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“Techno-economic assessment of the  
needs for improved technology for  
sour oil and gas management.”

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# INTRODUCTION

For Eni and other major operators, the development of major new hydrocarbon resources in the Middle East and the Caspian region represents a key element in their business strategy for the next few decades. A recurring characteristic of the oil and gas resources in this region is the presence of high concentrations of hydrogen sulphide, H<sub>2</sub>S, which can reach as much as 20 mol % or higher in the (associated) gas.

High concentrations of H<sub>2</sub>S and high total volumes to be processed can add greatly to the cost of these assets, as well as increasing the need for strict safety and environmental practices. Not surprisingly, much effort is underway in Eni and elsewhere, to develop new processes that can reduce the economic impact of H<sub>2</sub>S and sulphur management.

The objective of this study was to develop and apply a methodology for estimating the need for and the potential impact of new technology on the development of highly sour oil and gas resources.

This objective has been reached by first developing an economic simulation tool which guides the identification of the preferred H<sub>2</sub>S processing scheme (Claus or Lo-Cat) for a given resource and by estimating the cost of this process. By considering the absolute value of the H<sub>2</sub>S processing costs and by comparing these costs to the value of the hydrocarbon resources to be produced, it is shown that for certain field characteristics (not only those of the Caspian region, but also the on-shore field of Miglianico), H<sub>2</sub>S processing costs are indeed significant. This provides a very strong incentive to identify major improvements over the consolidated technologies. Several alternative processes (some still at the research stage) which aim to cut the cost of H<sub>2</sub>S processing or extract energy or chemical value from this by-product are considered in the light of the economic needs identified for the highly sour field developments.

# 1.OVERVIEW

The first part of the paper describes the technical and economical aspects of the traditional technologies, Amine/Claus/SCOT and Lo-Cat, for H<sub>2</sub>S processing. Next, a simulation tool for estimating the costs of H<sub>2</sub>S processing has been developed from an SRI report [1] by disaggregating the costs. The model implemented allows for the calculation of both investment and operating costs pertaining to the use of these technologies.

Application of this tool to assets in Italy and the Caspian region has allowed an evaluation of the economical weight of H<sub>2</sub>S management on the asset value to be made, assuming the use of the best consolidated technologies for H<sub>2</sub>S management.

Finally, innovative and emerging technologies for H<sub>2</sub>S processing are examined in the light of the demonstrated need for alternative, lower-cost routes for H<sub>2</sub>S processing in highly sour oil and gas resources.

## 2. CONSOLIDATED TECHNOLOGIES FOR SOUR GAS PROCESSING

### 2.1 TECHNICAL ASPECTS OF GAS REMOVAL AND SULFUR RECOVERY FROM NATURAL GAS.

Natural gas desulphurisation and sulphur (S) recovery processes consist of three segments:

- Acid gas absorption
- S recovery
- Tailgas treatment

The first step in the process chain consists in the removal of acid gases from process streams by absorption. Commonly used solvents are of three types: physical, chemical, and blends. After absorption, release of the acid gases from the solvent may be achieved through pressure reduction, thermal stripping, or a combination of both. The regenerated solvent is recycled to the absorber, while the released acid gases flow forward to the S recovery unit.

Usually the  $H_2S$  is converted to sulphur through the Claus process. The chemical principle of the process is to stoichiometrically burn one-third of  $H_2S$  with air to  $SO_2$  in a thermal reaction furnace. The converted  $SO_2$  reacts with the remaining two-thirds of  $H_2S$  to form elemental S. The S conversion reactions begin initially in the thermal furnace and then continue in two or three catalytic stages over an aluminum catalyst.

For low  $H_2S$  contents, as well as low total S recovery capacities, the conventional thermal stage is not appropriate. As such, for trace amounts to 2.0 mol%, and total S recovery capacity not exceeding 20 t/d, the process employed is LO-CAT. Unlike the Claus process, which involves thermal and catalytic reactions, LO-CAT is a direct  $H_2S$  oxidation in an iron chelate solution.

For cases in which the  $H_2S$  content of the reactor off-gas is higher than that permitted by regulations, there arises the need for a further reduction of the  $H_2S$  content with a tailgas treating process. As a result, the increasing need to reduce the  $H_2S$  content has led to the development of many new technologies, including the following: Clintox, SCOT, Hydrosulfreen, Beaven.

For our study, in which we take into consideration natural gas and oil reservoirs, we have chosen the desulfurization processes that best known and commercially well established for the removal of sulphur from natural gas.

Natural Gas Desulfurization by the Process Chain of Sulfinol absorption, Claus Sulfur Recovery, and SCOT Tailgas Treatment

The Claus process converts concentrated H<sub>2</sub>S derived from solvent absorption to elemental S by a combination of thermal and catalytic oxidation stages. In the absorption process, the solvent used is sulphinol, which consists of three components: water, sulfolane, which is a physical solvent, and a tertiary amine, which is a chemical solvent (di-isopropanolamine or methyldiethanolamine). Typically, a Claus unit is separated into a thermal reaction followed by a catalytic reaction. In other words, a reaction furnace is followed by a thermal reactor which in turn is followed by three catalytic beds. The catalytic reaction involved in the Claus process requires two moles of H<sub>2</sub>S to react with one mole of SO<sub>2</sub>. As such, only one-third of the H<sub>2</sub>S stream is burned in the reaction furnace, leaving two-thirds to react with the SO<sub>2</sub> in the thermal reactor.

The reactions that characterize the process are the following:



In treating the tailgas, the treatment process used is the SCOT tailgas treating system. The entrained S in the tailgas is converted to H<sub>2</sub>S by catalytic hydrogenation over a Co-Mo or Ni-Mo catalyst. The other gases, SO<sub>2</sub>, COS, and CS<sub>2</sub> are also hydrolyzed to H<sub>2</sub>S. Following moisture reduction, H<sub>2</sub>S in the converter gas is absorbed in an MDEA solution and then recycled to the upstream Claus reactors for S recovery.

## 2.2 NATURAL GAS DESULFURIZATION

This technology is used for desulphuring high-pressure natural gas containing low concentrations of H<sub>2</sub>S and moderate amounts of CO<sub>2</sub>. Initially, H<sub>2</sub>S and CO<sub>2</sub> in a natural gas stream are absorbed from the gas by an aqueous solution of diethanolamine (DEA), producing a marketable natural gas that is acid gas-free

and that meets pipeline specifications. Subsequently, the acid gases are thermally stripped from the DEA solution. The stripped lean solution is recycled to the absorber, while the acid gases are treated with LO-CAT II, a liquid phase, direct oxidation process. This procedure separates CO<sub>2</sub> and other inert gases from H<sub>2</sub>S, the latter is then subjected to a catalytical oxidation that involves using a mixture of iron chelates as the redox vehicle that convert it to elemental S.

### 2.3 OVERVIEW OF THE TECHNOLOGICAL SEQUENCE FOR THE DESULFURIZATION PROCESS

Regarding the Claus process, the data taken as reference are:

Sour gas type	Natural gas
Sour gas feed rate, million Nm <sup>3</sup> /d	1,5
Sour gas H <sub>2</sub> S content, mol%	22,5
S capacity base cases, t/d	431

