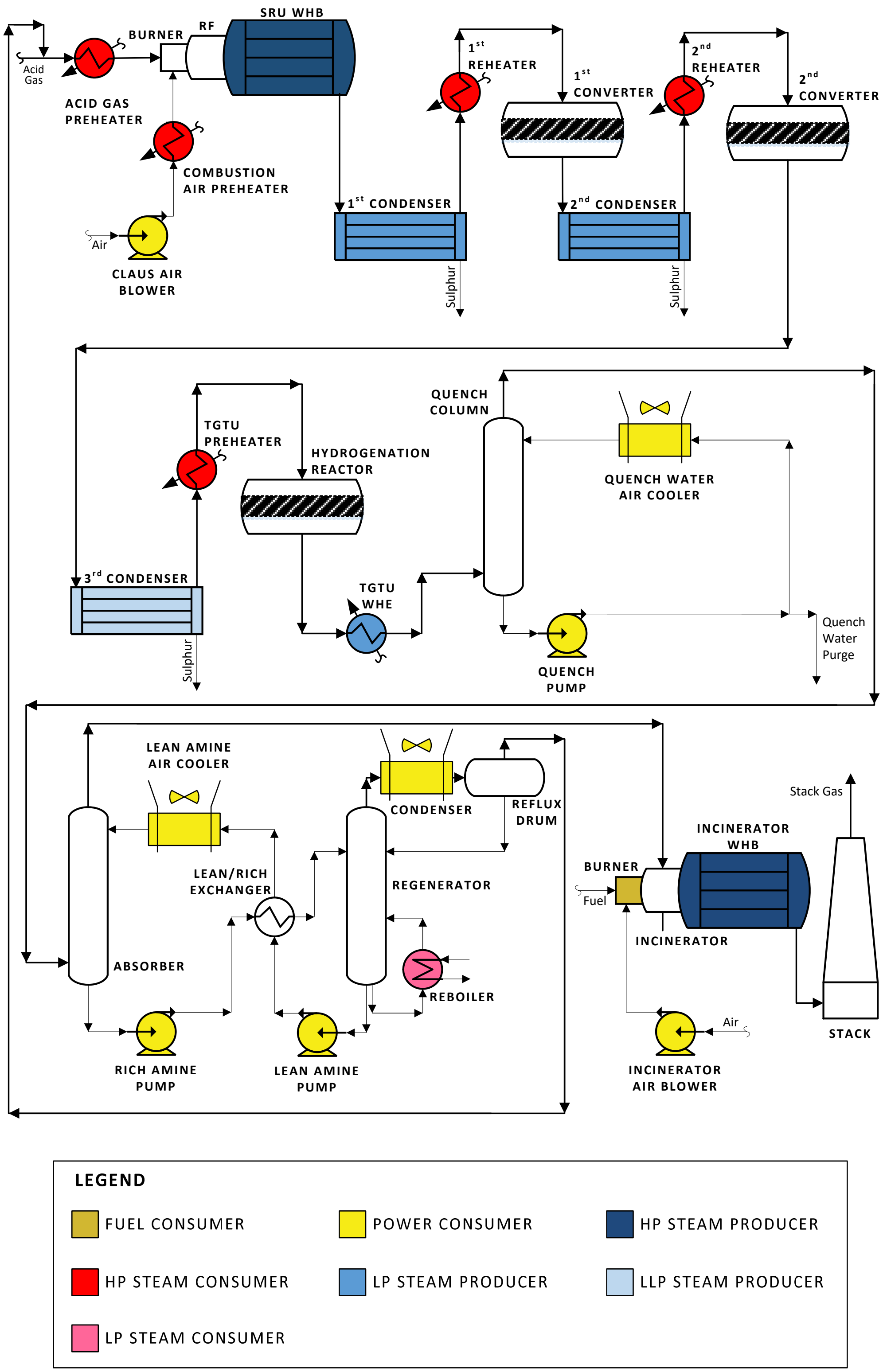
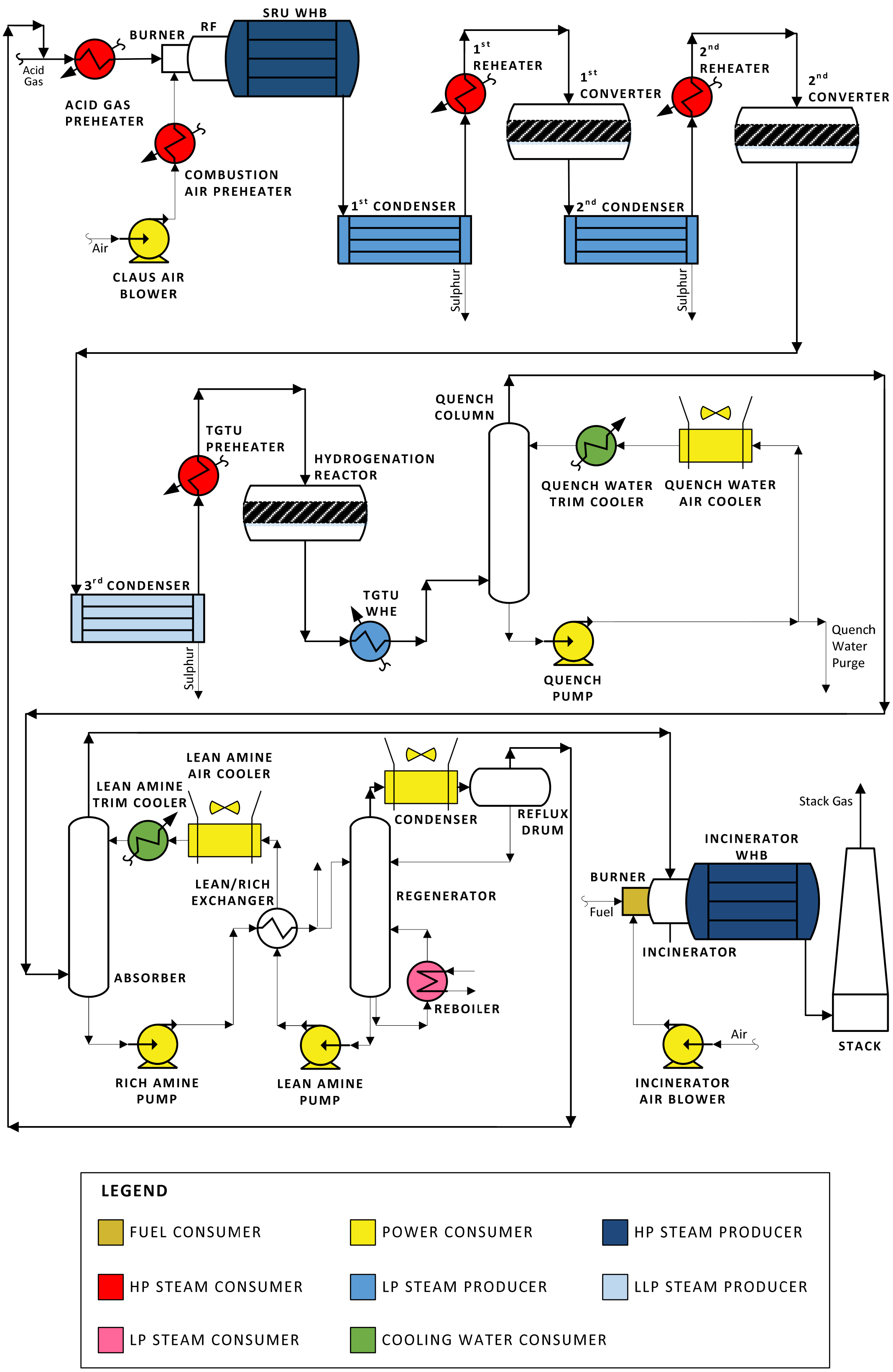


