CO$_2$ ABATEMENT IN OIL REFINERIES: FIRED HEATERS

Report Number PH3/31
October 2000

CO$_2$ abatement in oil refineries: fired heaters

Background

Previous work by IEA GHG (see report PH3/8) showed that CO$_2$ emissions world-wide from oil refineries totalled about 700 million tonnes in 1994. The baseline projection is that by 2020 refinery CO$_2$ emissions will be about 840 million tCO$_2$/year. This projected increase is based on a combination of: (i) growth in demand for transport fuels and (ii) continued pressure to produce 'cleaner' fuels e.g. with a low sulphur content (necessitating more processing and increase emissions of CO$_2$/barrel of product). The major source of CO$_2$ emissions in refineries is the fired heaters (accounting for about 65% of the total according to the earlier report). It is concluded in the report that abatement options focussed on energy efficiency could hold refinery emissions of CO$_2$ steady until about 2010 but, after that, a significant off-set of emissions can only be achieved by more a radical approach.

The objective of this study is to examine abatement options based on the application of CO$_2$ capture technology to fired heaters. These options could be used to make a major reduction in CO$_2$ emissions from oil refineries. CO$_2$ storage is not addressed.

There are numerous ways in which emissions of greenhouse gases associated with the production and use of transport fuels could be reduced. For example, various co-production schemes have been suggested in which synthesis gas is generated and used to produce both transport fuels and electricity. Such wider issues are not considered here.[1] The base-case assumption here is an existing refinery that produces a conventional spectrum of oil-derived products.

Approach adopted

The study was organised in stages as follows:

- A base-case design for a 'generic' refinery was established. The objective was to define emission levels against which abatement options could be compared. Although any specific refinery will be different, the generic refinery is sufficiently representative that the conclusions of the study will be broadly applicable. Much of this phase of work was prepared with the assistance of BP in advance of the contractor's study activities.

- Eight potential methods of CO$_2$ capture were evaluated in outline. Only technologies capable of being applied to an existing refinery site in the near future were considered.

- Three options for CO$_2$ capture were selected for more detailed development and evaluation.

The study was done by Foster Wheeler Energy Ltd., Reading, England.

Thanks are due to Keith Howard, BP, Refining Technology Group and his colleagues who assisted greatly in defining the generic refinery and provided technical input throughout the study.

To aid understanding, the report is expressed in terms of 'refinery' units, such as barrels of oil, in addition to metric units.

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[1] A forthcoming report assesses CO$_2$ capture options in Fischer-Tropsch processing to produce a fuel that can be used in diesel engines.

Results and discussion

generic refinery

Table S1 summarises the characteristics of the generic refinery (more detailed tables are presented in section 2 of the report). The 200 000 barrel/day refinery emits approximately 1.5 million tonnes/year of CO_2, of which, $^2/_3$ i.e.1 million tonnes/year, is from fired heaters. Releases of CO_2 from flares, incineration, effluent processing, etc. are not tabulated; they would add about 3% to the total (see report PH3/8).

Table S1:Characteristics of the generic refinery and fired heater CO_2 emissions

Process unit (number of fired heaters/stacks)	Throughput (barrels/day)	CO_2 emitted (tonnes/year)
Crude distillation (1/1)	200 000	271 000
Vacuum distillation (1/1)	68 000	115 000
Distillate hydrotreater (1/1)	53 000	29 000
Reformer (5/1)	45 000	161 000
Gas Oil hydrotreater (2/1)	37 000	54 000
Hydrocracker (3/1)	40 000	166 000
Alkylation unit (1/1)	2 000	9 000
Hydrogen unit (1/1)	685 000Nm^3/day output	142 000
Total emission from fired heaters:		947 000
Hydrogen unit (CO_2 from process)		116 000
Total CO_2 emissions considered for abatement		1 063 000
Additional CO_2 emissions not considered for abatement:		
Catalytic cracker 'coke burn'		167 000
'Shadow' carbon associated with imported power and heat		224 000
Total emissions of CO_2 (excluding, flares, incineration, etc)		1 454 000

heater fuel

The heater fuel in the base-case is a homogeneous mix of 60% refinery gas and 40% imported natural gas. It is assumed that a refinery considering the application of CO_2 capture technology would not be burning fuel oils in the refinery heaters.[2]

heater configuration

The generic refinery has a typical mix of forced and natural draft heaters. Details are given in section 2 of the report. The CO_2 concentration in the untreated flue gas is between 8% and 9% by volume (wet basis). Some of the fired heaters have shared stacks. In total the generic refinery has 15 fired heaters and 8 stacks.[3] This heater-to-stack ratio (i.e. approximately 2:1) is typical of older sites; a more recently built site is likely to have a ratio of about 4:1. The implications of heater-to-stack ratio for CO_2 capture are dealt with in the report.

'Shadow' carbon

The generic refinery is not self-sufficient in electrical power and steam; these utilities are generated outside the battery limit. The concept of 'shadow' carbon was adopted to include in the calculations the CO_2 associated with providing these services. All the capture options considered increase the demand for utilities, in particular electricity, and the shadow carbon can have a significant influence on the net level of CO_2 emission abatement. This topic is dealt with in the report by a sensitivity study.

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[2] The choice of refinery fuels is a complex issue that can be influenced by local tax regimes. Fuel oil is often one of the lower value products and fired heaters are frequently design to accept both gas and oil feeds. Substituting natural gas for liquid fuels is a likely initial step towards CO_2 emission reduction.

[3] Not counting the cat.cracker and the power station.

CO_2 capture options

The study assessed the cost and effectiveness of CO_2 abatement by the following process routes:

- Capture from fired heater flue gases using a regenerable amine solvent. (See figure 3-1 on page 3-3 of the report.)
- Use of oxygen produced in an air separation unit (ASU) to burn the heater fuel (oxyfuel combustion). Flue gas is re-circulated to control the combustion temperature. (See figure 3-2 on page 3-6 of the report.)
- Use of a hydrogen-rich fuel gas in the fired heaters. CO_2 capture takes place before the fuel is burnt. The H_2-rich fuel gas is made from the refinery-produced gases supplemented by natural gas. Various combinations of hydrogen production technology and mole fraction H_2 in the fuel gas were assessed. (See figures 3-3 to 3-6, pages 3-9 to3-12 of the report.)

flue gas scrubbing with amine solvent

Although the fired heaters are at various locations distributed around the refinery, capture from flue gases using amine solvents (based on MEA[4]) is a better option than it might appear at first sight. A separate local absorber is used for each heater stack thus minimising the flue gas ducting required. Amine regeneration and CO_2 treatment and compression is done at a convenient central location. Amine solvent can be readily circulated back-and-forth between the absorbers and CO_2 recovery unit. Separation of these activities reduces both the cost and space required. (Approximate plot areas are shown in figure A3-2 of the report.)

oxyfuel combustion

There are no known commercial applications of refinery fired heaters modified to burn fuel in an oxygen-rich environment. However, the arrangement is common in other industries, e.g. glass manufacture. Discussions with burner manufacturers and the contractor's furnace design experts established that there appears to be no fundamental technical barriers.

With oxyfuel combustion the flue gas volume is reduced by about 75% to 80% relative to combustion in air. This makes installation of flue gas ducts from the heaters to a centralised location feasible. The flue gas is predominantly a mixture of CO_2 and water which is readily separated. The amount of oxygen required (approximately 150 t/hr) is sufficiently large that an air separation unit (ASU) is needed. The ASU is treated as an on-site unit incurring capital expenditure and operating costs. Power for the ASU is imported and carries associated shadow carbon.

H_2-rich combustion

Several variations based on firing H_2-rich fuels in the heaters are examined. In all cases it was assumed that the refinery gases would be converted to hydrogen.

If only the refinery gases are converted to hydrogen, additional heater fuel is required. If the additional fired heater fuel is natural gas, the H_2-rich fuel is about 85% (vol) hydrogen. In these cases, the achieved reduction in emissions from fired heaters is less than 70%. Such schemes might be of interest but were considered, for this study, to result in insufficient overall reduction in CO_2 emissions from the refinery.

The cases selected for detailed assessment are based on a heater fuel that is essentially 100% hydrogen. Hydrogen is produced from refinery gases and imported natural gas.