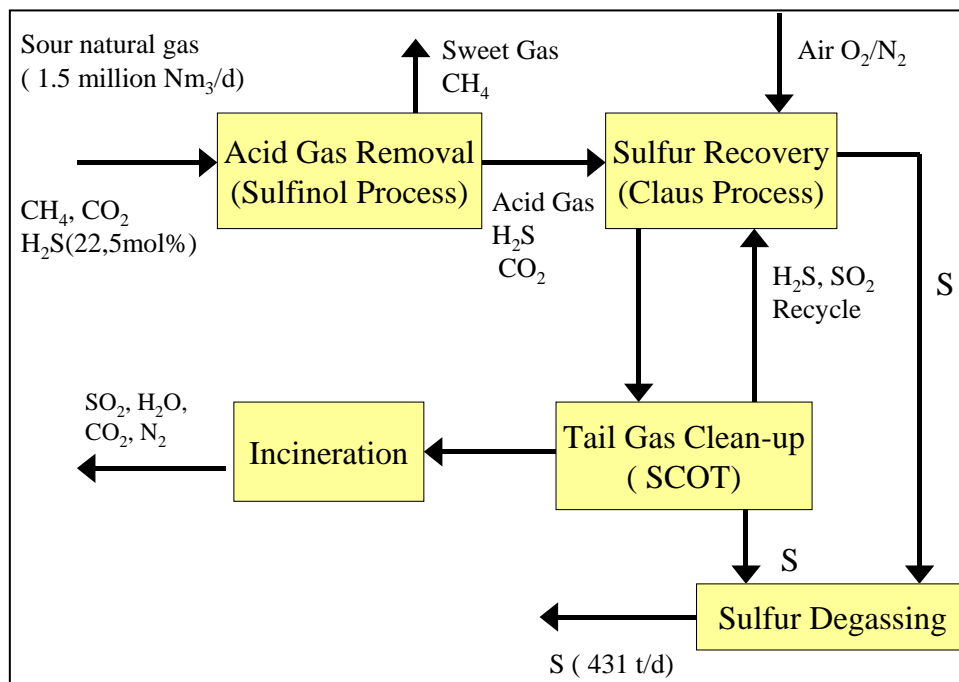
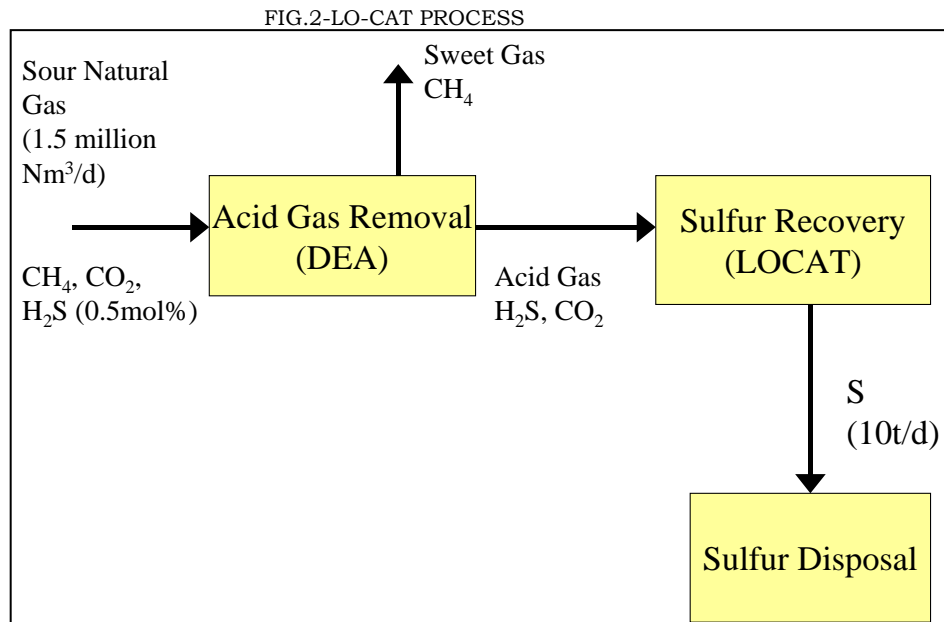


Fig. 1- Claus process

As for the LO-CAT, the data include:

Sour gas type	Natural gas
Sour gas feed rate, million Nm <sup>3</sup> /d	1,5
Sour gas H <sub>2</sub> S content, mol%	0,5



## 3. COST ANALYSIS

### 3.1 COST STRUCTURE

This section analyzes capital and production costs for desulphuring acid gas streams using either the Lo-Cat<sup>®</sup> II or Claus process. The analysis is based on data published from SRI Consulting [1]. The first step of the analysis is the definition of the plant capacity to be considered (for a description of the choice between Lo-Cat<sup>®</sup> or Claus process, see the technical section). For the data shown in this section, the plant is assumed to be located on the U.S. gulf coast. In the table below, three cases are presented for each of the two technologies; in the case of using Lo-Cat<sup>®</sup> II desulfurization technology, the base capacity of the plant considered is 8 million lb/yr (3500 t/yr) of elemental Sulfur, resulting from a gas flow rate of 53 million scfd with a H<sub>2</sub>S content of 0.5 vol%, while the half of base case has a capacity of 4 million lb/yr, and the twice of base case has a capacity of 15 million lb/yr. These examples will show how the amount of flow entering the process affects desulfurization costs.

The first two rows reported in the table indicate the estimated battery limits investment (BLI) and the estimated total fixed capital (TFC), respectively. These data are fundamental for the determination of the other desulfurization costs. The evaluation of these two data is as follow: the estimated FOB equipment cost for the base case of Lo-Cat<sup>®</sup> II technology is about \$ 0.93 million; adding the costs for direct and indirect installations, plus 10% for unscheduled equipment (e.g., mobile equipment, test instruments, portable tools), the estimated battery limits investment amounts to about \$ 5.3 million, including 25% contingency.

The off-site investment include water treatment systems, a steam boiler, and a general services facilities equal to 20% of the sum of the BLI and the utilities and storage investment, plus another 2.6% for wastewater treatment. The resulting total off-site investment is about \$ 3 million, including a 25% contingency.

Adding the battery limits and off-sites investment together, the estimated total fixed capital (TFC) investment is about \$ 8.3 million for the base case. The investment for the half of base and twice of base plants are extrapolated from the costs of the base capacity plant. The total fixed capital extrapolated from the base

case of the Lo-Cat<sup>®</sup> II technology (8 million lb/yr of sulfur) are respectively \$ 6.2 million and \$ 12 million.

The scaling exponents to be used for determining the TFC are strongly influenced by the plant capacity; for the Lo-Cat<sup>®</sup> II plant they are 0.52 and 0.42 for the twice of base and the half of base plant, respectively.

The table proceeds with the row showing the variable costs for materials and utilities. It is important to underline that for the Lo-Cat<sup>®</sup> II plant, more than 98% of the material cost is attributable to chemical costs in the direct oxidation section; on the other hand, the Lo-Cat<sup>®</sup> II section accounts for only 10% of the utilities costs, whereas the DEA adsorption-stripping section accounts for 90%.

Following with the cost analysis, the next component is the labor cost, which is mainly constituted from the sum of three factors: operating labor, maintenance labor, and control lab labor. The first of them needs to be treated with particular caution, because it represents the cost of the labor force and is therefore a variable strongly dependent upon the geographic zone where we are operating. In the example we show in the table, the plant is located on the U.S. part of the Gulf of Mexico, where the labor cost is assumed to be \$ 29.85 per hour (1997). Obviously, a plant located in another area of the world would mean a different impact of this cost on the overall expenses.

“Maintenance Labour” is calculated as a yearly percentage of BLI, and the explanation seems to be obvious, in fact as the equipment settled increases, the maintenance labour that should be done on it, increases proportionally. Generally this percentage is within the range of 1% ÷ 4% of BLI. “Control Labour” is referred to the control activities executed in laboratory during the process and they are determined as a percentage of the operating labour (it is usually about 20%).

Adding up these three terms together we obtain a value of 5.77 cent/Lb for the base case.

Adding to the Labour cost the maintenance materials expenses, which are a function of BLI, and the operating supply costs, which depends on the manpower in the field , we arrive at the “Total Direct Costs”, which in our case amounts to 20.95 cent/Lb. To the operating expenses considered so far we have to add general and administrative ones, grouped under the name “Plant Overhead”; these costs are generally linked to employee costs and therefore they are

expressed as a percentage of “Labour Cost”. The range of variation that this percentage may undertake is very wide (50% ÷ 150%) and depends on the specificity of the project we decide to examine. Normally, if precise sheets) of expenses are not available, we can consider a mean value of 80%.

The following cost, “Taxes and Insurance”, should not be confused with taxes on revenues eventually reached, but it is above all determined by insurance expenses on the equipment settled and it is expressed as a percentage of TFC (2%).

Summing up the terms listed so far we are now at the point to be able to define the Plant Cash Costs, which amount to 27.73 cent/Lb (base-case).

Considering for the plant an economic life of 10 years and a depreciation plan with fixed quota, we now have to take into account the term “depreciation”, equal to 10% of TFC, that should be added up to “Plant Cash Cost”: in this way “Plant Gate Costs” are obtained (38.58 cent/Lb). Obviously this value is strictly linked to the depreciation plan chosen.

Generally the costs of G&A, sales and research are considered out of the gate and they assume a forfeit value of 1% of the final cost in the year considered.