Figure 6. PFD for Benchmark Plant Cases E and F



A standard set of design parameters was employed for all cases to allow relative comparison on a consistent basis. Key design features are aimed at optimizing energy efficiency, as follows:

- Sulphur Recovery Unit
  - Air-only operation, without fuel co-firing
  - Motor-driven Claus air blowers
  - HP saturated steam (40 barg) produced in SRU waste heat boiler (WHB)
  - 2 Claus beds (3 for 97% SRE case) with promoted activated alumina catalyst
  - 2 additional sub-dewpoint beds for 99.0% SRE case
  - LP steam (3.5 barg) produced in 1st & 2nd sulphur condensers
  - LLP steam (1.0 barg) produced in 3rd & 4th sulphur condensers
  - HP saturated steam (40 barg) consumed in SRU preheaters and reheaters

- Tail Gas Treatment Unit
  - HP saturated steam (40 barg) consumed in preheater
  - Low temperature hydrogenation catalyst (232 ^{o}C inlet temperature)
  - LP steam (3.5 barg) produced in TGTU waste heat exchanger (WHE)
  - Lean solvent temperature of 50 ^{o}C for all except Cases E and F, which was reduced to 40 ^{o}C to achieve ultra-high SRE (air cooling to 50 ^{o}C with CW trim cooling)
  - LP steam (3.5 barg) consumed in regenerator reboiler
  - Solvent circulation rate for Case F assumed as 50% of Case E to approximate proprietary solvent[4]

- Incinerator
  - Operated at 815 ^{o}C (upper limit, required for achieving <5 mg/Nm^3 TRS)
  - 2% excess O_2 in stack gas
  - Fuel fired (LHV of 8,953 kcal/Nm^3)
  - Motor-driven incinerator air blowers
  - HP saturated steam (40 barg) produced in incinerator WHB
  - No sulphur pit ejector routed to incinerator

Key process parameters were compared for the range of SRE cases, as summarized in Table 2.

Table 2. 1,000 MTPD Benchmark Plant Process Parameters

CASE **SRE**	**A** **97%**	**B** **99.0%**	**C** **99.3%**	**D** **99.9%**	**E** **99.98%**	**F** **99.98%**
Stack Gas SO_2 (kmol/hr)	38.95	12.97	9.04	1.24	0.28	0.28
Stack Gas CO_2 (kmol/hr)	870.6	885.7	845.4	850.6	851.2	851.2
Total Stack Gas Flow (kmol/hr)	8,012	8,134	6,176	6,188	6,189	6,189
Amine Circulation (m^3/hr)	---	---	176	264	1,026	513

ENERGY BALANCE FOR BENCHMARK PLANT

Energy production/consumption for the top producers and consumers in the Benchmark sulphur recovery facility is summarized in Table 3, for each of the cases studied. It is observed that as SRE increases, energy export decreases, and the facility reverts from net energy export to import at ultra-high recovery efficiency (Case E), primarily due to LP steam consumption for MDEA solvent regeneration in the TGTU. When a highly-selective proprietary solvent is employed (Case F), LP steam consumption is reduced substantially, making the facility essentially energy-neutral.

Table 3. Energy Balance by Top Utility Producers/Consumers for Benchmark Plant (kW)

CASE	A	B	C	D	E	F
SRE	97%	99.0%	99.3%	99.9%	99.98%	99.98%
HP STEAM PRODUCERS						
SRU WHB	+53,921	+54,044	+55,256	+55,586	+55,561	+55,561
Incinerator WHB (815°C)	+40,460	+40,615	+30,379	+30,424	+30,429	+30,429
HP STEAM CONSUMERS						
Acid Gas Preheater	-3,994	-3,994	-3,994	-3,994	-3,994	-3,994
Combustion Air Preheater	-3,271	-3,279	-2,900	-2,916	-2,918	-2,918
1st & 2nd SRU Reheaters	-7,376	-7,210	-7,604	-7,638	-7,678	-7,678
3rd SRU Reheater	-2,352	-1,350	---	---	---	---
TGTU Reactor Preheater	---	---	-4,382	-4,401	-4,423	-4,423
LP/LLP STEAM PRODUCERS						
1st & 2nd Sulphur Condensers	+19,544	+19,620	+20,255	+20,353	+20,431	+20,431
3rd Sulphur Condenser	+4,954	+4,092	+5,282	+5,307	+5,329	+5,329
4th Sulphur Condenser	+2,677	+4,185	---	---	---	---
TGTU Hydrogenation WHE	---	---	+3,592	+3,609	+3,622	+3,622
LP STEAM CONSUMERS						
Regenerator Reboiler	---	---	-12,266	-18,610	-86,820	-43,410
MP/LP FUEL CONSUMERS						
Incinerator Burner	-43,838	-47,147	-38,261	-39,426	-39,572	-39,572
MAJOR ELECTRIC POWER CONSUMERS						
Claus Air Blowers	-1,516	-1,520	-2,001	-2,012	-2,014	-2,014
Quench Pumps	---	---	-75	-75	-75	-75
Amine Pumps	---	---	-83	-126	-486	-243
Quench Water Air Cooler	---	---	-102	-153	-593	-296
Lean Amine Air Cooler	---	---	-68	-105	-274	-137
Regenerator Ovh'd Condenser	---	---	-62	-101	-588	-294
Incinerator Air Blowers	-246	-247	-196	-196	-196	-196
MAJOR COOLING WATER CONSUMERS						
Quench Water Trim Cooler	---	---	---	---	-17,366	-8,683
Lean Amine Trim Cooler	---	---	---	---	-9,937	-4,969
ENERGY BALANCE						
Net Energy Import/Export	+58,963	+57,809	+42,770	+35,627	-60,974	-3,236
Comparison to Case A (Net Δ)	---	-2%	-27%	-40%	-203%	-105%

The energy balance for the various cases is summarized by unit operation in Table 4, which clearly illustrates that the SRU is always a net energy exporter whose energy production remains fairly constant, even for Case B, which employs a non-amine-based tail gas treating technology to achieve higher SRE. It is the amine-based TGTU that is responsible for increasing energy consumption as SRE increases.