There is little to choose between a number of hydrogen technology options; the preferred options is largely dependant on local factors, e.g. whether surplus steam has a significant value. In a specific refinery the choice would be influenced by the existing hydrogen units and any potential need for additional hydrogen

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[4] Monoethanolamine.

for oil processing. Options based on steam methane reforming (SMR) and partial oxidation are examined in the report. CO_2 capture is by conventional physical or physio-chemical solvent.

The option selected for detailed assessment is based on air-blown catalytic partial oxidation (CAPO). In this option, as with the other options based on burning hydrogen, a significant additional amount of carbon is imported across the refinery battery limits. This process does not have a fired heater as the reaction is autothermal and preheat is obtained by process integration. (figure 3-6 page, 3-12).

An option of some interest (option 4, figure 3-3, page 3-9) involves using H_2-rich gas in all the fired heaters <u>except</u> the SMR furnace. Natural gas and recycled tail gas[5] are used in the SMR furnace. This processing scheme has the effect of concentrating CO_2 in the SMR furnace. CO_2 is captured using MEA on the furnace flue gas. As the main purpose of this study was to compare different types of CO_2 capture options, this 'hybrid' case was not selected for more detailed assessment. In emission reduction and cost terms this option appeared to be similar in attractiveness to the options selected for more detailed assessment.

CO_2 emission reductions

These results are presented in terms of on-site and net (i.e. including off-site) CO_2 emissions. The difference depends on the amount of shadow CO_2 assumed to be associated with import and export of utilities. (Sensitivities to the assumptions used in this study are shown on page 8-3.)

Table S2 presents results for CO_2 abatement <u>on the refinery site</u> for each of the 3 processes evaluated in detail. As can be seen, all 3 technology options can be used to avoid over 90% of the emissions from refinery heaters.[6] In terms of the overall reduction in emissions within battery limits this is about 70% except for the case of H_2-rich fuels where the emission reduction is less because extra natural gas is imported to make hydrogen.

Table S2: Reduction in <u>on-site</u> refinery emissions for each abatement option.

	Amine scrubbing	Oxyfuel	H_2-rich fuel
Total base-case refinery CO_2 emissions (thousands of tonnes/year).	1454	1454	1454
Base-case refinery emissions available for abatement. (CO_2 thousands of tonnes/year).	1063	1063	1063
CO_2 emission avoided (thousands of tonnes/year).	1014	1019	963
% reduction in 'fired-heater' CO_2 emissions.	95	96	91
% reduction in total refinery emissions of CO_2.	70	70	66

However, the 3 CO_2 abatement processes require additional electrical power and other services, the availability of which depends on the refinery site. For the purposes of the study it is assumed that, where additional power and steam is needed, it is obtained by burning natural gas in relatively inefficient plant. The study includes a sensitivity analysis to assess the impact of these assumptions. Although for any specific refinery the overall implications could be significant, the shadow CO_2 for each abatement option does not vary greatly. Table S3 presents net results for CO_2 abatement for the overall refining process.

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[5] The tail gas is from a pressure-swing adsorption (PSA) unit used to clean the hydrogen product. It contains 75% hydrogen and 19% methane (by volume). Addition of PSA appeared an effective way of reducing CO_2 emissions.

[6] For studies on capture processes IEA GHG normally specify a minimum CO_2 capture level of 85%. For this study the engineering judgement was made that higher levels of capture would be optimum.

Table S3: Reduction in overall emissions from refinery processing.
(Including the effects of shadow CO_2).

	Amine scrubbing	Oxyfuel	H₂-rich fuel
Total refinery CO_2 emissions (thousands of tonnes/year)	1454	1454	1454
Base-case refinery emissions available for abatement. (CO_2 thousands of tonnes/year)	1063	1063	1063
Net CO_2 emission avoided (thousands of tonnes/year.)	685	681	610
% reduction in 'fired-heater' emissions.	64	64	57
% reduction in total refinery emissions of CO_2.	47	47	42

As can be seen from table S3, even if the additional services are obtained from relatively inefficient plant, the net emissions of CO_2 from the refining operation can be reduced by over 40% if any of the 3 technologies are applied to the refinery's fired heaters.

The net extent of CO_2 abatement for a specific refinery is likely to lie within the range of results shown in tables S2 and S3, say, 45%-70%. It seems reasonable to assume that the reduction achieved would be towards the upper number as major reductions in refinery emissions of CO_2 would not be attempted if the additional services required carried a heavy penalty in shadow carbon.

costs

All the options for major reductions in CO_2 emissions incur a cost. The costs for CO_2 abatement using each of the 3 technologies are presented in table S4. A 10% discount rate is used and natural gas is assumed to cost 2$/GJ. Other assumptions are given in detail in section 6 of the report.

Table S4: The cost of avoiding CO_2 emissions in the generic refinery

	Amine	Oxyfuel	H₂ –rich fuel
Installed cost (inside battery limits) inUS$ million	146	137	203
Cost of CO_2 avoided $/t$CO_2$ (within refinery site boundary)	18	20	22
Net cost of CO_2 avoided $/t$CO_2$ (including shadow CO_2 for imported utilities)	27	28	42

The capital cost estimate does not allow for investment costs outside battery limit costs to supply the increased utility requirements.[7] This cost will depend on the existing capacity of utility systems and the extent of any required modifications. The cost of electricity is assumed to be 5 cents/kWh. (The costs assumed for utilities are given on page 7-2 of the report.)

For the amine scrubbing and oxyfuel firing options the cost of avoiding CO_2 emissions is less than 30$/tonne, even with conservative assumptions for the amount of shadow CO_2 associated with imported utilities. The relatively high cost of avoidance for the H₂-rich fuel option (particularly taking shadow CO_2 into account) is a consequence of the additional natural gas needed to make the hydrogen fuel.

For oxyfuel firing, the cost of sealing the fired heaters to obtain minimal leakage was assumed to be negligible. In practice, it may be impractical or uneconomic to seal some heaters. Any significant leakage would increase the cost of this option.

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[7] Note for instance, that the amine treatment option requires a large quantity of cooling water (see page A2-1 of the report).

Expert Group and other comments

There were no major comments from the Expert Group. This is to be expected as BP took an active interest in the progress and direction of the study.

Major conclusions

The major conclusions are as follows:

1. Abatement of CO_2 emissions from oil refinery fired heaters can reduce overall emissions from the refining process by between 40-70%. This abatement can be achieved by application of existing or near-term technology and does not require a major change in the nature and supply of transport fuels.

2. The abatement options all require additional utilities e.g. electricity. The net emission reduction depends on the amount of 'shadow' CO_2 associated with the provision of additional utilities. Clear accounting of emissions inside and outside battery limits is required. If imported utilities are essentially carbon-free the refinery emissions can be reduced by approximately 70%.

3. There are 3 near-term technologies available for CO_2 capture, each of which has a similar effectiveness. They are:
   - amine scrubbing of flue gas
   - oxyfuel combustion
   - H_2-rich fuels

The preferred option would probably be refinery specific; more than one option or hybrid options could be used.

4. The cost of CO_2 abatement lies in the range 20-40 US$/$tCO_2$ emission avoided for reduction in refinery emission of about 40%.

5. For the generic refinery used as a reference in the study, amine-based CO_2 capture from the flue gas of fired heaters is marginally more attractive than oxyfuel firing. This difference is not significant; the economics of scrubbing and oxyfuel are very similar.

Recommendations

The conclusions of this study need to be confirmed by consideration of their application in specific refineries.

The net level of CO_2 emission abatement achieved can depend significantly on the increased demand for utilities, in particular electricity. This should be a major consideration in any proposed scheme for CO_2 emission reduction.

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1.0 EXECUTIVE SUMMARY
(For further definition of terms used refer to attachment 6)

1.1 STUDY OBJECTIVE

The main objective of this study was to assess processing options that may be used to obtain major reductions in the emissions of CO_2 from fired heaters in oil refineries.

The study assessed the cost and effectiveness of CO_2 abatement by the following processing routes:

- Capture using amine solvents
- Use of oxygen for combustion
- Use of H_2 rich fuel

A reduction of 75% of the total base case CO_2 emissions was selected as the abatement target for the study.

1.2 STUDY SCOPE

The study comprised of three distinct phases referred to as "tasks":

Task 1
A base case design for a generic refinery was established together with a total a CO_2 emission level against which abatement options could be compared. This was a minor part of the overall study and included review and agreement of the generic refinery data and emissions.