In determining the final cost linked to the desulphurisation process it is usual to consider a recovering level of invested capital, defined as “ROI before taxes”, and expressed as a percentage of TFC. Nevertheless this is only one of the possible approaches in defining a recovery of the invested capital in the determination of the final cost, and perhaps it is the most usual methodology in the Anglo-Saxon system. In Europe the way usually followed is more cautious than the previous one; it suggests determining the recovering quota as in the following formula:

$$R = \frac{C(1+i)^{e.l.} * i}{(1+i)^{e.l.} - 1}$$

where:

**e.l.** is the “economic life” of the plant (in our example it is 10 years);

**C** is the initial investment;

**i** represents the expected rate of return, which is determined by three factors:

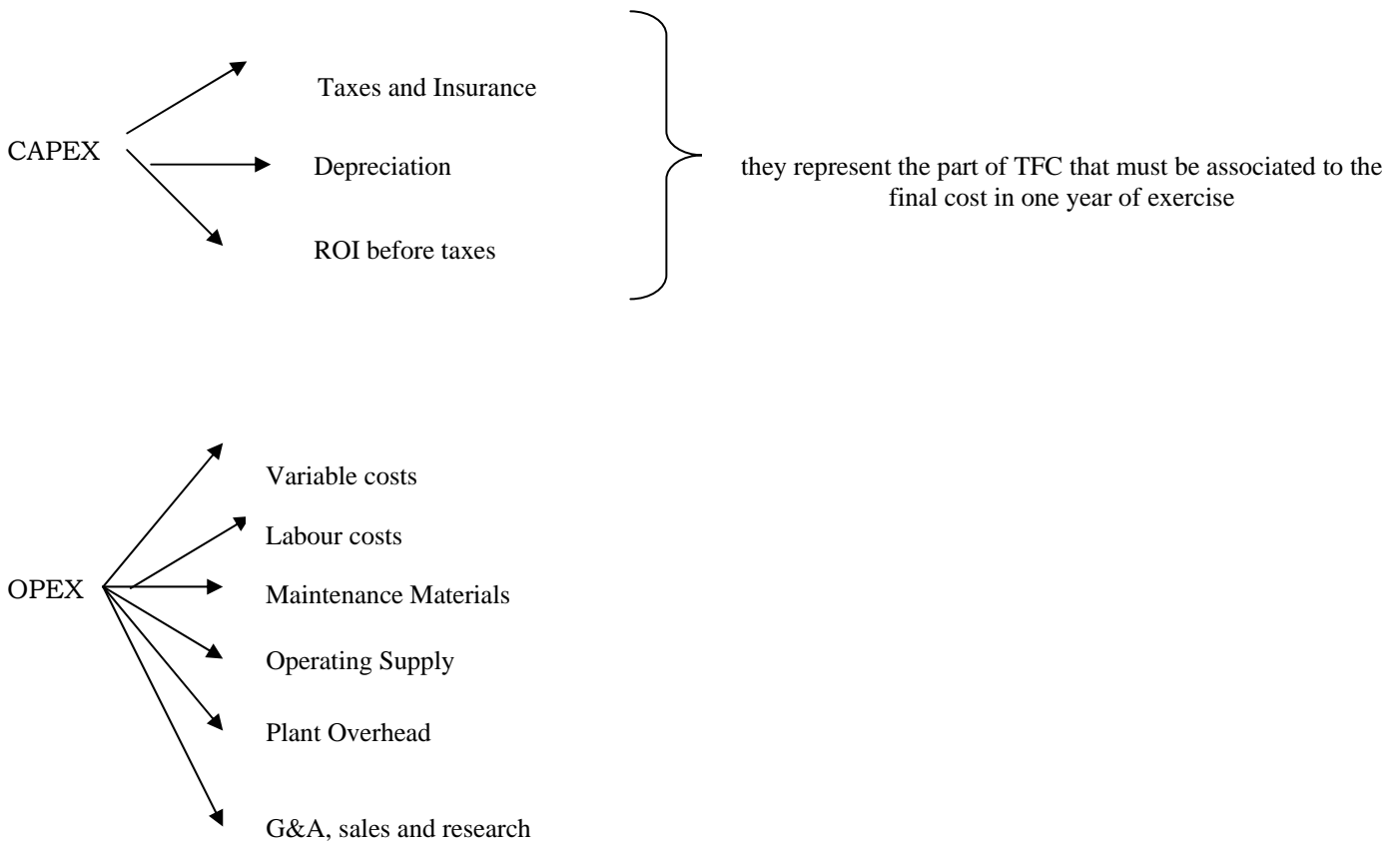
- the risk-free rate of interest;
- an adequate premium for the risk managed in executing such a project;

- a margin sufficient to assure the recovery of the payment of taxes.

In general, with this last methodology we obtain lower capital recovery, and consequently a lower “final cost”, with respect to the one in the table below, that amounts to 66.37 cent/Lb.

We have preferred to reclassify the costs examined above in terms of “Capex” and “Opex” costs as the scheme below:

reclassify the costs examined above in terms of “Capex” and “Opex” costs as the scheme below:



Tab.1-Production costs. From [1].

<b>production costs [c/lb]</b>				
<i>Capacity [Mlb/y]</i>				<b>7,65</b>
<i>BLI [M\$]</i>				<b>5,30</b>
<i>TFC[M\$]</i>				<b>8,30</b>
			<b>un. capex</b>	<b>un.opex</b>
<b><i>Variable Costs</i></b>				<b>13,73</b>
	Operating Labor			<b>3,42</b>
	Maintanance Labor (%/Y of BLI)	2,40		<b>1,66</b>
	Control Labor (% of Op Labor)	20		<b>0,68</b>
<b><i>Labor Costs</i></b>				<b>5,77</b>
	MaintananceMaterials (%/Y of BLI)	1,6		<b>1,11</b>
	Operatin Supply (% of Open Labor)	10		<b>0,34</b>
<b><i>Total Direct Costs</i></b>				<b>20,95</b>
	Plant OverHead (% of Lab Costs)	80		<b>4,61</b>
	Taxes_Insurance (%/Y of TFC)	2		<b><u>2,17</u></b>
<b><i>Plant Cash Costs</i></b>				<b>27,73</b>
	Depreciation (%/Y of TFC)	10	<b><u>10,85</u></b>	
<b><i>Plant Gate Costs</i></b>				<b>38,58</b>
	G&A, Sales Research			<b>0,67</b>
<b><i>Net Production Costs</i></b>				<b>39,25</b>
	ROI before Taxes (%/Y of TFC)	25	<b><u>27,12</u></b>	
<b><i>Product Value</i></b>				<b>66,37</b>
<i>unitary capex [c/lb]</i>			<b><u>40,14</u></b>	
<i>unitary opex [c/lb]</i>				<b>26,23</b>
<b><i>Product Value</i></b>				<b>66,37</b>

### 3.2 STRUCTURE OF COSTS- A MORE ANALYTIC APPROACH

Based on the SRI data [1], we have developed an analytical “model” to estimate all the costs for both technologies.

Indicating with  $q$  [tS/d] the daily S production, the desulphurisation CAPEX (investments) depends on  $q$  as:

$$capex(q) = K_c \cdot q^\alpha$$

The coefficient  $K_c$  depends on the country and the change with US \$; the  $\alpha$  depends only on technology chosen.

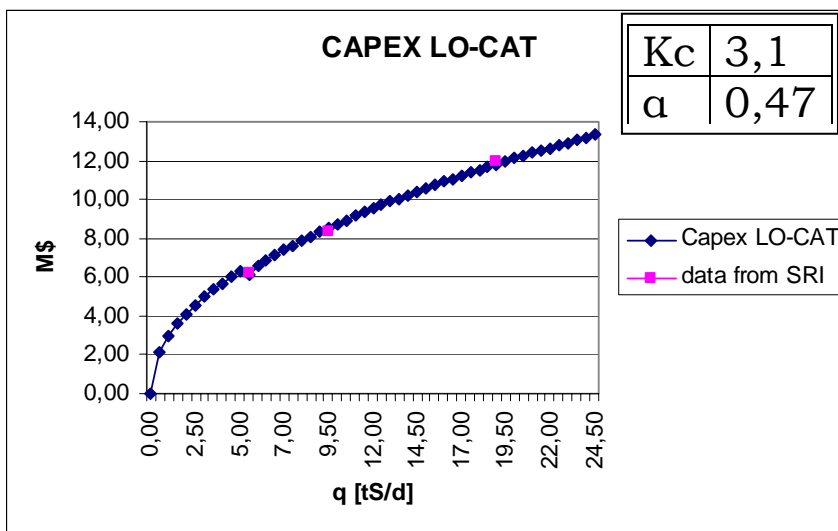
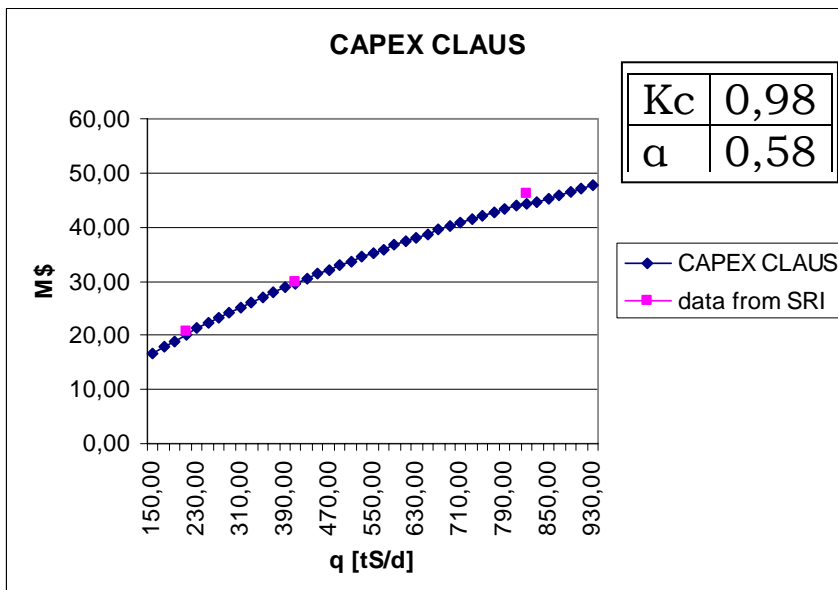