Highly-selective solvents can offer reductions in LP steam and solvent cooling requirements but the overall impact of the TGTU on sulphur recovery energy production is still significant. The world average sulphur recovery efficiency for new plants is around 99.9% (Case D), which results in a 40% energy penalty on the standalone Claus plant, as shown in Tables 3 and 4. Despite its detrimental energy impact on the overall SRU/TGTU energy balance, the amine-based tail gas treating process is currently the only technology available for achieving guaranteed SRE in excess of about 99.3%.

Table 4 also shows that incinerator energy consumption increases slightly as SRE increases, simply due to a reduced quantity of H_2S in the tail gas. This results in lower tail gas heating value and therefore greater fuel requirements, although the overall impact is marginal.

Table 4. Benchmark Plant Energy Balance by Processing Unit (kW)

CASE SRE	A 97%	B 99.0%	C 99.3%	D 99.9%	E 99.98%	F 99.98%
SRU	+62,587	+64,588	+64,294	+64,686	+64,717	+64,717
TGTU	---	---	-13,446	-19,962	-116,940	-58,908
INCINERATOR	-3,624	-6,779	-8,078	-9,198	-9,339	-9,339
NET	+58,963	+57,809	+42,770	+35,627	-60,974	-3,236

ENERGY KEY PERFORMANCE INDICATORS (KPIs)

The net energy balance figures provided in Tables 3 and 4 are converted to “thumb rule” targets that can be used to assess a sulphur recovery facility’s energy performance, as provided in Table 5. These key performance indicators (KPIs) can be utilized by operators to evaluate whether their facilities are operating in accordance with best energy efficiency standards. While such metrics may not have been viewed as particularly important previously, the authors are observing an increasing trend, in this low oil price climate, of operators wishing to make the best use of the sulphur recovery unit’s energy benefits. As a result, SRU/TGTU energy efficiency is being increasingly evaluated and scrutinized, particularly for those sour facilities with relatively large sulphur recovery requirements.

Table 5. Energy Performance KPIs for Benchmark Plant

CASE SRE	A 97%	B 99.0%	C 99.3%	D 99.9%	E 99.98%	F 99.98%
kWh per Metric Ton ‘S’ Produced	+1,458	+1,401	+1,033	+853	-1,477	-85
kWh per Nm^3 H_2S in Acid Gas Feed	+2.02	+1.98	+1.47	+1.22	-2.00	-0.12

It is important to keep in mind specific feed conditions and plant design configuration when applying this information. A different configuration, feedstock and/or operating philosophy can lead to significant variations in KPIs, as we will see in the Case Study which follows. For example, a plant which is equipped with an incinerator WHB can generate up to 40% more HP steam than one that is not, as illustrated in Table 3. Some other examples that can lead to widely varying KPIs include fuel gas co-firing in the SRU, the use of an RGG in the TGTU, installation of low-temperature catalyst in the TGTU and TGTU solvent chilling requirements, to name a few.

For the most part, the energy efficiency of the Benchmark Plant design has been optimized across all Cases A-F, with the exception of incinerator operating temperature, which could be reduced by around 165 ^{o}C, depending on the emission regulations. However, since an incinerator WHB is employed in the Benchmark Plant design, additional waste heat is recovered at the higher temperature and the overall impact on energy efficiency is negligible. There are some other minor opportunities for improving energy efficiency that are not included in the Benchmark Plant design, such as BFW preheat in the final condenser and sulphur cooler to maximize HP steam production (rather than generating LLP steam), but the overall impact on energy export/import for these items is not expected to significantly impact the results of this study.

CASE STUDY - LARGE SULPHUR RECOVERY FACILITY IN THE MIDDLE EAST

To illustrate the potential energy enhancements that can be gained by scrutinizing plant operation to ensure efficient operation, a specific Case Study is presented for a large sulphur recovery facility in the Middle East. This Real World Plant is compared to the Benchmark Plant KPIs to assess its energy performance.

Case Study Facility Description

In this example, the capacity of a single SRU/TGTU train is 1,300 MTPD, processing lean acid gas with H_2S content less than 50 mol%. Natural gas co-firing is employed in the SRU burner to ensure sufficient temperature for BTEX destruction. The recovery efficiency specification is slightly greater than 99.9%. A highly-selective proprietary solvent is utilized in the TGTU and propane refrigeration is required to achieve a lean solvent temperature of 49 ^{o}C, due to high ambient temperatures that do not allow air cooling below 59 ^{o}C. The Real World Plant Case Study is most similar to Case D from the hypothetical Benchmark Plant analysis, with key differences as highlighted in bold, italic, underlined font in Table 6.

Table 6. Comparison of Benchmark Plant (Case D) to Real World Plant Configuration

	Benchmark Plant (Case D)	**Real World Plant (Design Case)**
CAPACITY & KEY DESIGN CONDITIONS	• 1,000 MTPD • 60 mol% H_2S • 99.90% SRE • Air cooling process temp. – 50 °C (min)	• 1,300 MTPD • 46 mol% H_2S • 99.91% SRE • Air cooling process temp. – 59 °C (min)
SRU	• Air-only operation, without fuel co-firing • Motor-driven Claus air blowers • HP saturated steam (40 barg) produced in SRU WHB • 2 Claus beds with promoted activated alumina catalyst • LP steam (3.5 barg) produced in 1st & 2nd sulphur condensers • LLP steam (1.0 barg) produced in 3rd sulphur condenser • HP steam (40 barg) consumed in SRU preheaters and reheaters	• Air-only operation, ***with fuel co-firing*** • Motor-driven Claus air blowers • HP saturated steam (***41.5 barg***) produced in SRU WHB • 2 Claus beds with promoted activated alumina catalyst – ***TiO_2 layer in 1st bed*** • LP steam (***5.5 barg***) produced in 1st & 2nd sulphur condensers • LLP steam (1.0 barg) produced in 3rd sulphur condenser • HP steam (***41.5 barg***) consumed in SRU preheaters and reheaters, ***except for air preheater which is fuel fired***
TGTU	• HP saturated steam (40 barg) consumed in preheater • Low temperature hydrogenation catalyst • LP steam (3.5 barg) produced in TGTU WHE • MDEA solvent • Lean solvent temperature of 50 °C, achieved via air cooling • LP steam (3.5 barg) consumed in regenerator reboiler	• ***RGG (rather than steam preheater) in TGTU*** • ***Conventional hydrogenation catalyst*** • LP steam (***5.5 barg***) produced in TGTU WHE • ***Highly-selective, proprietary solvent*** • Lean solvent temperature of 49 °C, ***achieved via air cooling (to 59 °C) and refrigerant trim cooling (to 49 °C)*** • LP steam (***5.5 barg***) consumed in regenerator reboiler
INCINERATOR	• Operated at 815 °C (upper limit, required for achieving <5 mg/Nm^3 TRS) • 2% excess O_2 in stack gas • Fuel fired (LHV of 8,953 kcal/Nm^3) • Motor-driven incinerator air blowers • HP saturated steam (40 barg) produced in incinerator WHB • No sulphur pit ejector routed to incinerator	• Operated at 815 °C (upper limit, required for achieving <5 mg/Nm^3 TRS) • ***2.5%*** excess O_2 in stack gas • Fuel fired (LHV of 8,953 kcal/Nm^3) • Motor-driven incinerator air blowers • ***HP steam from SRU WHB superheated to 400 °C in incinerator WHB superheater*** • ***Sulphur rundown vessel vent routed to incinerator***

The most significant differences between Case D and the Real World Plant are as follows:

1. **Fuel gas consumption** – The Real World Plant utilizes fuel gas in three services where Case D does not; a fired combustion air preheater, continuous co-firing in the SRU burner and an RGG in the TGTU. Moreover, the RGG operates at 40 °C higher discharge temperature than the steam preheater in Case D due to the use of conventional hydrogenation catalyst in the downstream reactor. These additional fuel users will have an undesirable impact on the Real World Plant energy KPIs in relation to Case D. A slightly higher excess oxygen concentration in the incinerator stack gas (0.5 mol%) will also marginally increase fuel consumption for the Real World Plant.