Task 2
Potential methods for CO_2 abatement were investigated and evaluated. Only technologies capable of near- term application (requiring limited development) were considered. Vendor information was sought where required and indicative prices were calculated for all options considered. A summary tabulation comparing costs and CO_2 reduction potential of each option was prepared.

Task 3
Three options from task 2 were selected for further development. Detail developed for each option included process description, PFD, material balance, capital cost detail, levelised cost of CO_2 capture, DCF analysis and block flow diagram illustrating refinery carbon flows.

1.3 TASK 2 RESULTS

Eight options were evaluated during task 2 and the key results obtained are summarised in table 1-1 overleaf.

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The following points are noted:

- Capital costs were limited to the new process plant and existing plant modifications directly associated with the option ie the ISBL (Inside Battery Limits) costs. Capital expenditure for new / additional utilities which could be required for the option (ie located Outside Battery Limits, OSBL) eg additional cooling water facilities, power and steam generation, were not included in this generic work as these were considered refinery specific.

- The concept of 'Shadow CO$_2$' was introduced for the options to include for the CO$_2$ associated with providing electrical power and steam to the CO$_2$ abatement process. It was assumed these utilities were generated in facilities outside the battery limit with no CO$_2$ abatement ie producing 'shadow CO$_2$'. Similarly any steam generated by an option was credited with the avoided shadow CO$_2$ from the same facilities. 'Net' shadow CO$_2$ was thus the sum of the shadow produced and shadow avoided.

Table 1-1 Task 2 Results Summary

Option No	Option Description	Capital Cost MM$	Refinery CO$_2$ emissionsTPA (excludes shadow)	% Reduction from base case*	MM$ / % Reduction	Net Shadow TPA
1	Amine Treatment	146.2	47,356	96	1.5	330,200
2	Oxygen rich burning	136.6	43,372	96	1.4	338,429
3	Hydrogen Rich Fuel - SMR - 85% of RFG as H2 ex SMR	170.8	348,661	67	2.5	180,540
4	Hydrogen Rich Fuel - SMR - 100% of RFG as H2 ex SMR	230.1	21,166	98	2.3	273,059
5	Hydrogen Rich Fuel - SMR - 100% of RFG as H2 ex SMR H2 as Reformer Fuel	271.0	204,029	81	3.4	290,922
6	Hydrogen Rich Fuel - CAPO - 85% of RFG as H2 ex CAPO	148.7	452,726	57	2.6	208,597
7	Hydrogen Rich Fuel - CAPO - 100% of RFG as H2 ex CAPO	206.9	173,860	84	2.5	322,125
8	Hydrogen Rich Fuel - CAPO - Without Fired Heater Preheat	202.6	100,279	91	2.2	352,980

*1,062,780 TPA)

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Amine Treatment and Oxygen rich burning were found to be the most effective options in terms of reduction in refinery CO_2 emissions and capital cost per % CO_2 reduction. However it is acknowledged that both these options will attract a larger outside battery limit cost element than the hydrogen rich options (No's 3 through 8) to support the large increase in utilities required.

Amine Treatment and Oxygen rich burning also have the advantage of introducing no additional carbon directly into the refinery. In fact option 2 actually increases heater efficiency reducing the overall CO_2 produced. The hydrogen rich options by contrast all introduce additional carbon into the refinery ie CO_2 captured is greater than CO_2 avoided. Options 1 and 2 were therefore developed in task 3.

Of the hydrogen rich options considered options 3 and 6 were rejected because they did not reach the study target reduction.

Option 4 includes a Steam Methane Reformer fired on natural gas. It is attractive in terms of overall emission reduction and is similar to CAPO in terms of cost per % CO_2 reduction. It also has the advantage of concentrating the CO_2 recovery to one location i.e. the reformer flue gas. However, this option was not developed in task 3 because it was not considered appropriate to develop an option which created a large natural gas fired heater in the context of CO_2 abatement.

Option 5, a Steam Methane Reformer fired on hydrogen, is the highest capital cost per % CO_2 reduction option and the overall CO_2 reduction is not particularly attractive, therefore this option was not carried forward to task 3.

Option 8 gives the fourth best overall CO_2 reduction and, together with option 4, yields the best capital cost per % CO_2 reduction of all the hydrogen rich gas options. However this option was developed in task 3 rather than option 4 (which gives the best overall CO_2 reduction) as option 8 does not include a large natural gas fired heater.

The net shadow CO_2 was highest for options 2 and 8 where electrical power consumption is high for the cryogenic and feed gas/air compression demands respectively. However, the net shadow difference between the options which met the CO_2 reduction target (No's 1,2,4,5,7 and 8) did not vary greatly and did not significantly change the relative cost per % CO2 reduction when accounted for.

The options requiring a high LP steam demand for amine regeneration ie all options except option 2 will benefit significantly, in terms of reduced shadow CO_2, if large quantities of excess LP steam is available at the refinery in question.

The optimum CO_2 abatement solution implemented at an actual refinery, consideration of which is outside the scope of this report, is likely to involve a combination of elements taken from the various options considered above.

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1.4   TASK 3 RESULTS

The results of the DCF analysis and levelised cost of CO_2 removal for the three options carried forward from task 2 are shown in table 1-2 below:

Table 1-2

	Option 1 Amine Treatment	Option 2 Oxygen Rich Burning	Option 8 CAPO without preheat
Fixed Capital Cost MM$	-146.2	-136.6	-202.6
Operating Cost per year MM$	-28.071	-32.595	-31.483
Feed Cost per year MM$ (natural gas)	0	2.573	-7.8
DCF Net Present Value (NPV) MM$	-414.93	-421.74	-578.48
CO_2 Captured, t/ plant lifetime	22,813,695	21,148,088	26,440,628
CO_2 Avoided, t/ plant lifetime	22,813,695	22,936,680	21,656,273
CO_2 Avoided including shadow t/ plant lifetime	15,417,540	15,322,028	13,714,223
Levelised Cost of CO_2 removal (NPV/ t CO_2)			
$ / t CO_2 captured	-18.2	-19.9	-21.9
$ / t CO_2 avoided	-18.2	-18.4	-26.7
$ / t CO_2 avoided including shadow	-26.9	-27.5	-42.2

The NPV (Net Present Value) for the three options show that option 8 is the highest loss maker in line with its highest capital costs and operating plus feed costs. The NPV ranking of options 1 and 2 is reversed from the capital cost comparison. This is because although option 2 capital cost is less, this is outweighed by the operating cost which is 4.5 MM$ more per annum than option 1 mainly due to the ASU and CO_2 recovery high electrical power demand.

The levelised cost of CO_2 recovery has been defined here as the NPV divided by the tonnes of CO_2 either captured or avoided over the lifetime of the plant. The ranking of the options remains the same as a comparison on an NPV basis.

For option 1 the levelised cost of removal is the same for CO_2 captured as for CO_2 avoided because the carbon flow into the refinery remains unchanged compared with the base case. For option 2, more CO_2 is avoided than captured due to the improvement in heater efficiency with oxygen rich burning making the avoided cost less. By contrast, in option 8, more CO_2 is captured than avoided as more carbon is introduced to the refinery as fuel to the CAPO reactor, resulting in a higher avoided cost.

The quantity of CO_2 avoided decreases when the shadow emission is taken into account and the levelised cost of removal increases accordingly for all three cases. This is most significant for option 8 which combines the least profitable NPV, the lowest rate of CO_2 avoided and the largest CO_2 shadow.

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1.5   CONCLUSIONS & RECOMMENDATIONS

The study conclusions for CO$_2$ abatement at the generic refinery are summarised as follows:

- Amine Treatment and Oxygen rich burning are the most attractive options when comparing capital cost per % CO$_2$ reduction. However this excludes capital costs to support utilities for these options, most significantly cooling water, which will probably exceed those of the hydrogen rich burning alternatives. It is noted that the cooling water requirement of options 1 and 2 is of similar magnitude to the total capacity of typical refinery indirect seawater cooling systems.
  The hydrogen rich burning options are least attractive due to the additional CO$_2$ created within the process and the higher (inside battery limit) capital cost.

- The NPV of Amine Treatment is marginally more attractive (less loss making) than oxygen rich burning. The hydrogen rich burning option (CAPO) is significantly worse in line with its greater capital cost.