Fig 3.(a) Capex of Lo-Cat. (b) Capex of Claus



The production cost is divided in a unitary capex and a unitary opex (see previous chapter). We indicate the unitary capex as  $cx$ ; this is an annual percentage of capex total; it depends on following economic data:

- $r$  = return on Investments, ROI
- $d$ =depreciation
- $t$ =taxes

Therefore to have the unitary capex ( $cx$ ) expressed in [\$/Kg]:

$$cx = \frac{K_c}{365} \cdot \frac{capex(q) \cdot (r + t + d)}{q}$$

with

$$\chi = \frac{K_c (r + t + d)}{365 \cdot 1000}$$

$$cx[\$/Kg] = \chi q^{\alpha-1}$$

The opex cost expressed in [\$/y] show a  $q$  dependence as:

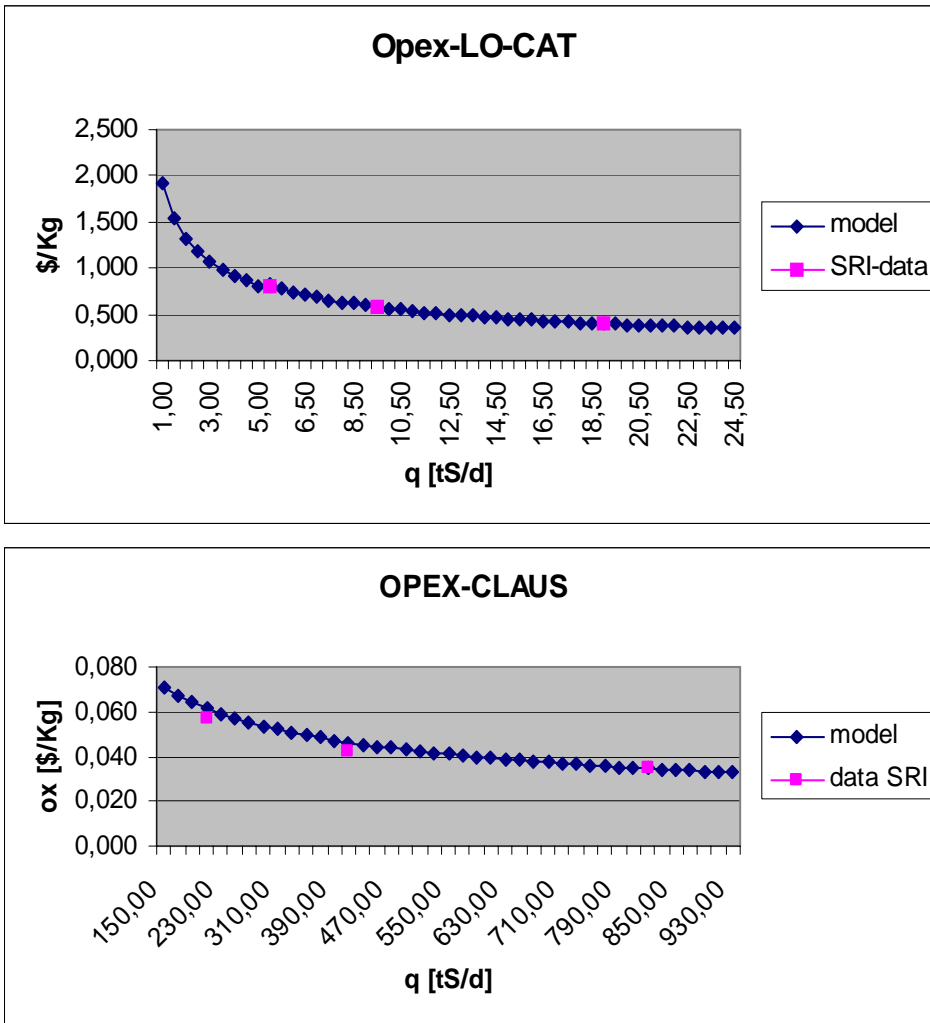
$$Opex(q) = K_0 q^\alpha$$

The unitary opex ( $ox$ ) expressed in [\$/Kg] are a similar dependence on  $q$ :

$$ox(q) = \frac{K_0 q^\alpha}{q * 365 * 1000} = \mathcal{G} q^{\alpha-1}$$

The important difference between the two classes of cost is that in the unitary capex there are the specifications of “economics”.

Fig 4.(a) Unitary Opex of Lo-Cat. (b) Opex of Claus



The unitary cost  $z$  [\$/Kg] will be given by the sum of unitary capex and opex and shows a dependence with  $q^{\alpha-1}$ . Because of  $\alpha$  is less than 1 for both the technologies, the impact of desulphuring is particularly onerous for small scale operations.

A summary of data from SRI report is showed in following tables. In Table 2.a there are the parameters for both the technologies. In Table 2.b and 2.c there are the costs calculated for 3 cases for each technology.

---

1 1Nm<sup>3</sup>=35,31 scf

Table 2.a

	Lo-CAT	CLAUS
q [tS/d] range	0-25	150-930
$\alpha$	0,47	0,58
$K_C$	3,91	0,98
$K_O$	1,91	0,58

Table 2.b

LO-CAT technology	1	2	3
q [tS/d]	4,75	9,49	19
[H <sub>2</sub> ]	0.5%	0.5%	0.5%
$V_{in}^2$ [Nm <sup>3</sup> ]	0,75 10 <sup>6</sup>	1,5 10 <sup>6</sup>	3 10 <sup>6</sup>
z [\$/1000scf]	0,42	0,29	0,22

Table 2.c

CLAUS technology	1	2	3
q [tS/d]	205	409	817
[H <sub>2</sub> ]	22.5%	22.5%	22.5%
$V_{in}^1$ [Nm <sup>3</sup> ]	0,7 10 <sup>6</sup>	1,42 10 <sup>6</sup>	2,83 10 <sup>6</sup>
z [\$/1000scf]	1,32	1,00	0,8

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<sup>2</sup> 1Nm<sup>3</sup>=35,31 scf

### 3.1 THE PROGRAM EXCEL “H2S COSTS”

#### 3.3.1 DETERMINING OPTIMAL THE SIZE OF A THE PLANT

The Excel program “H2S\_cost” has been developed to provide an estimate of the desulphurisation cost ( $z$ ) for a gas or gas/oil field. We suppose the technology choice is only between the LO-CAT process or CLAUS technology. Other technical alternatives will be discussed in the next chapter.

To estimate  $z$  it necessary to know:

- physical data from field
- economics
- technology parameters

The physical data from field are:

- The production rate PR [bbl/d]
- Gas to Oil Ratio (GOR) [m<sup>3</sup>/m<sup>3</sup>]
- H<sub>2</sub>S molar concentration [H<sub>2</sub>S]

From them is possible to work out the volume feed of natural gas going in the plant, that is indicated as  $V_{in}$  [Nm<sup>3</sup>/d].

In a oil field it can express as:

$$V_{in} = PR \cdot GOR \cdot 0,159 [m^3 / bbl]$$

For a gas field  $V_{in}$  is directly the production rate of natural gas expressed as [Nm<sup>3</sup>/d] or as [boe/d]. From  $V_{in}$  and the molar concentration of H<sub>2</sub>S in the gas (**[H<sub>2</sub>S]**) the daily S production ( **$q$  [t/d]**) will be given as:

$$q = \beta \cdot V_{in} \cdot [H_2S]$$

The calculus of coefficient  $\beta$  (~1,2 10<sup>-5</sup>) is described in Appendix B.

The value of  **$q$**  allow to chose the technology and, therefore, the parameters:

- K<sub>c</sub>
- K<sub>o</sub>
- $\alpha$

Only the coefficient  $\alpha$  depends exclusively on technology; the other parameters refer to the costs in USA \$ of 1995. Therefore it is necessary to update the value

of the coefficient with a scale factor (**sf**) that considers the change, the inflation and the country of installed plant.

In addition to the scale factor, there are other economics parameters whose mean has been explained in previous chapter. Definitely the input parameters are:

- (r)eturn on ROI
- (d)epreciation
- (t)axes of plant
- (s)cale (f)actor

that with the technology choice is necessary to estimate the desulphurisation cost (expressed in different measure units).