2. **HP steam production** – Case D assumes saturated HP steam production in the SRU and incinerator WHBs, while the Real World Plant generates HP saturated steam in the SRU WHB but only employs

an HP steam superheater coil in the incinerator, with no downstream HP saturated steam generation. Because the process gas temperature leaving the superheater coil is approximately 400 ^{o}C warmer than from the exit of the incinerator WHB for Case D, overall heat recovery, and therefore HP steam export, will be negatively impacted.

A slightly higher HP steam pressure will result in a reduction in HP steam production for the Real World Plant; however, the energy content of higher pressure steam will be greater and therefore this is not expected to impact the energy KPIs. This statement is also true for the differences in LP steam pressure.

3. **LP steam consumption** – Case D employs generic MDEA in the TGTU, while the Real World Plant utilizes a highly-selective proprietary solvent. The proprietary solvent should result in a solvent circulation rate reduction on the order of 50% of the MDEA value (for the same capacity and recovery efficiency), with corresponding LP steam reduction in the reboiler. However, it should be noted that the concentration of the solvent was only about 55% of the design value at the time of operating data collection, which results in lower LP steam consumption for the same circulation rate (due to lower acid gas loading).

4. **Power consumption** – Case D achieves the required lean solvent temperature of 50 ^{o}C via air cooling, while the Real World Plant requires air cooling (to 59 ^{o}C) followed by trim cooling (to 49 ^{o}C) due to hot ambient temperatures. Power consumption associated with refrigeration will increase the overall power import requirement for the Real World Plant.

5. **Sulphur recovery efficiency (SRE)** – Case D is based on 99.9% SRE, while the Real World Plant requires 99.91% to meet the stack SO_2 specification. While this may seem like a minor difference, TGTU solvent circulation requirements can increase disproportionately due to reduced H2S partial pressure in the tail gas. Because of this difference in SRE, combined with significant differences in ambient conditions, the TGTU for the Real World Plant cannot be compared on an exactly consistent basis with Case D.

6. **Process gas flowrate** – for the Real World Plant Design Case there are multiple factors which will increase the process gas flowrate through the unit versus Case D. The differences in capacity and acid gas composition will be compensated by evaluating KPIs on a 'per ton of S' or 'per Nm^3 of H_2S' basis and therefore should not have a significant impact. However, other differences such as continuous fuel co-firing in the SRU, and the employment of an RGG in the TGTU will dilute the concentration of reactants in the process gas and increase the volumetric flow through the unit. The additional flow may lead to increased steam generation in some cases, but likely at the expense of additional energy consumption required for heating or cooling a larger volumetric gas flow rate.

Other differences between the Real World Plant and Benchmark Case D are minor and not expected to have a substantial impact on the comparison of energy KPIs between the two.

Case Study Basis

The primary purpose of the Case Study is to evaluate energy performance of the Real World Plant and suggest changes to achieve an expected operating condition that is as close to the Hypothetical Benchmark Plant (Case D) energy KPIs as possible. To do this, it was first necessary to build the simulations listed below.

Design Case Simulation based on design information

Actual Case Simulation based on daily average DCS data for a selected typical representative operating condition

Expected Case Simulation based on optimized energy performance of the existing hardware at the selected representative operating condition

The top utility consumers/producers for the HP steam, LP steam, fuel gas and electrical power systems were included in the study, comprising ~98% of all energy utilization in the unit. The key parameters for each of the above cases are provided in Table 7. It should be noted that a previous study was conducted to determine the minimum fuel gas co-firing rate to ensure BTEX combustion and this is reflected in the Expected Case. Since the time of the study, the unit has been operating at a higher co-firing rate for unknown reasons.

Table 7. Process Parameters for Real World Plant Case Study

	Design	Actual	Expected
Acid Gas Feed (Nm^3/h)	82,035	50,208	50,208
Acid Gas H_2S Content (mol%)	46.15	49.95	49.95
Fuel Gas Co-firing (Nm^3/h)	3,452	1,960	300
Sulphur Production (MTPD)	1,300	860.5	860.6
Solvent Circulation Rate (% of design)	---	100%	100%
Solvent Concentration (% of design)	---	55%	100%

Case Study Methodology

Design Case energy production/consumption values were obtained from the project design basis, H&MB and utility summary documents. To determine the energy values for the other cases, certain assumptions were made to simulate current actual performance and modifications were applied to simulate future expected performance.

The following assumptions were applied to the Design Case simulation to approximate current plant performance and produce the Actual Case simulation:

- Acid gas composition based on recent sample analysis
- Fuel gas composition based on recent sample analysis
- Air composition based on design heat and material balance
- Plant DCS data matched (as close as possible), except where measured values were obviously incorrect

The following changes were applied to the Actual Case to achieve optimal energy performance with the existing hardware and produce the Expected Case simulation:

- Temperature of acid gas pre-heat increased by 9 ^oC to match design value of 240 ^oC
- Temperature of combustion air pre-heat increased by 17 ^oC to match design value of 360 ^oC
- Fuel gas co-firing reduced by approximately 1,600 Nm^3/h, with subsequent 60 ^oC reduction in reaction furnace temperature; temperature still adequate for BTEX destruction
- Outlet temperature of 2nd reheater reduced 4 ^oC based on a sulphur dewpoint margin of 15°C
- RGG burn stoichiometry reduced by 1.7% to match design value of 84.6%
- Stack gas O_2 content reduced by approximately 5 mol%, to match design value of 2.5 mol%
- Incinerator operating temperature reduced to 568 ^oC to match Actual Case (it is unknown whether adequate emissions values are achieved at this temperature)
- Lean solvent temperature increased by 5 ^oC (note that past performance has shown that lean amine temperature can be increased but this will require a separate study; increase of 5 ^oC has been assumed for the purpose of this evaluation)
- TGTU solvent concentration increased from 55% to 100% of design value

Case Study Results

Tables 7 and 8 indicate the net energy balance for the Design, Actual and Expected Cases for the Real World Plant, by major equipment and by unit, respectively. The net energy consumption of the Actual Case is only about 25% of the Design Case. This is partially because the plant is operating at only about 66% of the design rate; however, this does not fully account for the energy reduction, and thus this point will be examined in detail. The Expected Case results in more than a 100% reduction in energy consumption versus the Design Case, which converts the plant from a net energy consumer to a net energy exporter.