- Comparing the levelised cost of CO$_2$ capture of the options (NPV / tonnes CO$_2$ removed) results in the same ranking as comparing NPV alone. Hydrogen rich burning becomes significantly more expensive when its large 'shadow' CO$_2$ emission is accounted for.

Areas recommended for further study are shown as follows:

- Examine CO$_2$ abatement options against a range of utility balance / cost scenarios for a specific refinery.

- Examine options for increasing the availability of low grade heat for LP steam production in order to make the options involving amine regeneration more favourable

- Investigate integration with external power plant with respect to steam and power, heat balance and flue gas treatment.

- For the generic refinery:

  - Investigate abatement of FCC / RCC coke burn CO$_2$ (excluded from this study scope).

  - Investigate refinery process unit configuration for opportunities for CO$_2$ abatement.

- Establish and agree (with all relevant parties) a comprehensive and standardised formula for deriving shadow CO$_2$ to allow for variations in power station fuels and efficiency.

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2.0   BASIS OF DESIGN

2.1   INTRODUCTION

The main objective of the study was to evaluate potential modifications to fired heaters in a refinery that would significantly reduce their emissions of CO_2. The technologies considered for CO_2 abatement were as follows:

- Capture using amine solvents
- Use of oxygen for combustion
- Use of H_2 rich fuel

The initial study activity (Task 1) was to confirm and develop a 'generic' refinery based on information included in the Technical Study Specification IEA/CON/99/61 and subsequent discussions between IEA and Contractor. The developed design basis is included at the end of this section (Tables 2-3 and 2-4).

The subsequent tasks undertaken for the study (tasks 2 and 3), in which the above technologies were evaluated, are described in sections 3 and 4. Further design basis information specific to these tasks eg basis for hydrogen production unit size are included in these sections.

2.2   GENERIC REFINERY

The developed generic refinery is shown in Tables 2-3 and 2-4 at the end of this section.

The following points are noted:

- **Fuel Gas Composition**
  The base case heater fuel composition relates to a homogeneous mix of 60% refinery gas and 40% natural gas to all heaters. Any additional fuel gas consumption required was assumed to be natural gas ie refinery gas production is limited.

- **Exclusions from base case total CO_2 emission**
  CO_2 emissions from the Combined Heat and Power station and Catalytic Cracker coke burning have been excluded as the study scope is limited to emissions from conventional process heaters.

- **CO_2 for Reduction Target**
  For this study the percentage reduction achieved by application of various abatement technologies has been measured against a base case total CO_2 value. This value is the total CO_2 emissions of the generic refinery excluding CO_2 from the combined heat and power station and from the Catalytic Cracker coke burning. The target reduction is a minimum of 75%.

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- **CO_2 from Existing Hydrogen Plant**
  The existing refinery hydrogen plant has an atmospheric release of CO_2 from a CO_2 scrubbing system. This CO_2 is considered ready for capture and compression directly without further treatment.

- **Heater efficiencies, FD/ND and flue gas temperatures**
  An attempt has been made to distinguish between the forced draft and natural draft heaters. The generic refinery has a typical mix of heater configurations due to different duties and dates of installation. A generic efficiency value of 90% is used for all forced draft heaters with a corresponding flue gas exit temperature of 180 °C. For all natural draft heaters, an efficiency of 70.6% is used with a flue gas exit temperature of 350 °C.

- **Excess air**
  The quantity of excess air used for combustion affects the composition and quantity of the flue gases. The quantity of excess air required will vary from heater to heater dependent on its configuration. However, in order to provide a generic basis a value of 3% excess oxygen (by volume in flue gases) is used for all forced draft heaters and a value of 4% is used for all natural draft heaters.

2.3   SHADOW CO_2

Electrical power and steam requirements of abatement options considered were related to equivalent 'shadow CO_2' produced / avoided as follows:

- It was assumed any electrical power required was generated from a conventional natural gas fired power station (without CO_2 abatement) operating outside of the refinery fence at an efficiency of 35%.

- It was assumed any steam required was generated from a natural gas fired boiler operating outside of the refinery fence at an efficiency of 85%. (Similarly any steam generated 'avoided' shadow CO_2 from an equivalent boiler).

2.4   SITE CONDITIONS

Location: Netherlands coast
Ambient air temperature: 9°C
Ambient air relative humidity: 60%
Ambient air pressure: 1.013 bara

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2.5   PRODUCTS

Recovered CO$_2$:

Target recovery - flue gas (using Amine)   95%

Target recovery
 – CAPO/SMR effluent gas (using Amine)   99.95%

Delivery point   plant limit
Purity   >99%
Pressure (barg)   110
Temp ^{o}C   ambient

2.6   RAW MATERIALS

2.6.1   Natural Gas

Table 2-1

Component	Mol%
CH$_4$	94.8
C$_2$H$_6$	3.8
C$_3$H$_8$	0.2
CO$_2$	0.5
N$_2$	0.7
Total	100.0

H2S normal   vol ppm   1 - 2
H2S max   vol ppm   10
Pressure (@ process BL)   20 barg

2.6.2   Refinery Fuel Gas

Table 2-2

Component	Mol%
H$_2$	17.4
CH$_4$	44.4
C$_2$H$_6$	9.3
C$_3$H$_8$	9.9
C$_4$H$_{10}$	16.1
C$_5$H$_{12}$	2.1
C$_6$H$_{14}$	0.3
CO$_2$	0.5
Total	100.0

H2S normal   vol ppm   100
H2S max   vol ppm   3000
Pressure (@ process BL)   2 barg

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2.7 UTILITIES

2.7.1 Cooling water

Indirect sea water cooling is assumed. Summer average sea water inlet temperature 12°C, maximum temperature rise 7°C.

2.7.2 Steam

LP steam at 3.5 barg / 160 ^{o}C

MP steam at 20 barg / 300 ^{o}C

HP steam at 40 barg / 400°C

2.8 CO2 CAPTURED / AVOIDED

This definition is used in the analysis of the task 3 options and is included in section 4.2.

Table 2-3: Generic Refinery Study - Heater and CO2 Study Basis (English Units)   REV 1

Unit	Throughput	No Of Heaters	Stack	FD/ND	Flue gas Temp	Heater Description	Estimated Heat Absorbed	Efficiency	Calculated Heat Released	Fuel Consumption (Note 6)	(Notes 7 and 9)	Dioxide (Note 10)
	bpsd				(Note 14) F		10^6BTU/h	%	10^6BTU/h	lb/h	TPA	TPA
Crude Unit	200000	1	A	FD	350	Charge Heater	478.0	90.0%	531.1	25725	96084	271369
Vacuum Unit	68000	1	B	FD	350	Charge Heater	202.0	90.0%	224.4	10871	40605	114679
Distillate HDT (Note 1)	53000	1	C	ND	660	Charge Heater	40.0	70.6%	56.6	2744	10249	28945
Reformer (Note 2)	45000	5	D	FD	350	Charge Heater	83.0	90.0%	92.2	4467	16684	47121
			D	FD	350	1st Interheater	96.0	90.0%	106.7	5166	19297	54501
			D	FD	350	2nd Interheater	46.0	90.0%	51.1	2476	9247	26115
			D	FD	350	3rd Interheater	22.0	90.0%	24.4	1184	4422	12490
			D	FD	350	Debut Reboiler	37.0	90.0%	41.1	1991	7437	21006
Gas Oil Hydrotreater	37000	2	E	ND	660	Main Heater	50.0	70.6%	70.8	3430	12811	36181
			E	ND	660	Reboiler	25.0	70.6%	35.4	1715	6405	18091
Hydrogen unit (Note 3 & 13)	24.2 MSCFD	1	F	ND	660	Reformer Furnace	196.0	70.6%	277.6	13445	50218	141830
	-	-	-	-	-	CO2 Recovery	-	-	-	-	-	115660
Hydrocracker	40000	3	B	ND	660	Hot Hydrogen recycle	63.0	70.6%	89.2	4322	16142	45588
			B	ND	660	Reboiler 1	102.0	70.6%	144.5	6997	26134	73810
			B	ND	660	Reboiler 2	64.0	70.6%	90.6	4390	16398	46312
Alkylation Unit (Note 4)	2000	1	G	FD	350	Reboiler	16.0	90.0%	17.8	861	3216	9083
Total CO2 available for capture							**1520**		**1854**	**89783**	**335349**	**1062781**
Cat Cracker Coke Burn (Note 11)		1	H	FD	350	Regenerator/CO Boiler					50700	167209
Total CO2 for Reduction												**1229989**
Combined Heat & Power(Note 5&12)	45MWe	1	I						439.1	21268	79437	224354
TOTAL											**750134**	**1454343**