If *q* exceeds the maximum capacity of a plant, more plant have to be built and this number is indicated as *n*. Definitely the final formula to calculate the total unitary cost *z* is :

$$z[\$/Kg] = fs \cdot n \cdot \left[ \frac{K_c \cdot (r + t + d)}{365 \cdot 1000} + \frac{K_0 \cdot 365}{1000} \right] \cdot \left( \frac{q}{n} \right)^{\alpha - 1}$$

A Excel worksheet has been implemented to work out the unitary cost (expressed in different measure units) from the economical and physical data. The form of this software is shown in Fig.5

Physical Field Data			Economics Data				LO-CAT	Technology	CLAUS	
PR [boe/c]	Vin [Nm3]	[H2S]	(r)eturnROI	(d)epreciation	(t)axes	fs	(n) plants	Kc	Ko	alfa
9,0E+03	1,00E+06	1,0	15%	10%	2%	1	1	1,91	0,47	0,58
	<i>q</i> [tS/d]			capex [M\$]		opex [M\$/y]				
	12,7			8,3		0,75				
single plant	12,7			8,3		0,75				
				unitary capex[\$/Kg]		0,4859				
				unitary opex [\$/Kg]		0,1616				
				des. cost z [\$ /Kg]		0,65				
				des. cost z [\$ /t]		647,48				
				des. cost z [\$ /1000Nm3]		8,22				
				des. cost z [\$ /1000 scf]		0,23				
				des. cost z [\$ /boe]		0,92				

FIG 5- A VIEW OF EXCELL WORKSHEET TO WORK OUT THE DESULPHURING COST

### 3.3.2 THE ANALYSIS OF DESULPHURING COST

Another approach is studying the desulphuring cost in function of physical variables. For instance, it is possible, fixed the volume of gas, investigate the dependence of  $z$  on percentage of  $H_2S$ . In this example fixed  $V_{in}=1 \text{ Mnm}^3$  the cost per  $1000 \text{ Nm}^3$  is shown if fig.6:

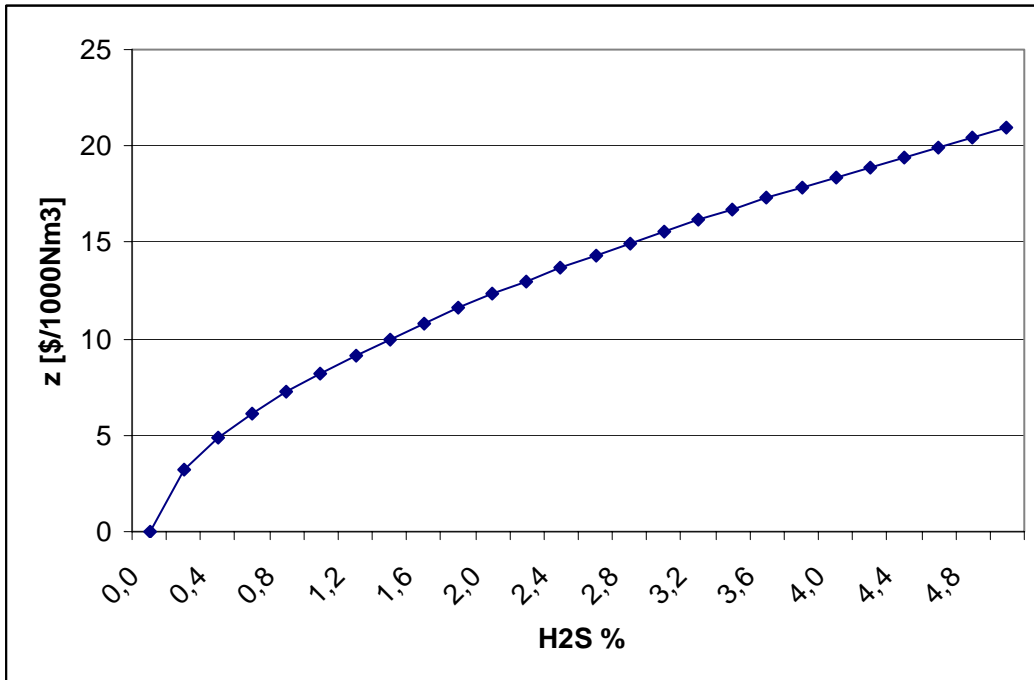


Fig.6 Cost  $z$  versus  $H_2S$  concentration

### 3.3.3 AN INVESTMENT ANALYSIS

Let  $p_z$  be the value of  $S$  in the market expressed in [\$/Kg]. A simple investment analysis show that the net present value of the discounted cash-flows has the form of:

$$NPV = -capex(q) + \sum_{t=1}^T \frac{(p_{z_t} - z_t) \cdot q_t}{(1+i)^t}$$

In this case the plant is considered simply as a plant that produced a product ( $S$ ) that can be sold in the sulphur market. If the NPV value would be positive, the desulphurisation process will be a self-sustaining investment. Roughly speaking we can compare the average cost of the desulphurisation process [0.1 \$/Kg] with a typical recent value for  $S$ , ie., less than 30 \$/t. It is clear that this is not a self-

sustaining investment. The problems related to sulphur market will be treated in section 3.

Where the sulphur finds no outlet on the market (a situation found in several areas of the world), the cost of desulphurisation will be charged entirely to the gas value, or to the oil value if there is no market for the associated gas. Supposing where  $capex_{field}$ ,  $c$  (the cost of extracted gas) are related to the gas field.

In this case  $z$  is expressed in  $[\$/1000 Nm^3]$ .

$$NPV = -capex_{field} - capex(q) + \sum_{t=1}^T \frac{[p_t - (c + z(q)) \cdot V_{in}]}{(1+i)^t}$$

If the cost is charged on all the hydrocarbon production, we have:

$$NPV = -capex_{field} - capex(q) + \sum_{t=1}^T \frac{[p_t - (c + z(q))] \cdot Q_{boet}}{(1+i)^t}$$

where  $z$  is expressed in  $[\$/boe]$ .

For this analysis it is necessary to know the expectations on prices, the costs, the capex and the profile of production that is not easy to have.

Therefore a sensitivity approach has been used to estimate the convenience of developing field. A worksheet “NPV-sensitivity” allows this kind of procedure (fig.7)

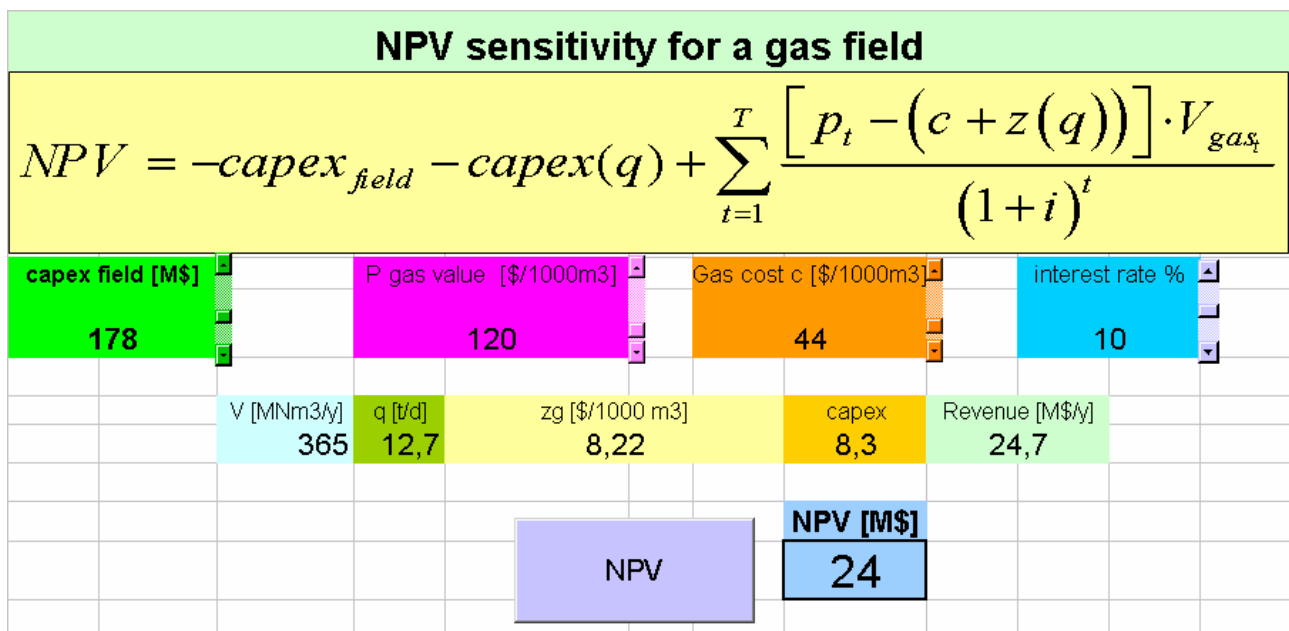


FIG.7- A WORKSHEET THAT ALLOWS THE SENSITIVITY ANALYSIS ON OIL/GAS FIELD

### 3.4 APPLICATION TO HYPOTHETICAL FIELDS

We start applying the model before described to some field with different characteristics.