Table 8. Energy Balance by Top Utility Producers/Consumers for Real World Plant (kW)

	Design	Actual	Expected
SATURATED / SUPERHEATED HP STEAM PRODUCERS			
SRU WHB	+102,148	+68,079	+58,088
HP Steam Superheater	+14,876	+6,225	+3,472
HP STEAM CONSUMERS			
Acid Gas Preheater	-7,857	-4,628	-4,801
1st & 2nd SRU Reheaters	-8,101	-5,187	-4,421
LP/LLP STEAM PRODUCERS			
1st & 2nd SRU Sulphur Condensers	+30,776	+18,606	+16,509
TGTU Hydrogenation WHE	+14,086	+9,056	+7,494
LP STEAM CONSUMERS			
Regenerator Reboiler	-37,482	-19,819	-19,819
MAJOR ELECTRIC POWER CONSUMERS			
Claus Air Blowers	-3,533	-3,087	-3,040
Quench Pumps	-110	-68	-68
Amine Pumps	-235	-156	-156
Quench Water Air Cooler	-395	-33	-22
Lean Amine Air Cooler	-265	-19	-19
Regenerator Overhead Condenser	-196	-68	-68
Refrigerant Compressor	-5,778	-6,995	-4,615
Refrigerant Condenser	-660	-246	-510
Incinerator Air Blowers	-1,231	-1,013	-360
MP/LP FUEL CONSUMERS			
Combustion Air Preheater	-21,742	-10,210	-10,004
Acid Gas Burner	-34,510	-19,600	-3,000
Reducing Gas Generator Burner	-19,500	-10,160	-9,203
Incinerator Burner	-97,725	-41,220	-20,176
ENERGY BALANCE			
Net Energy Export/Import	-77,434	-20,543	+5,281
Comparison to Design Case (Net Δ)	---	+73%	+107%

Table 9. Energy Balance by Processing Unit for Real World Plant (kW)

	Design	Actual	Expected
SRU	+57,181	+43,973	+49,331
TGTU	-50,535	-28,508	-26,986
INCINERATOR	-84,080	-36,008	-17,064
NET	-77,434	-20,543	+5,281

Table 9 illustrates that the largest contributors to reduced energy consumption for the Actual and Expected Cases occurs in the TGTU and incinerator. The reasons for this will be investigated in detail.

The energy consumption figures in Tables 8 and 9 are converted to Key Performance Indicators (KPIs) in Table 10 and Figure 7, which can be used for comparison with Benchmark Plant Case D. Although the KPIs are greatly improved from the Design Case to the Expected Case, the Real World Plant still falls short of the Benchmark Case D KPIs, for reasons which can be explored by looking at the major utility producers/consumers in more detail in the discussion that follows.

Table 10. Energy Performance KPIs for Real World Plant

Utility	kWh per Metric Ton of 'S'			kWh per Nm^3 of H_2S		
	Design	Actual	Expected	Design	Actual	Expected
HP Steam	1,865	1,799	1,460	2.67	2.57	2.09
LP Steam	136	219	-23	0.19	0.31	0.03
Fuel Gas	-3,202	-2,264	-1,182	-4.58	-3.24	-1.69
Power	-221	-337	-250	-0.54	-0.67	-0.56
TOTAL	**-1,421**	**-584**	**5**	**-2.24**	**-1.04**	**-0.56**
Comparison to Benchmark Case D	**853**			**1.22**		

Figure 7. Overall Energy Performance KPIs for Real World Plant vs. Benchmark Case D

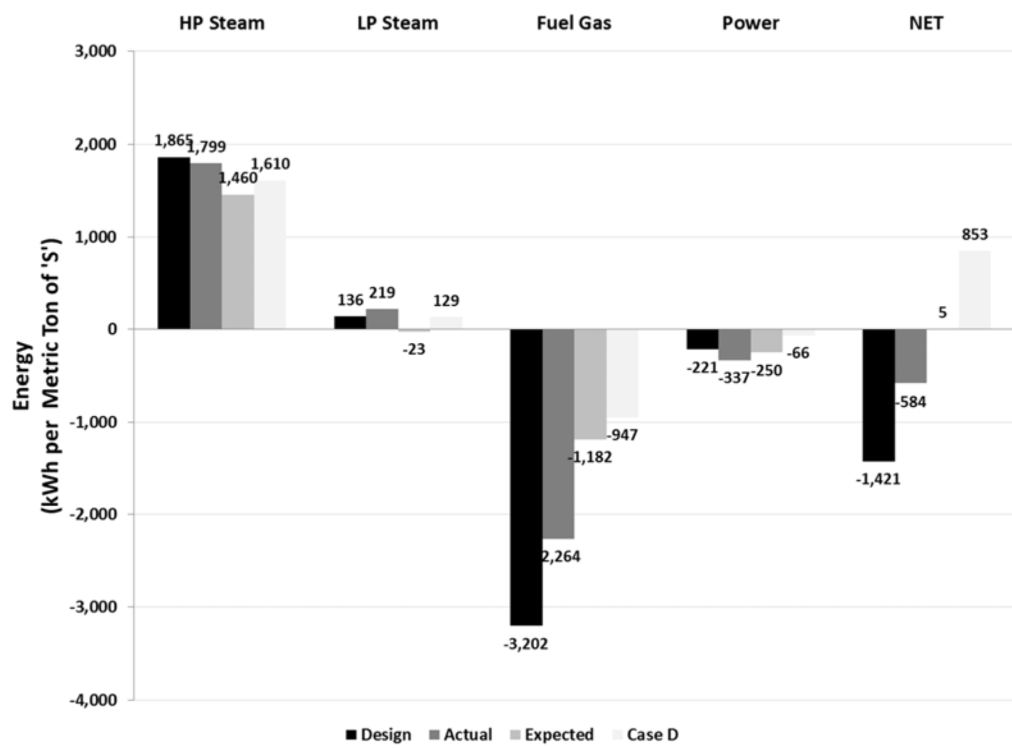


Figure 8 illustrates fuel gas consumption KPIs by major equipment and shows that fuel consumption in the combustion air preheater, co-firing in the SRU burner and RGG contribute significantly to overall fuel gas requirements in the Real World Plant. For the Benchmark Plant Case D, these fuel gas consumers are not required, resulting in a considerably lower fuel consumption KPI. In all cases, the incinerator is a significant fuel consumer.

For the Expected Case, fuel consumption in the SRU burner was reduced in accordance with the results of BTEX destruction testing, resulting in a significant improvement in the KPI for that equipment item.

Incinerator fuel consumption was significantly reduced for the Expected Case due to substantial reductions in operating temperature and % excess O_2 in the stack gas.