NOTES:

Note 1: Reboil provided by heat integration

Note 2: Average over a cycle

Note 3: Hydrogen Unit assumes 85% purity; H2 quoted is pure H2

Note 4: "Throughput" is Alkylate produced

Note 5: Estimate based on BP Amoco experience; assumes overall efficiency of 35%

Note 6: FW calculated the fuel gas LHV as 48.024 MJ/kg (20646 Btu/lb). Refinery
    fuel gas is 60% refinery fuel gas with the remainder being natural gas.
    Compositions were provided in email from H Audus 21/3/00

Note 7: Time efficiency is 94%

Note 8: Plot contains 17 heaters and 9 (A-I) stacks. This ratio ofapproximately 2:1
    is typical of older generation sites.
     With more recent build sites the heater/stack ratio is more likely to be approximately 4:1

Note 9: Fuel Total excludes Cat Cracker Coke Burn. CO2 produced includes FCC
    Coke burn and assumes that 90wt% of coke is carbon which is 100%
    converted to CO2

Note 10: All carbon in the fuel is assumed to be 100% converted to CO2

Note 11: Cat Cracker flue gas not considered for capture as part of this study. However,
    associated CO2 will be included in calculation of overall refinery reduction.

Note 12: CO2 emission from the power station is excluded from the study basis

Note 13: Hydrogen plant CO2 from process heater firing normal flue gas mix. In addition
    there is a CO2 vent from an amine scrubbing system on the H2 plant
    (115660 tpa) which can be captured directly for compression.

Note 14: Typical flue gas temperatures are estimated. It is assumed forced draft heaters
    incorporate preheat (I.e. FD and ID fans).

Table 2-4: Generic Refinery Study - Heater and CO2 Study Design Basis (Metric Units)   REV 1

Unit	Throughput	No Of Heaters	Stack	FD/ND	Flue gas Temp	Heater Description	Estimated Heat Absorbed	Efficiency	Calculated Heat Released	Fuel Consumption (Note 6)	(Notes 7 and 9)	Dioxide (Note 10)
	m3/ day				(Note 14) C		kW	%	kW	kg/h	TPA	TPA
Crude Unit	31794	1	A	FD	180	Charge Heater	140109	90.0%	155677	11670	96097	271369
Vacuum Unit	10810	1	B	FD	180	Charge Heater	59209	90.0%	65788	4932	40610	114679
Distillate HDT (Note 1)	8425	1	C	ND	350	Charge Heater	11725	70.6%	16605	1245	10250	28945
Reformer (Note 2)	7154	5	D	FD	180	Charge Heater	24329	90.0%	27032	2026	16686	47121
			D	FD	180	1st Interheater	28139	90.0%	31266	2344	19300	54501
			D	FD	180	2nd Interheater	13483	90.0%	14981	1123	9248	26115
			D	FD	180	3rd Interheater	6449	90.0%	7165	537	4423	12490
			D	FD	180	Debut Reboiler	10845	90.0%	12050	903	7438	21006
Gas Oil Hydrotreater	5882	2	E	ND	350	Main Heater	14656	70.6%	20756	1556	12812	36181
			E	ND	350	Reboiler	7328	70.6%	10378	778	6406	18091
Hydrogen unit (Note 3 & 13)	685300	1	F	ND	350	Reformer Furnace	57451	70.6%	81364	6099	50225	141830
	-	-	-	-	-	CO2 Recovery	-		-	-	-	115660
Hydrocracker	6359	3	B	ND	350	Hot Hydrogen recycle	18466	70.6%	26153	1961	16144	45588
			B	ND	350	Reboiler 1	29898	70.6%	42342	3174	26137	73810
			B	ND	350	Reboiler 2	18759	70.6%	26568	1992	16400	46312
Alkylation Unit (Note 4)	318	1	G	FD	180	Reboiler	4690	90.0%	5211	391	3217	9083
Total CO2 available for capture							**445535**		**543336**	**40731**	**335393**	**1062780**
Cat Cracker Coke Burn (Note 11)		1	H	FD	180	Regenerator/CO Boiler					50700	167209
Total CO2 for Reduction												**1229989**
Combined Heat & Power(Note 5&12	45MWe	1	I						128705	9648	79448	224353
TOTAL											**750235**	**1454342**

NOTES:

Note 1: Reboil provided by heat integration
Note 2: Average over a cycle
Note 3: Hydrogen Unit assumes 85% purity; H2 quoted is pure H2
Note 4: "Throughput" is Alkylate produced
Note 5: Estimate based on BP Amoco experience; assumes overall efficiency of 35%
Note 6: LHV of refinery fuel gas is 47.79 MJ/kg (20546 BTU/lb). LHV of natural gas is
    48.593 MJ/kg (20891 BTU/lb) fuel required; the remainder is natural gas)
    Fuel gas is 60% of total fuel required with the remainder being natural gas
    Hence average LHV of fuel fired = 20684 BTU/lb
Note 7: Time efficiency is 94%
Note 8: Plot contains 17 heaters and 9 (A-I) stacks. This ratio ofapproximately 2:1
    is typical of older generation sites.
    With more recent build sites the heater/stack ratio is more likely to be approximately 4:1

Note 9: Fuel Total excludes Cat Cracker Coke Burn. CO2 produced includes FCC
    Coke burn and assumes that 90wt% of coke is carbon which is 100%
    converted to CO2
Note 10: All carbon in the fuel is assumed to be 100% converted to CO2
Note 11: Cat Cracker flue gas not considered for capture as part of this study. However,
    associated CO2 will be included in calculation of overall refinery reduction.
Note 12: CO2 emission from the power station is excluded from the study basis
Note 13: Hydrogen plant CO2 from process heater firing normal flue gas mix. In addition
    there is a CO2 vent from an amine scrubbing system on the H2 plant
    (115660 tpa) which can be captured directly for compression.
Note 14: Typical flue gas temperatures are estimated. It is assumed forced draft heaters
    incorporate preheat (I.e. FD and ID fans).

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3.0   TASK 2 DISCUSSION

3.1   INTRODUCTION

In this task, potential modifications to fired heaters and/or their fuel supply were developed in outline with the objective of reducing overall emissions of CO$_2$ from the refinery; these modifications were then assessed.

The modifications considered during this task are shown in Table 3-1 overleaf.

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Table 3-1

Option No	Option Title	Option objective
1	Amine Treatment	95% recovery of CO_2 from flue gas
2	Oxygen rich burning .	95% recovery of CO_2 from flue gas
3	Hydrogen Rich Fuel – Steam Methane Reforming (SMR) 85% of refinery fuel gas as H_2 ex SMR (Firing natural gas plus PSA tail gas in SMR heater, amine treat SMR flue gas)	No recovery of CO_2 from process heater flue gases. CO_2 recovery within SMR as follows: • 95% recovery of CO_2 from SMR heater flue gas • 99.5% recovery of CO_2 from reformer gas ex LT shift reactor
4	Hydrogen Rich Fuel – Steam Methane Reforming (SMR) 100% of refinery fuel gas as H_2 ex SMR (Firing natural gas plus PSA tail gas in SMR heater, amine treat SMR flue gas)	As above
5	Hydrogen Rich Fuel – Steam Methane Reforming (SMR) 100% of refinery fuel gas as H_2 ex SMR (Firing hydrogen product in SMR heater)	No recovery of CO_2 from process heaters or SMR heater flue gas CO_2 recovery within SMR limited to 99.5% recovery of CO_2 from reformer gas ex LT shift reactor.
6	Hydrogen Rich Fuel – Catalytic Air Partial Oxidation (CAPO) 85% of refinery fuel gas as H_2 ex SMR (using hydrogen product fired heater to preheat feed)	No recovery of CO_2 from process heaters or CAPO feed heater flue gas CO_2 recovery within CAPO limited to 99.5% recovery of CO_2 from gas ex separator.
7	Hydrogen Rich Fuel – Catalytic Air Partial Oxidation (CAPO) 100% of refinery fuel gas as H_2 ex SMR (using hydrogen product fired heater to preheat feed)	As above
8	Hydrogen Rich Fuel – Catalytic Air Partial Oxidation (CAPO) 100% of refinery fuel gas as H_2 ex SMR (using reactor effluent to preheat feed, no feed heater)	No recovery of CO_2 from process heaters CO_2 recovery within CAPO limited to 99.5% recovery of CO_2 from gas ex separator

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Three options were investigated for each of the SMR and CAPO processing routes whereas only one option was considered for amine treatment and oxygen rich burning. This is because, for hydrogen rich burning, different hydrogen production rates were selected to characterise different levels of hydrogen enrichment of the fuel gas (options 3,4,6 and 7 – refer to section 3.4.1) and additional process configurations were developed to eliminate SMR natural gas burning (option 5 – section 3.4.4) and eliminate the CAPO feed heater requirement (option 8 – section 3.4.7).