#### 3.4.1 SMALL ON-SHORE ITALIAN FIELD

The data of this field give an annual sulphur production of  $q=7,6$  tS/d. The peak of production will be in 2012: 8000 bbl/d of oil and 0,15 MNm<sup>3</sup> of gas. This corresponds to a total hydrocarbon production<sup>3</sup> of 8960 boe/d.

This suggests the choice of the Lo-Cat technology and, therefore, allows using the procedure to work out the capex and the opex of the desulphurisation plant. The value of return on ROI used is 15% and the factor of scale (fs) has been kept at the same value of  $sf=1$ .

The value of gas in this area is estimated in 2,5 [\$/1000 scf] ( 88,31 \$/1000Nm<sup>3</sup>) [2]. This is a conservative estimation of gas value that could be worth more but it is enough to provide a qualitative estimation of the impact of desulphurisation on total costs.

Table 3 summarizes this results:

Table 3

Field Data	
PR [boe/d]	8960
V <sub>in</sub> [Mnm <sup>3</sup> d]	0,15
[H <sub>2</sub> S]	4%
q [tS/d]	7,6
Capex field [M\$]	164
Opex field [M\$/y]	14
Desulphuring costs	
Capex [M\$]	6,2
Opex [M\$/y]	0,56

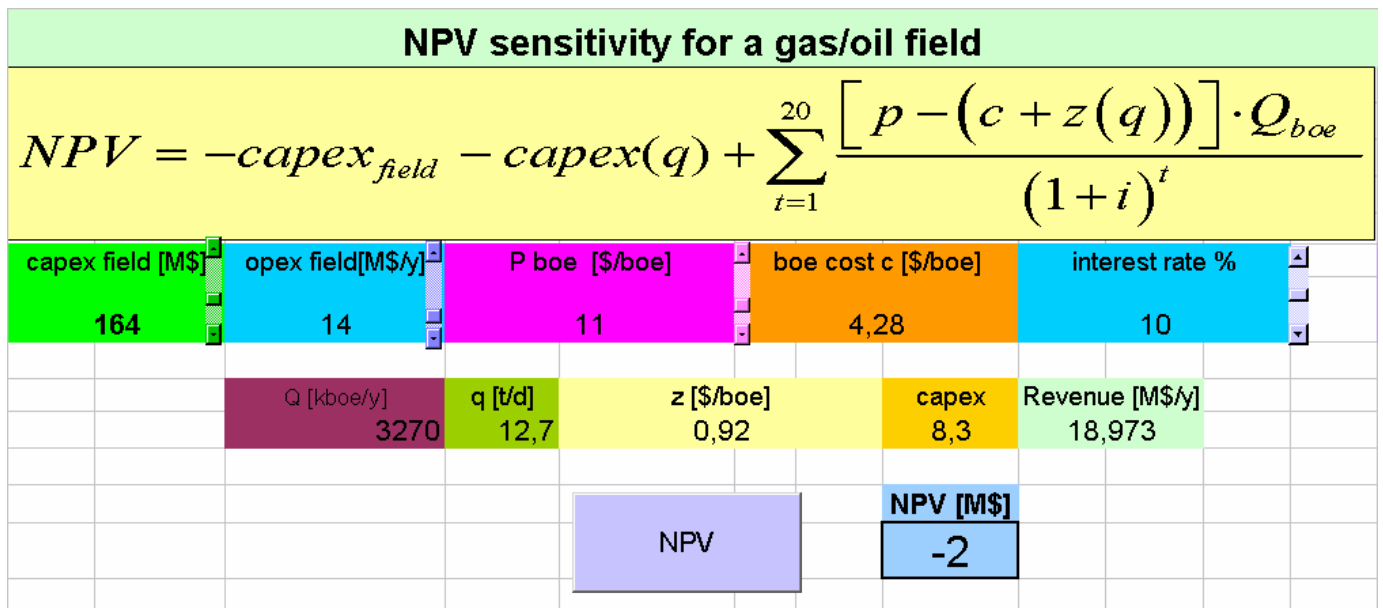
<sup>3</sup> For Italian gas 1 Nm<sup>3</sup>=6,3x10<sup>-3</sup> boe

Gas Value	88
[\$/1000Nm3]	
z [\$/Kg]	0,8
z [\$/1000Nm3]	40,86
z [\$/boe]	0,68

Our analysis show the high cost of desulphurisation on total cost of production (5 \$/boe) may reach more than 10%.

For this field the capex and opex are supposed to know and therefore is possible to make a sensitivity analysis on boe price that make non economic the development of field. Using the Excel worksheet “NPV-sensitivity” is possible estimating the price in 10-11 \$/boe. In fig.8 the application of Excel worksheet is shown.

Fig.8 Sensitivity Analysis for Miglianico field



### 3.4.2 CASPIAN REGION FIELDS

Evaluating H2S processing costs is of particular interest for the Caspian region, given the characteristically high H2S levels, and the strategic value of the this region and middle east for Eni and other major oil companies. We begin by analyzing the field of Karachaganak; a description of which is found in [8].

The costs of Kazakhstan have been calculated both with a factor scale  $fc=1$ , although it does not seem plausible because of in this remote region all the cost are higher than those in USA.

The value of  $r$  used is 15%. The field data used [9] refer to a production of 200000 boe/d, a production level for the years until 2004. In fact the production rate for this field will have the peak of 500000 boe/d in 2010.

The daily production of 200000 boe/d is split into:

- 120000 bop/d of liquids
- 12,8 MSm<sup>3</sup>/d (equivalent to 80000 boe/d)

Only one-half of the gas extracted is sold to the market, the rest is re-injected into the reservoir. With this volume of gas and [H<sub>2</sub>S]=3,5% we can calculate the daily S output of 284 t/d which suggests using the Claus technology.

The value of gas in this area has been estimated for the sake of this exercise to be in 0,5 \$/1000 scf [2]. The cost of desulphurisation is charged both on oil and gas sold (that is the half of the total gas extracted) to finally have the  $z$  expressed in [\$/boe]

Tab.4 summarizes the data from the Karachaganak oil-gas field

Table 4. Karachaganak

Field Data	
PR total sold[boe/d]	160000
V <sub>in</sub> gas [Mnm <sup>3</sup> d]	6,4
[CO <sub>2</sub> ]	5,5%
[H <sub>2</sub> S]	3,5%
q [tS/d]	284,5
Capex field [M\$]	----
Opex field [M\$/y]	----
Desulphuring costs	
fc	1
Capex [M\$]	26
Opex [M\$/y]	5,61

Gas Value [\$/1000Nm3]	18
z [\$/Kg]	0,12
z [\$/1000Nm3]	10,81
z [\$/boe]	0,43

In Tab.5 there are the data from Kashagan oil-gas field. In this analysis we suppose a production of 100 kbbld with a concentration of [H2S]=16%. The physical fields data give a S output of 1605 tS/d; this means using 2 Claus Plants of 803 tS/d capacity. Therefore the total capex and opex will be the sum of 2 opex and capex of a 803 tS/d plant. The parameters of cost are supposed to be the same as for Karachaganak. In this case the gas extracted is supposed not to be sold to the market. Therefore the cost of desulphurisation is attributed entirely to the price of the crude oil. In fact even if the gas were flared, it would still have to be desulphurized to respect environmental laws.

Therefore in Kashagan there is a conjunction of two difficult situations: a very high percentage of H2S and the absence of a market for the large volumes of associated gas.

Table 5 summarizes the data:

Table 5- Kashagan

Field Data	
PR [boe/d]	100000
GOR [Nm <sup>3</sup> /bbl]	79,3
V <sub>in</sub> [Mnm <sup>3</sup> d]	7,9
[H2S]	16%
q [tS/d]	1605
Capex field [M\$]	-----
Opex field [M\$/y]	-----
Desulphuring costs	
Number plants	2 x 803 [tS/d]

Capex [M\$]	56,1
Opex [M\$]	20,48
Gas Value[\$/1000Nm3]	18
z [\$/Kg]	0,07
z [\$/1000Nm3]	14,2
z [\$/boe]	1,12

### 3.5 THE SULPHUR STORAGE ISSUE

For all three cases examined, application of the consolidated technologies for H<sub>2</sub>S processing (in one case, Lo Cat, in the other cases, Claus) lead to high total costs. Because the Lo-Cat process for the removal of low quantities of S (scavengers) have relative high cost, the cost of removing S can be a high percentage on the total cost of production (small Italian field).