Combustion air preheater and RGG fuel consumption remain largely unchanged for the Expected Case due to the fact that they are already operating as efficiently as possible, which is an improvement over the Design Case.

It is noteworthy that the Case D incinerator fuel consumption KPI is higher than for the Expected Case. This is due to the fact that the Expected Case incinerator operating temperature was reduced to 568 ^{o}C in accordance with current operation, while the Case D incinerator operating temperature is 815 ^{o}C, which is the upper limit. Despite a higher operating temperature, most of the extra energy consumed in the incinerator in Case D is recovered as HP steam, which is evident in Figure 9 and the discussion that follows.

Figure 8. Fuel Gas KPIs for Real World Plant vs. Benchmark Case D

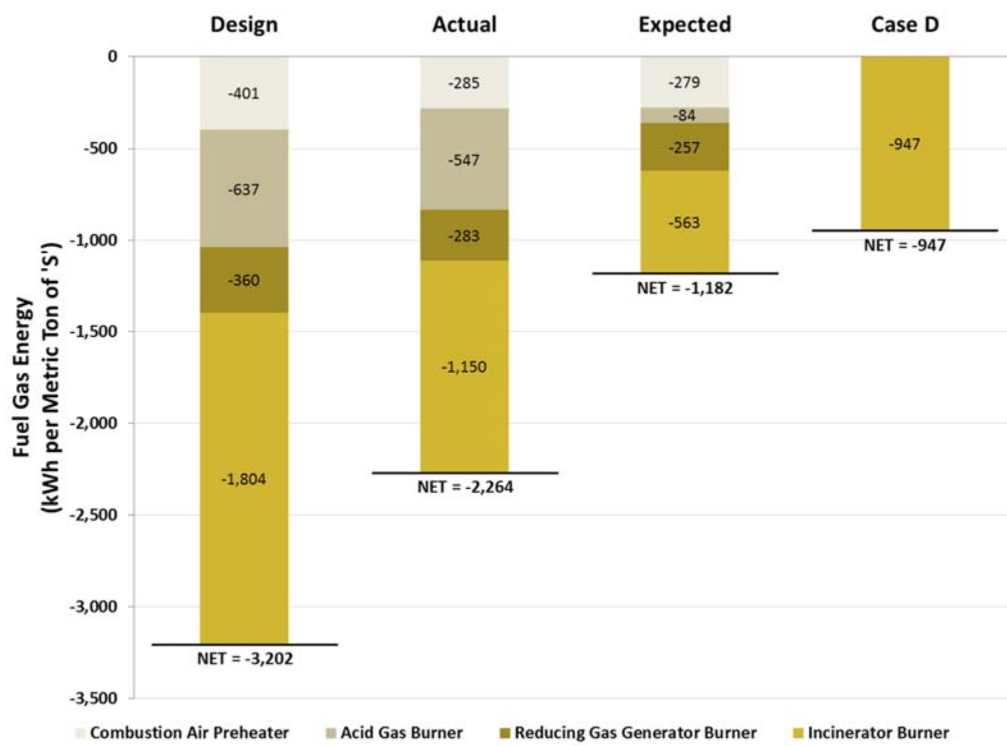


Figure 9 illustrates HP steam KPIs by major producer/consumer and shows that the SRU WHB contributes significantly to HP steam production. The incinerator is also potentially a significant contributor to HP steam export energy but this depends on the configuration of the incinerator waste heat recovery system.

For the Expected Case, HP steam production in the SRU WHB is reduced due to the reduction in SRU burner co-firing described above. Case D does not employ SRU co-firing and also has a reduced combustion air preheat temperature; thus, SRU WHB steam production is decreased further. Nevertheless, it is still the largest contributor to HP steam production.

The Real World Plant is equipped with only an HP steam superheater coil downstream of the incinerator. The energy KPI for this exchanger decreases for the Actual and Expected Cases due to a lower incinerator operating temperature and reduced HP steam export from the SRU WHB. For the Benchmark Plant Case D, maximum waste heat recovery in the incinerator WHB greatly increases total HP steam export and has a positive impact on the overall HP steam KPI.

While the Case D overall HP steam production KPI is greater than the Expected Case, it is lower than the Design Case. The reason for this is that there are two additional HP steam consumers for Case D – the combustion air preheater and TGTU preheater. In the Real World Plant, these two services utilize fuel gas rather than HP steam, which is less efficient from an energy perspective. This is illustrated via comparison of energy KPIs shown in Figures 8 and 9 for these two services.

Figure 9. HP Steam KPIs for Real World Plant vs. Benchmark Case D

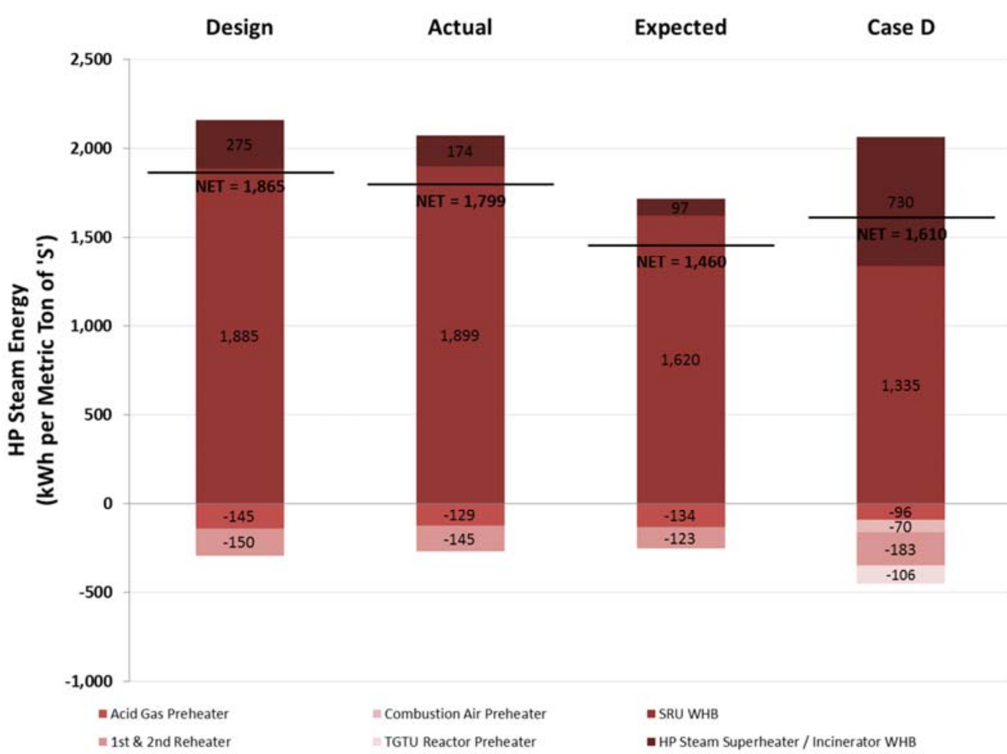


Figure 10 illustrates LP steam KPIs by major producer/consumer and shows that the TGTU reboiler contributes significantly to LP steam consumption. LP steam production in the sulphur condensers and TGTU WHE typically exceeds the reboiler requirement, but this may not always be the case for ultra-high SRE, as illustrated in Benchmark Case E.