At the conclusion of task 2 the assessment of all the modifications listed in 3.1 were summarised in a tabulation giving capital cost against CO$_2$ emissions (including 'shadow' CO$_2$ emissions). This summary is included in attachment 1.

The following subsections provide background to the key aspects of design work done to support the study and assess each modification. Process descriptions are included in section 5.

3.2   AMINE TREATMENT (OPTION 1)

3.2.1 Discussion

The CO$_2$ flows associated with this option are shown in Figure 3-1 below:

Figure 3-1: Amine Treatment (Option 1)
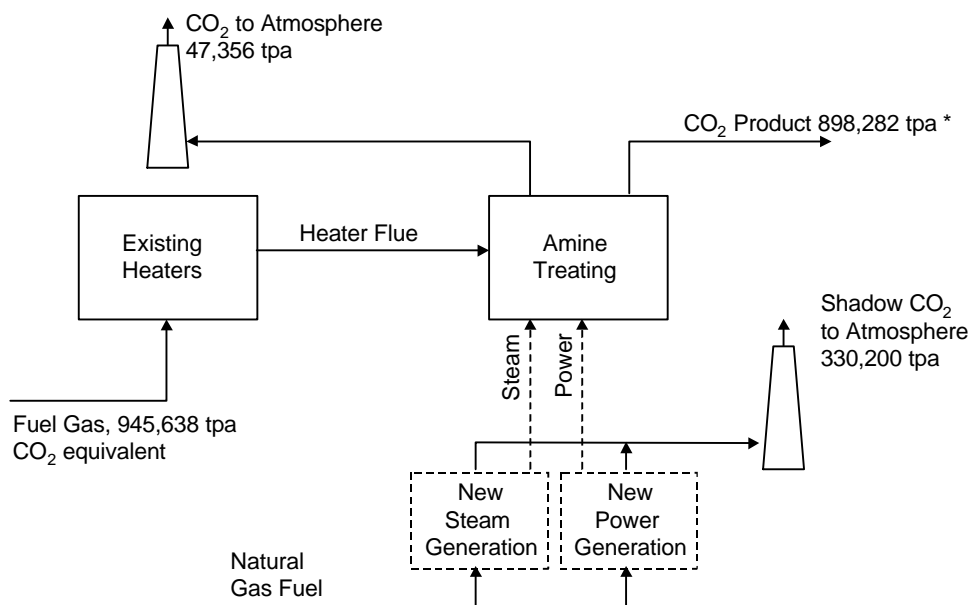
 * An additional 115,660 tpa CO$_2$ product from the existing refinery H$_2$ plant is also captured as product

95% of the equivalent CO$_2$ in the fuel gas to the process heaters is recovered from the amine regenerator. This yields an overall CO$_2$ reduction of 96% before the shadow CO$_2$ is accounted for (see Attachment I). The shadow CO$_2$

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lowers this reduction to 64%, however this is improved to 86% if the shadow associated with LP steam is excluded. Whereas it might be argued that some excess LP steam could be available it is highly unlikely that this will amount to the 118 MW required.

Amine treatment and oxygen rich burning give a lower capital cost per % of CO$_2$ reduction than the hydrogen rich fuel cases. However the outside battery limits investment in terms of utility provision for these options, which are excluded in this report, will probably be greater than the hydrogen rich options.

A circulating raw water stream, with a small make up and purge, is used to direct contact cool the flue gas. The raw water leaving the direct contact cooler is cooled against cooling water. This was the amine vendor's standard configuration and there is obvious potential here for energy recovery and reduction of the CO$_2$ shadow, especially for the higher temperature flue gas from the natural gas furnaces.

The CO$_2$ equivalent input into the refinery for amine treatment is the same as the base case i.e. the heaters run at the same efficiency and no additional fuel is burnt apart from that associated with the shadow production.

Amine treatment therefore appears to offer a relatively attractive means of CO$_2$ abatement especially if LP steam for amine regeneration is in excess in the refinery considered. This option was therefore carried forward for development in task 3.

3.2.2 Heater / stack ratio

The generic refinery heater / stack ratio is 2.14:1 which is a typical value for older installations. The implications of the amine treatment option having fewer stacks were considered as part of task 2.

In the base case, the 15 process heaters are served by 7 stacks. Combination of the following stacks reduces the heater / stack ratio to 3.75:1 i.e.:

B&F
C, E and G
(Stacks A and D remain as independent stacks)

A smaller number of larger absorbers are required to serve the B/F and C/E/G stacks, similarly the ducting to/from the absorbers increases in size but decreases in length.

The total installed cost of amine treatment with the two heater stack ratios is given in the table overleaf:

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heater / stack ratio	Total installed cost (MM$)
2:1	146.2
4:1	132.5
Saving	13.7 (9.4%)

3.3   OXYGEN RICH BURNING (OPTION 2)

3.3.1 Selection of oxygen production route

In arriving at a conventional cryogenic air separation unit for this option, the following alternatives were considered and rejected:

- **PSA:** 93 –95% pure oxygen (typical purity for oxygen enriched combustion) can be produced in vacuum pressure adsorption (VPSA) systems. These are most cost effective in the 20 to 150 tonne / day oxygen capacity range. This option requires about 150 tonne per hour.

- **Membrane:** Membrane technology is more typically used to obtain a high purity nitrogen product yielding a waste enriched oxygen stream (30 to 50%). In addition special membrane systems can be designed to produce up to 85% oxygen. These systems are competitive with PSA technology up to 25 tonne / day of equivalent oxygen, again too small for this study's requirement.

3.3.2 Discussion

The CO$_2$ flows associated with this option are shown in Figure 3-2 overleaf. The heater flue gases are combined and routed to a common CO$_2$ Recovery Unit and the tail gas is vented to atmosphere via a suitable adjacent stack.

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Figure 3-2: Oxygen Rich Burning (Option 2)
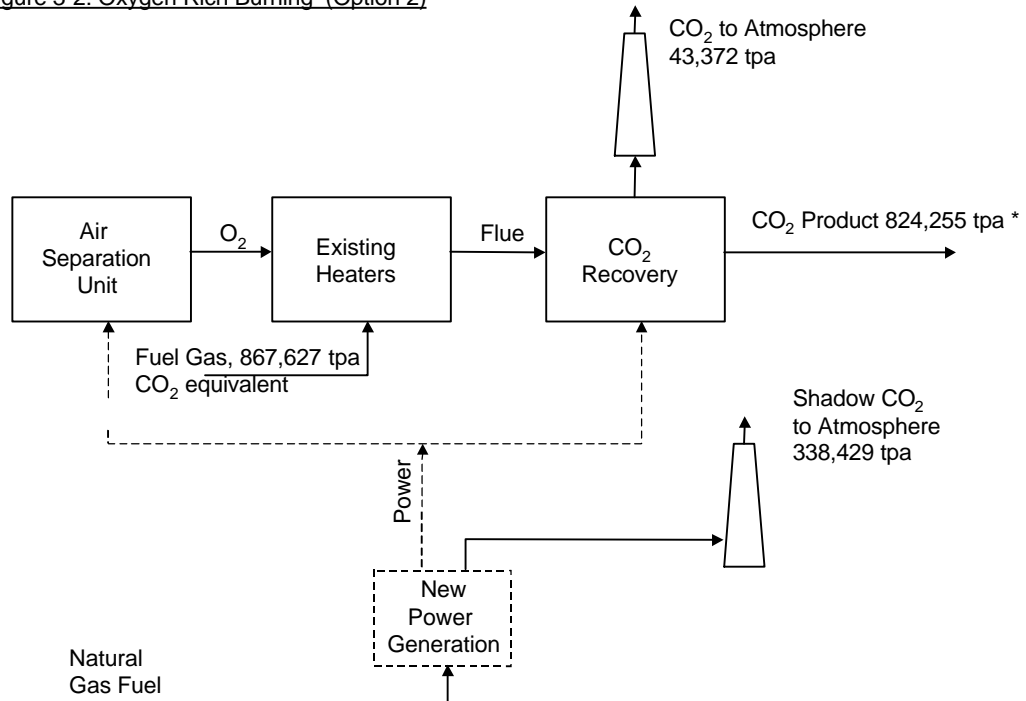

* An additional 115,660 tpa CO_2 product from the existing refinery H_2 plant is also captured as product

95 % of the equivalent CO_2 in the fuel gas to the process heaters is recovered in the CO_2 Recovery Plant. This yields an overall CO_2 reduction of 96% before the shadow CO_2 is accounted for (see Attachment I). The shadow CO_2 lowers this reduction to 64%, demonstrating the high electrical power demand of this option.