Application of the Claus process for the removal of large quantities of S leads to low unitary cost per Kg of sulphur removed. However, the large volumes of H<sub>2</sub>S to be handled and the limited market for the large volumes of produced gas for the two Caspian region fields studied lead to a very significant contribution of H<sub>2</sub>S processing costs to total cost of extraction. For Karachagank, this is 0.41\$/boe and for Kashagan 1,12 [\$/boe]. A rough estimation for Kashagan field when the production will be around 1 Million of bbl/d give a daily cost of more 1 M\$/d.

Another consideration for the Caspian fields regards the distance of the sulphur produced from the market, and the present world-wide over-supply of sulphur on the market. For these reasons, not only must the Caspian region sulphur be considered as a cost but it is likely that the sulphur produced will be destined for long-term storage.

In this case there are a lot of open questions.

The problem is producing a huge amount of solid S the cannot be sold: the market of S has an excess of supply and the S price is dropped to zero value [5].

Storing the S produced seems to be the only feasible solution but this has a cost. Producing 1600 t of solid S per day, as before described in our example, means reaching an annual production of 0,58 million of tons that have to be stored. The cost of storing sulphur as a solid block is estimated in 13\$/tS, for an annual cost of sulphur storage of 7,6 M\$. Other costs are related to monitoring and maintenance of the large sulphur blocks in order to avoid environmental damage. It has been reported that the annual cost for avoiding acid runoff from old and unstable sulphur blocks has reached 3 \$ t/y in some cases [5]. This would correspond, for the above annual production, to an additional 1,7 M\$ per year. It is easy to understand the importance of this issue for the H<sub>2</sub>S rich gas of Kazakhstan fields.

In table 6 are reported the costs before described and the cost per year due to different contribution in Kashagan field with a production rate of 1M boe/d:

		Cost for Kashagan [M\$]
Z [\$/boe]	1.12 <sup>4</sup>	408
Storage cost [\$/tS]	13	76
Other costs [\$/y]	3	17
<b>total</b>		<b>501</b>

With this estimation become easy to understand the importance of economical impact that the H<sub>2</sub>S removal issue will have on developing fields in this area.

The problem related to the sulphur storage in Kazakhstan' Tengiz field is well discussed in [6].

In next section some alternatives technologies to storage will be discussed.

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<sup>4</sup> prudent estimation calculating the cost with fs=1

## 4. TECHNOLOGY ALTERNATIVES

The economical impact of S storage before analyzed make necessary studying other approach to avoid the storage. The technological alternatives that will be discussed in this chapter are based on following ideas:

- avoiding Claus and Lo-Cat improving the acid gas re-injection
- using H<sub>2</sub>S as a source of value (energy, H<sub>2</sub>)

### 4.1 ACID GAS INJECTION

The acid gas from the amine regeneration column, is at low pressure, typically 5 to 7 psig. The acid gas is made largely of H<sub>2</sub>S and CO<sub>2</sub>, but it also contains small amounts of light hydrocarbons and nitrogen. Because of the nature of the stripping process, the acid gas is saturated with water at these conditions. In order to get it through the pipeline and into the injection, the acid gas has to be compressed to a sufficient pressure. The compression takes place in a multi-stage compressor with interstage cooling until 49°C. At this temperature, an aqueous phase is likely to form and is removed using interstage knock-out scrubbers. Next the acid gas is transported to an injection well in a pipeline.

Based on safety considerations, the pipeline is generally kept as short as possible. Finally, the fluid is injected usually into a suitable underground formation. The formation is selected based on geological criteria such as the size of the disposal reservoir, and the containment of the injected acid gas.

In an acid gas injection scheme, pressures can range from near atmospheric up to 30 Mpa or more; the upper limit dictated by the selected reservoir. The temperature can range from about 30°C up to as much as 200°C; again the upper limit is being the reservoir conditions.

Acid gas injection is essentially a zero-emission process because during normal operation all of the H<sub>2</sub>S from the produced gas is re-injected. This solution requires adequate technologies since there are several important physical properties and phase equilibria to be considered in the design of an acid gas injection scheme.

These include, but are not limited to:

- non aqueous phase behaviour
- water content of the acid gas and the possible formation of an aqueous liquid.
- hydrate formation
- the density and viscosity of the acid gas

Perhaps the most significant problem with the calculation for the acid gas injection is that much of the process occurs at conditions that are near the critical point of the mixture.

Typically the range of temperature is  $0.75 < Tr < 1.25$ , where  $Tr$  is the reduced temperature, the ratio of the temperature to the critical temperature. The vicinity of the critical point is a region where it is difficult to accurately calculate the physical properties. Not only do the property estimation tend to be poor, but in addition, the properties are strong functions of the temperature and the pressure. In this region, small changes in pressure or temperature can have a dramatic effect on the physical properties.

Acid gas injection in suitable formations is a technology so far only applied to low capacity reservoirs. It is not yet possible to use this method to stock the quantities involved in order to significantly tackle the issue of sulphur generation.

The main obstacles to overcome are:

- the technological gaps on compression plants.
- the availability and selection of suitable formations to stock such quantities.
- the security risks arising from the presence of tonnes of high-pressure liquid hydrogen sulphide on the plants.

In Tab. 7 and Tab. 8 we compare the data of a typical small size LO-CAT recovery plant with a Gas Injection Plant for the same data field.

Vin [Mnm3d]	0,5
[H2S]	0,44%
q [tS/y]	2,8

Tab. 7

Tab. 8

Capex [M\$]	5	Capex [M\$]	2,25
Opex [M\$/y]	1,13	Opex [M\$/y]	0,3
z [\$/Kg]	2,44	z [\$/Kg]	0,3

The technology of re-injection show a lower cost of desulphuring. The use of this technology could be competitive the gas extracted.

#### 4.2 XERGY PROCESSING.

A new H2S removal technology has been developed by Xergy processing for removing a S in a range of 50 Kg/d to 20 t/d. It is the range where the usual desulphurisation system is quite expensive [3].

Is possible compare capex and opex of this technology with LO-CAT. The data of field are showed in Tab 9.a. For this production of S (q=1,8 tS/d) we can compare the cost for Lo-CAT from [3], the cost of Lo-CAT from our model and the cost of Xergy method. The results are shown in Tab 9.b and it's clear the convenience of this new technology ( and the precision of our model whose results are similar to costs reported in [3]).

Tab 9.a

V <sub>in</sub> [Mnm <sup>3</sup> d]	0,28
[H2S]	0,5%
q [tS/y]	1,8

Tab 9.b

	LO-CAT from [3]	LO-CAT Our model	Xergy method
Capex [M\$]	2,6	2,7	2
Unitary Opex [\$/Kg]	0,37-0,72	0,36	0,062-0,085

Some discussion of other technologies that have been developed to treat high pressure sour gas are in [4]

### 4.3 HYDROGEN PRODUCTION

Because of the high cost of removing S for small-scale process, there is interest in identifying alternative S recovery approaches. One of these is the production of hydrogen ( $H_2$ ) from  $H_2S$ , for which many experimental processes have been devised. The  $H_2$  is a value in the petrolchemistry and will be more and more important in the future hydrogen economy.

The different processes are quite diverse, and many of them are very interesting. A summary of the processes follows.

In the early 1980s, the I.V. Kurchatov Institute in Moscow began experiments to convert  $H_2S$  to hydrogen using microwave radiation. Within a short time they had installed a large-scale pilot plant at a gas plant at Orenburg, Russia. Follow-up work took place in North America at Argonne National Laboratory and in a joint Canadian study (Alberta Energy/Shell Canada/Atomic Energy of Canada).

Several laboratories are working on electrolysis of  $H_2S$ , either as a gas or as liquefied  $H_2S$  or as an aqueous solution. Major projects have been conducted at Georgia Tech, CANMET, Texas A&M and elsewhere.

Some novel processes use electrical discharges. Again there are several variants. Thus "hot" electric arcs or the cooler discharges from ozonizers can be used.

Other processes that are being looked at use thermal decomposition -- basically strong heating. Alberta Sulphur Research Ltd. is piloting a thermal process in which heat from a Claus reaction furnace is used as the source of energy. The economics of thermal processes have been studied at the University of Calgary.

Ingenious and elaborate thermal cycles have been investigated. In most of these, at least one step does not give good yields, and economic separation of the products is an issue.