For the Real World Plant Actual Case, the reboiler LP steam consumption KPI is reduced from the Design Case, which is a result of the lower-than-design solvent concentration; lower acid gas loading results in reduced steam consumption for the same circulation rate. Low solvent concentration is compensated by reducing lean solvent temperature by 5-7 ^{o}C, which increases refrigeration requirements as indicated in Figure 11. For the Expected Case, it is assumed that solvent concentration is increased to the design value and therefore the LP steam consumption KPI increases and refrigeration KPI decreases.

The Case D (MDEA) LP steam KPI for the reboiler is lower than for the Expected Case (proprietary solvent). This is likely because the Real World Plant circulation rate contains considerable design margin, which is evidenced by the fact that the unit is currently achieving around 99.95% SRE, even at low solvent concentration. If the circulation rate for the Real World Plant was reduced to just meet an SRE of 99.9%, the circulation rate, and LP steam consumption, would be significantly lower than for Case D.

The LP steam production KPI for the sulphur condensers is reduced for the Expected Case due to minimum co-firing in the SRU burner, which results in lower process gas flow through the unit. The net energy impact of reduced co-firing is beneficial, when fuel consumption is taken into account. Because the acid gas H_2S concentration for Case D is higher than for the Expected Case (60 mol% vs. 46 mol%), the concentration of reactants throughout the process is higher, resulting in higher temperatures and a slight improvement for the LP steam KPI for the sulphur condensers.

LP steam production in the TGTU WHE is lower for Case D than for the Real World Plant due to the use of low temperature hydrogenation catalyst, which also results in less energy consumption in the upstream preheater; the net impact is a reduction in energy consumption.

The net result of the above is that the overall LP steam production KPI for Case D is slightly greater than for the Real World Plant Expected Case.

Figure 10. LP Steam KPIs for Real World Plant vs. Benchmark Case D

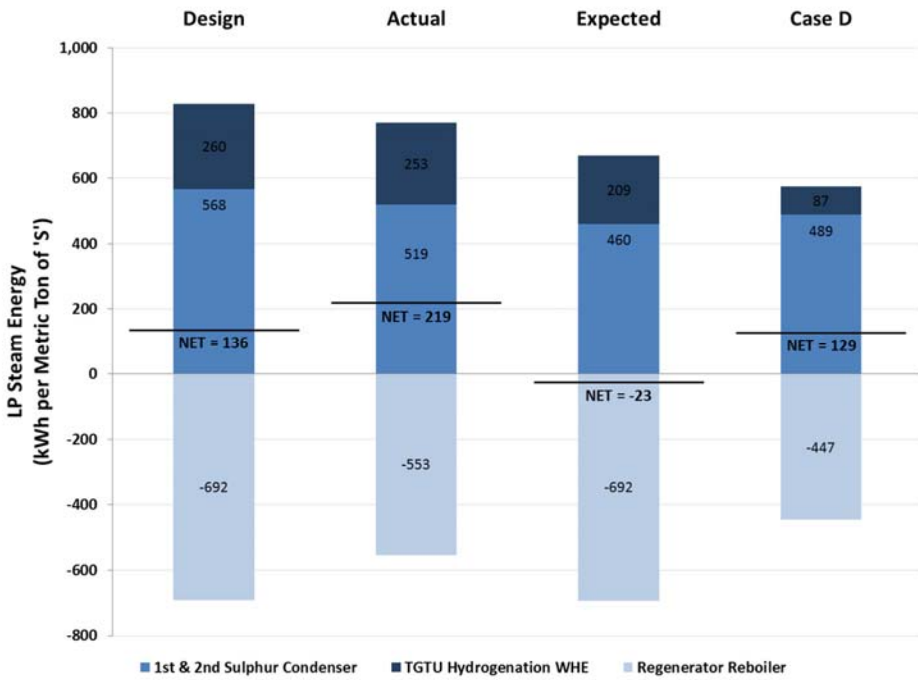


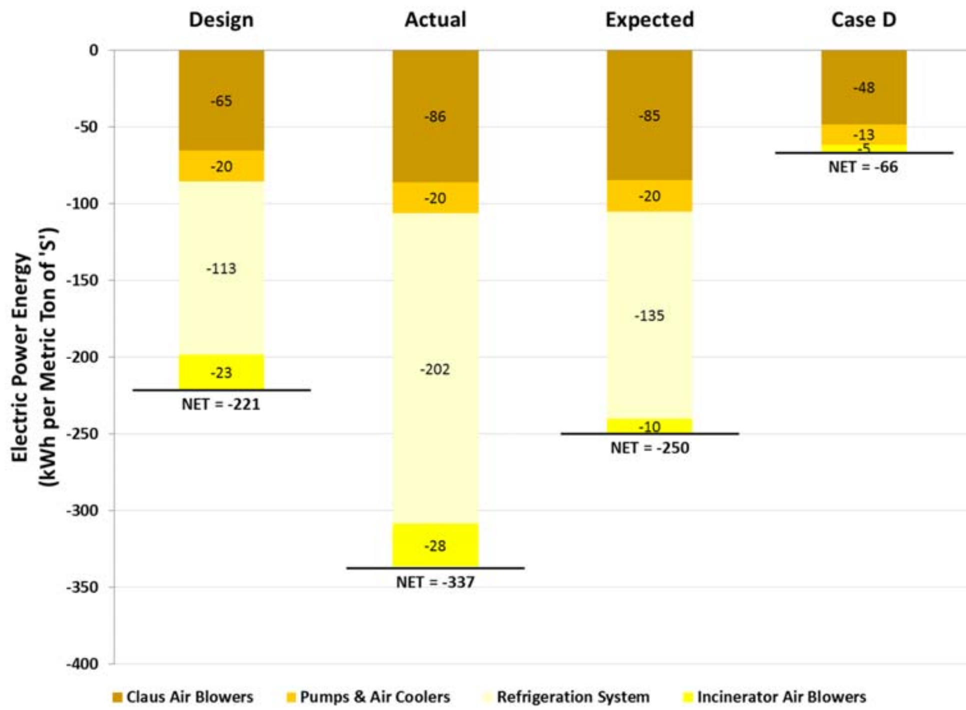
Figure 11 illustrates electric power consumption KPIs by major equipment and shows that the refrigerant compressor contributes most significantly to power consumption, followed by the Claus combustion air blowers. Case D does not require refrigeration due to lower ambient temperatures that allow process cooling to 50 ^{o}C with air cooling alone.

In going from the Design to Actual Case for the Real World Plant, refrigeration power consumption increases due to the fact that the solvent is cooled 5-7 ^{o}C below the design temperature to compensate for lower-than-design solvent concentration. Also, the refrigeration compressor runs at full capacity, despite a 33% reduction in plant throughput, which negatively impacts this KPI. For the Expected Case, the refrigeration KPI improves due to a reduction in load on the system, as the solvent concentration is increased to the design value and lean solvent temperature is increased 5 ^{o}C over the design value.