It is noted that air leakage into the process heaters is negligible and is not taken into account when determining the flue gas composition.

Oxygen rich burning gives a lower capital cost per % CO_2 reduction than the hydrogen rich fuel cases. This capital cost includes for the ASU, heater modifications (new burners, ducting, and flue gas re-circulation blowers), changes to the stacks, the CO_2 Recovery Unit and CO_2 compression. However, as for the amine treatment option, the outside battery limit investment for oxygen rich burning, which has been excluded in this report, will probably be greater than the hydrogen rich alternatives.

The CO_2 input into the refinery is about 8% less than the base case. This is because the efficiency of natural draft furnaces in the generic refinery increases to 90% as a result of the oxygen rich firing modifications.

The flue gas is direct contact cooled with water (see section 5) for consistency with the amine treatment case. There is obvious potential here for energy recovery and reduction of the CO_2 shadow, especially as the flue gas from the

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oxygen rich furnaces is some 20°C higher than the base case natural draft furnace temperature.

Although, currently, there are no commercial applications of refinery furnaces modified to burn fuel in an oxygen rich environment, the arrangement is common in other industries e.g. glass and steel manufacture. In addition on cost and CO$_2$ reduction potential the option is attractive and was therefore selected for development in task 3.

3.4 HYDROGEN RICH FUEL (OPTIONS 3,4,5,6,7 & 8)

3.4.1 General

The SMR and CAPO heat and material balances have been based on typical reformer/reactor inlet and outlet temperatures, steam / carbon ratio and approach to equilibrium temperature and do not represent optimised solutions. There is a trade off between these criteria, hydrogen production and capital cost that needs to be further studied for the actual configuration investigated. The choice of hydrogen technology for hydrogen rich fuel is a fine one that is largely dependent on local factors particularly the steam balance.

The SMR design / costing work done to support the study has been based on a Terraced Wall (TW) Reformer (i.e. the Foster Wheeler proprietary design) as it is typical for this service in refineries. For option 5, where hydrogen product is fired in the SMR heater, the capacity of the unit is 142 MMSCFD of product hydrogen which is at the capacity limit of a single (dual cell)TW Reformer. At this capacity, a top fired furnace, typically employed for very high throughputs, could be installed at lower cost than a TW Reformer depending on the economic criteria of the particular refinery considered.

For the hydrogen rich options two cases were selected to characterise enrichment of the heater fuel gas with hydrogen:

- 85mol % of the total refinery gas derived from hydrogen product from the hydrogen production unit. The remaining 15% is natural gas

- 100mol% of total refinery gas derived from hydrogen product from the hydrogen production unit.

It was considered these two cases would give a good range of hydrogen plant size against which reductions in CO$_2$ emissions could be judged. Table 3-2 overleaf summarises the effect of supplementing fuel gas with pure hydrogen, the selected cases are highlighted:

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Table 3-2: Supplementing Fuel Gas with Hydrogen

Mol frac H_2	Mol frac Ref gas	Mol frac nat gas	Mol Wt of combined Gas	H_2 requirement from new H_2 plant kg/h	% CO_2 reduction in emissions from refinery heaters
1	0.00	0	2.02	16354	**100.0**
0.95	0.00	0.05	2.76	13878	**85.7**
0.85	0.00	0.15	4.24	10232	**64.5**
0.75	0.00	0.25	5.73	7678	**49.9**

It was determined that the 85% hydrogen in fuel gas options were not effective in reducing CO_2 emissions in terms of capital cost per percentage CO2 reduction (see sections 3.4.3 and 3.4.6). This was because, compared with the 100% Hydrogen in fuel gas options, a substantial quantity of carbon remains in the fuel gas at 85% enrichment which is not compensated for by the reduced capital cost of the smaller SMR/ CAPO units.

All the hydrogen rich fuel options include amine treatment of CO_2 rich streams which are tabulated below for comparison.

Table 3-3: Supplementing Fuel Gas with Hydrogen

Option No.	Amine Treatment of CO₂ Rich Stream		
	Refinery Process Heater Flue Gas	SMR / CAPO Heater Flue Gas	Shift Reactor Effluent Gas
3	No	Yes	Yes
4	No	Yes	Yes
5	No	No	Yes
6	No	No	Yes
7	No	No	Yes
8	No	N / A	Yes

As for option 1, the hydrogen rich options will benefit in terms of reduced CO_2 shadow if excess LP steam for amine regeneration is available at the refinery considered.

3.4.2 Option 4 (SMR – 100% RFG as H_2, with PSA, NG to reformer heater)

The CO_2 flows associated with this option are shown in Figure 3-3 overleaf:

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Figure 3-3: SMR, 100% RFG as H$_2$, with PSA, NG to Reformer Heater (Option 4)
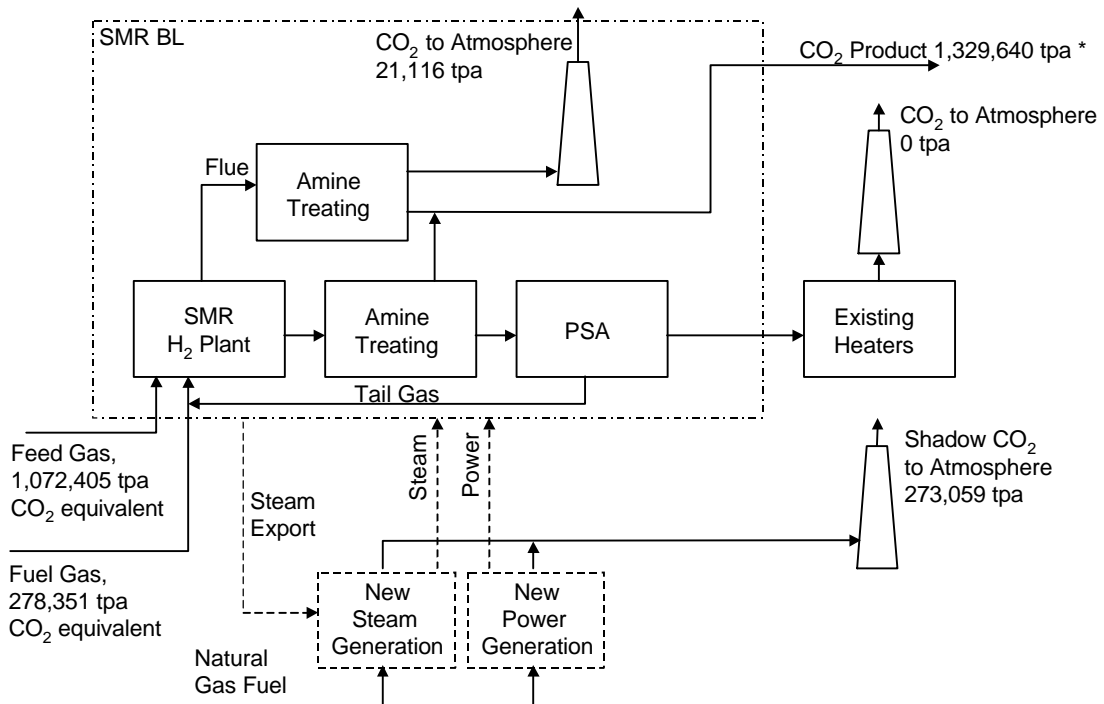
* An additional 115,660 tpa CO$_2$ product from the existing refinery H$_2$ plant is also captured as product

The single source of CO$_2$ emission in this option is the reformer furnace, 95 % of the CO$_2$ in this furnace flue gas is recovered. The refinery furnaces, firing 99.9% hydrogen, produce no CO$_2$ emission. This option yields an overall CO$_2$ reduction of 98% before the shadow CO$_2$ is accounted for (see Attachment I). The shadow CO$_2$ lowers this reduction to 72%.