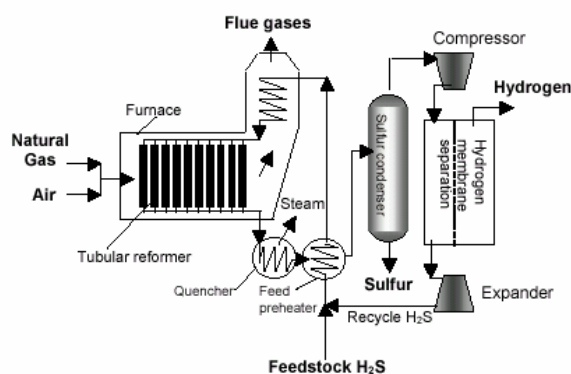
An oxidation-reduction cycle using anthraquinone derivatives is being pursued by Marathon Oil. It is quite advanced and is heading for the pilot stage.

There are processes that use light energy. Some of these use the sun's thermal energy, while others use photochemical principles. Research in the latter is focused on finding better photo catalysts and on ways to obtain sulfur in a salable form.

Considering the intensity of research in this field, a commercial H<sub>2</sub>S splitting process could emerge soon. Such a process could be practiced anywhere there is a source of H<sub>2</sub>S. However, transporting hydrogen is costly. Hydrogen should preferably be consumed at the site. Therefore the most suitable locations would be refineries or heavy oil upgraders, including the Alberta oil sands.

According to Cox (1998), using an efficient H<sub>2</sub>/H<sub>2</sub>S separation system, the thermal decomposition of H<sub>2</sub>S is able to produce hydrogen at a cost approaching that of the conventional SMR process. The analysis of Cox et al. showed that the most economic route for hydrogen production by direct decomposition of H<sub>2</sub>S is one in which CH<sub>4</sub> is burned to supply the decomposition heat and unconverted H<sub>2</sub>S is recycled until extinction. This scheme would produce H<sub>2</sub> at a cost of about **\$4.50/10<sup>6</sup> BTU** (corrected to 1998 US dollars). This figure compares favorably with **\$4.75/10<sup>6</sup> BTU** (corrected to 1998 US dollars) for a Claus plant to treat the same amount of H<sub>2</sub>S plus a conventional SMR plant to generate an equivalent amount of H<sub>2</sub> gas. In principal, this process can be integrated with a non-polluting heat source (for example, solar) to eliminate emission of greenhouse gases from the combustion furnace. Alternatively, part of the hydrogen gas produced in the process can be rerouted and burned in the furnace as fuel without any emission of greenhouse gases.

Fig.8



#### 4.4 ELECTRICITY GENERATION FROM SOUR GAS

Peter Clark, professor of Chemistry at the university of Calgary and director of research at Alberta Sulphur Research Ltd. (ASR), has recently launched an innovative idea by declaring that sulphur can be consider as a source for electricity production. He even goes so far as to say that by using sulphur energy, it would be possible to supply 25% of the electricity required in the entire Canadian province of Alberta.

Dr. Clark has recently revealed the results of his research; he shows that sulphur can be used as a source of fuel to produce enough energy in the form of water vapour to feed turbines that would produce electricity. What Dr. Clark is proposing, and what has already been proven in the laboratory, is that the reservoirs of sour gas can be used more profitably. As usual, after removal of the methane,  $H_2S$  and  $CO_2$  are brought into a pressurized chamber where they are burned. The products stemming from this combustion are sulphur dioxide ( $SO_2$ ), water vapour and the remainder of the  $CO_2$ . The energy released from this reaction in the form of water vapour is considerable, and can be used to drive turbines to produce electricity: by way of comparison, coal combustion (the primary source of electricity in Alberta) produces 8000 BTU/pound, whereas hydrogen sulphide combustion produces 6000 BTU/pound. Moreover, the resulting  $CO_2$  and  $SO_2$  will be redirected to the sour gas reservoir, where sulphur dioxide reacts within the reservoir with  $H_2S$  to yield liquid sulphur.

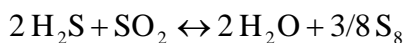
This means that a highly polluting substance such as  $SO_2$ , is recovered in the form of sulphur, and the  $CO_2$  is never released into the atmosphere but rather kept in internal circulation. In addition, the liquid sulphur thus formed can also be passed into a combustion chamber to form  $SO_2$  while releasing energy of 4500 BTU/pound. Once the  $H_2S$  reserves are exhausted, the underground reservoir will contain only liquid sulphur, to which the surplus of sulphur from tar sands petroleum refineries can be added for safer storage. This sulphur reserve will represent very little risk for the environment compared with the outside storage; in addition, this buried sulphur can be available for future generations.

AN INTERESTING ASPECT OF THIS PROCESS IS THAT IS GENERATES CONSIDERABLE ENERGY WITHOUT CO-PRODUCTION OF  $CO_2$ , THUS IT IS OF

## ADDITIONAL INTEREST IN AN ERA OF INCREASED CONTROL OVER GREENHOUSE GAS EMISSIONS.

Although there are still important technological issues to cope with, the process may not be very far from implementation: once the mixture of H<sub>2</sub>S/CO<sub>2</sub> gases has been burned to produce steam, the sulphur and carbon dioxides (mixed with water vapour, and nitrogen and excess oxygen coming from the air used for the combustion) are brought to a condenser and then to a stripper, where they are separated from the condensed water. Well-known graphite lined vessels can be employed for this step. After this, the SO<sub>2</sub> is divided from the rest of the gases by means of an absorber, and after the step of SO<sub>2</sub> solvent regeneration it is compressed and then injected

Down into the reservoir, at a pressure of about 3.000 – 12.000 psi and a temperature greater than 90 °C (depending on the reservoir conditions), the following reaction takes place, which realized a reduction in volume and pressure:



At this stage, the sulphur produced by the reaction will be subjected to a gravity drainage and it migrates toward the bottom of the reservoir. It could be questioned if the liquid sulphur can represent a risk to the productivity of the field, because the reservoir can become plugged with sulphur; Clark and colleagues reply that the produced sulphur cannot fill the pore structure of the reservoir, due to the density increase of the product; moreover, 100 years of experience available on movement of gas through liquid sulphur enable the researchers to exclude this hypothesis.

Another interesting feature of this technology is the possibility of using it also to produce energy from the **combustion of solid sulphur** which yields SO<sub>2</sub> and heat (used once again to generate water vapour); in this way, sulphur coming from stockpiles or from new field production can be burned and than reinjected, avoiding the great environmental problems of storing solid sulphur. The process of burning solid sulphur is interesting also because it does not produce water vapour in the reaction chamber, simplifying the design and the number of steps of the plant.

In conclusion, as well as supplying a considerable amount of energy and electrical potential, this new chemical technology is very ecological for it prevents the emission of polluting gases such as SO<sub>2</sub> and CO<sub>2</sub>. A number of additional chemical engineers and others are joining Clark's team, and with the help of an industrial partnership, they are beginning the erection of a test plant to be used as a prototype for a first commercial plant in 2004.

## CONCLUSION

Development of Kazakhstan field as Karachaganak and above all, Kashagan is a challenge because of the very high percentage of H<sub>2</sub>S concentration and very high reservoir pressure.

The high expected production rate of this field will give enormous quantities of S that will have to be removed from gas. The traditional technologies, as Claus process seem to be too expensive relating to the problem of storing the solid sulphur. The economical and environmental impacts could be affected the profitability of the whole project.

Other technologies, the H<sub>2</sub>S re-injection above all, will be able to guarantee the economic development of the fields where the problem of H<sub>2</sub>S is economically relevant.

## ACKNOWLEDGMENTS

We would like to express our gratitude for the assistance rendered by the following persons Mr. Novelli, Mr. Chiaramonte and Mr. Lockhart.

## APPENDIX A

The moles extracted are function of environmental pressure e temperature:

$$n_{ng} + n_{H_2S} = \frac{p_0 V_{in}}{RT_0} \quad [1]$$

The concentration can be expressed as:

$$\frac{n_{H_2S}}{n_{H_2S} + n_{ng}} = [H_2S] \quad [2]$$

Substituting [2] in [1] we obtain:

$$n_{H_2S} = \frac{p_0}{RT_0} V_{in} [H_2S] \quad [3]$$

The mass of gas is:

$$m_{H_2S} = n_{H_2S} PM_{H_2S} \quad [4]$$

The recovery factor of desulphuring plant is expressed as:

$$\eta = \frac{q}{m_{H_2S}}$$

Therefore the daily production of S can be expressed in function of volume of gas extracted and the molar concentration of H<sub>2</sub>S:

$$q = \eta PM_{H_2S} \frac{p_0}{RT_0} V_{in} [H_2S] = \beta V_{in} [H_2S]$$

## CONVERSION UNITS

$$1000 \text{ Nm}^3 = 6,2 \text{ boe}$$

$$1 \text{ t of S} = 714 \text{ Nm}^3$$

$$1 \text{ MBTU} = 293 \text{ KWh} = 27,2 \text{ Nm}^3 \text{ natural gas}$$

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