For the Real World Plant, the Claus combustion air blower power KPI increases for the Actual and Expected Cases, due to the fact that the blowers operate at full capacity despite a reduction in throughput. For Case D, the Claus combustion air blower power consumption KPI is reduced due to no co-firing in the SRU and higher acid gas H_2S concentration.

The net result of the above is that the electric power consumption KPI for Case D is substantially lower than that which can be achieved by the Real World Plant.

Figure 11. Electrical Power KPIs for Real World Plant vs. Benchmark Case D



Case Study Summary

The key operational changes listed below can be implemented in the Real World Plant to convert it from a significant energy consumer (Actual Case) to a net energy producer (Expected Case), without implementing any hardware changes.

E1. Reduce fuel co-firing in the SRU burner
E2. Reduce fuel consumption in the incinerator by reducing excess air
E3. Increase TGTU solvent concentration (match design concentration)
E4. Increase lean solvent temperature (5 °C assumed – higher value may be possible)

When comparing the Real World Plant to the Benchmark Cases explored in Table 5, the conditions should be closest to Case D; however, it is observed that there is a significant difference between the two KPIs. This is primarily due to differences in ambient conditions, feedstock and plant configuration. If the Real World Plant were modified to approximate the Benchmark Plant (Case D) energy performance KPIs, the following hardware changes would need to be implemented:

B1. Increase H_2S in acid gas to ~60 mol% (modify AGRU or install AGE; associated energy impact)
B2. Eliminate fuel co-firing in the SRU burner
B3. Replace fired combustion air preheater with HP saturated steam preheater
B4. Replace TGTU RGG with HP saturated steam preheater (requires H_2 source for upsets)
B5. Replace conventional tail gas hydrogenation catalyst with low temperature catalyst
B6. Install incinerator WHB to recover waste heat; minimum process gas outlet temperature ~350 °C

B7. Carry out refrigeration study to determine specific limits of the facility; it may be possible to increase solvent temperature and/or to turndown or shut down refrigeration system if solvent concentration is increased to design value

A cost-benefit analysis would need to be carried out for each of the above recommendations to determine whether the benefits of implementing the modification would justify the cost. Such a study has not yet been performed.

ENERGY EXPORT POTENTIAL FROM UAE SULPHUR RECOVERY FACILITIES

In 2016 the Middle East became the world's largest sulphur producing region. Current UAE sulphur production is approximately 19,000 MTPD (7 MMTPA), equivalent to roughly 11% of global production. As indicated in Table 5, for every ton of sulphur produced, a Claus SRU is capable of exporting HP and LP steam equivalent to approximately 1,400 kWh of energy. Approximately 75% of this energy (1,050 kWh) is exported as HP steam. Thus, current thermal power potential from the UAE's 19,000 MTPD of sulphur production is around 830 MW. Assuming a steam turbine efficiency of 60% gives 500 MW of mechanical power generation potential from all UAE SRUs.

As indicated by the Benchmark Plant analysis in Table 3, amine-based tail gas treating may consume between 40 and 100% of the energy produced by the SRU. For a 99.9% SRE facility utilizing a proprietary solvent, in a moderate climate, the TGTU should consume roughly 50% of the energy produced by the SRU. Thus, **the overall power generation from UAE SRUs should currently be in the range of 250 MW** (or greater, as not all UAE SRUs are equipped with amine-based tail gas treaters). This energy would be extremely beneficial for supplementing the power requirements of the sour gas processing facilities in which they are installed. However, as illustrated by the Real World Plant Case Study, certain factors may be eroding these SRU energy benefits in some cases. It may be possible to implement operational and/or hardware changes aimed at restoring some of the energy subsidies offered by these plants, but physical plant modifications to existing facilities would need to be justified with appropriate cost/benefit analyses.

Within the next decade, UAE sulphur production could double, which presents the possibility for producing around 500 MW of power from all UAE SRUs, and potentially more if technologies which maximize energy efficiency are employed.

CONCLUSIONS

Sulphur recovery facilities provide significant energy benefits and should be leveraged to their fullest potential via astute design and optimized operation, deliberately focused on energy conservation. This is especially important in the current climate of low oil price and reduced margins. Tail gas treating technologies can significantly erode energy benefits provided by Claus SRUs and therefore should be designed and operated to achieve lowest acceptable SRE (highest acceptable stack gas SO_2 content). Given the huge energy requirements of amine-based TGTUs, there may be a case to lobby for relaxed SO_2 emissions regulations in the future, but that is the topic of another paper.[5]

NOMENCLATURE

°C	degrees Celsius	**Mg**	Milligram
°F	degrees Fahrenheit	**MMTPA**	million metric tons per annum
AGE	acid gas enrichment	**mol%**	mole percent
AGRU	acid gas removal unit	**MP**	medium pressure
barg	bar gauge	**MTPD**	metric tons per day
BFW	boiler feed water	**MW**	megawatt
BTEX	benzene, toluene, ethylbenzene, xylene	**Nm^3**	normal cubic metres
BTU	British thermal unit	**O_2**	oxygen
CO	carbon monoxide	**PFD**	process flow diagram
CW	cooling water	**RF**	reaction furnace
DCS	distributed control system	**RGG**	reducing gas generator
h	hour	**SO_2**	sulphur dioxide
H&MB	heat and material balance	**SRE**	sulphur recovery efficiency
H_2O	water	**SRU**	sulphur recovery unit
H_2S	hydrogen sulphide	**S_x**	elemental sulphur
HP	high pressure	**TGTU**	tail gas treating unit
kmol	kilomole	**TiO_2**	titanium dioxide
KPI	key performance indicator	**TRS**	total reduced sulphur
kW	kilowatt	**UAE**	United Arab Emirates
kWh	kilowatt-hour	**vol%**	volume percent
LHV	lower heating value	**WHB**	waste heat boiler
LLP	low-low pressure	**WHE**	waste heat exchanger
LP	low pressure	**Δ**	change
m^3	cubic metres	**ΔH**	enthalpy of reaction
MDEA	methyl diethanolamine		

REFERENCES

1. U.S. Energy Information Administration, "Spot Prices for Crude Oil and Petroleum Products," September 2016.

2. International Monetary Fund, "Primary Commodity Prices, Spot Crude Petroleum, Dubai Fateh," September 2016.

3. Gas Processors Suppliers Association, "Sulfur Recovery – GPSA Engineering Data Book," 11th ed., 2000.

4. Slavens, A. and Aikins, T., "FLEXSORB®-SE Solvent – A Guide for Deployment in the Right Applications," SOGAT 2012, March 2012, Abu Dhabi, UAE.

5. Slavens, A., Dreitzler, L., Khan, S. and Kuppuswamy, T., "Dwindling Sulphur Emissions, at What Cost?," Sulphur Conference 2015, November 2015, Toronto, Canada.