This option gives a higher capital cost per % CO$_2$ reduction than the amine treatment or oxygen rich cases and is similar to the CAPO options.

The CO$_2$ equivalent input into the refinery is 43% more than the base case. Of this 43%, 30% is included in the fuel gas to the SMR furnace and 13% is in the feed gas.

Option 4 is attractive in terms of overall emission reduction and is similar to CAPO in terms of cost per % CO$_2$ reduction. An SMR with PSA therefore appears an effective route for CO$_2$ capture, provided that the PSA tail gas can be burnt in the reformer heater allowing CO$_2$ recovery to be concentrated in one location i.e. the reformer flue gas. However, this presupposes flue gas treatment is available and considered acceptable as part of a hydrogen enrichment option. This process route was not developed in task 3 because it was not considered appropriate to develop an option which created a large natural gas fired heater in the context of CO$_2$ abatement.

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3.4.3   Option 3 (SMR – 85% RFG as H_2, with PSA, NG to reformer heater)

This option is more expensive than option 4 in terms of capital cost per % CO_2 reduction, it also does not meet the CO_2 reduction target. It was therefore not further developed in task 3.

3.4.4   Option 5 (SMR – 100% RFG as H_2, without PSA, H_2 product to reformer heater)

The CO_2 flows associated with this option are shown in Figure 3-4 below:

Figure 3-4: SMR, 100% RFG as H_2, without PSA, H_2 Product Reformer Heater (Option 5)
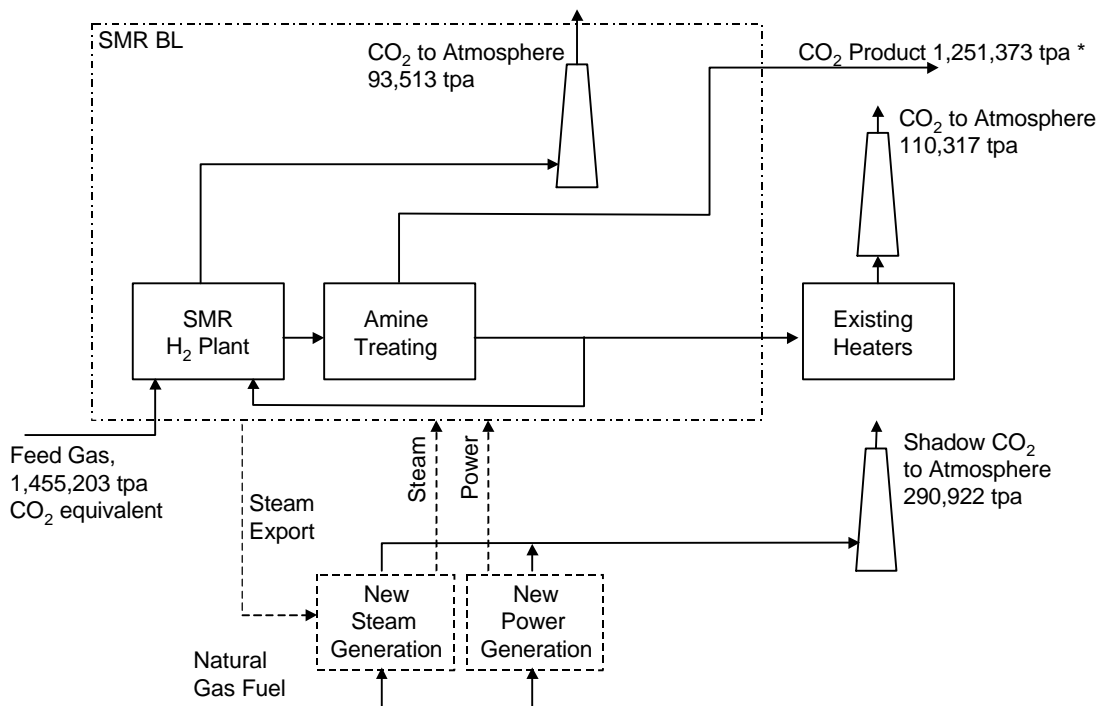
* An additional 115,660 tpa CO_2 product from the existing refinery H_2 plant is also captured as product

The two areas of CO_2 emission in this option are the reformer furnace and the refinery furnaces where the flue gases resulting from hydrogen rich gas combustion are not amine treated. The SMR hydrogen rich gas leaving the reactor effluent amine treatment, which removes 99.95% of the CO_2, contains a small quantity of carbon based gases.

Option 5 gives the highest capital cost per % CO_2 reduction of all the options. In addition, this option puts the greatest equivalent CO_2 into the process at 54% more than the base case.

This configuration is disadvantaged by the fact that although the hydrogen rich gas is more than 95% hydrogen, the large quantity of it required by the refinery and reformer heaters generates a significant quantity of CO_2. Capture of this CO_2 is not economically justified because of its low concentration. The

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CAPO equivalent option (option 7) is similarly disadvantaged. By contrast option 4 concentrates all the CO$_2$ in one location ensuring a better recovery.

As this is the highest capital cost per % CO$_2$ reduction option and the overall CO$_2$ reduction is not particularly attractive, this option has not been carried forward to task 3.

3.4.5   Option 7 (CAPO – 100% RFG as H$_2$, with fired heater for preheat)

The CO$_2$ flows associated with this option are shown in Figure 3-5 below:

Figure 3-5: CAPO, 100% RFG as H$_2$, with Fired Heater for Preheat (Option 7)
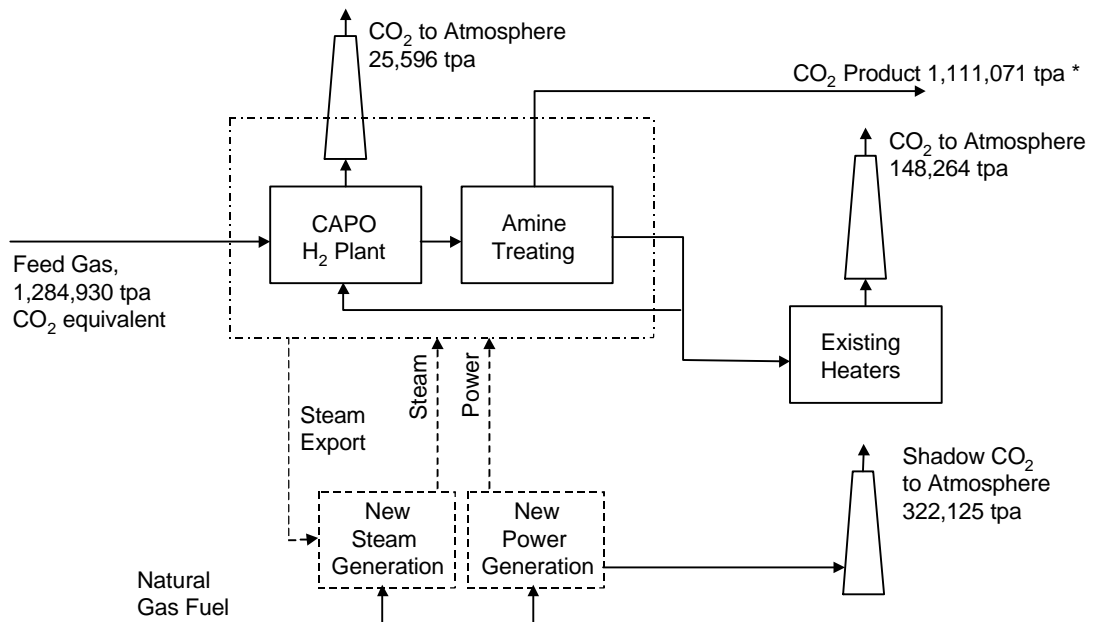
* An additional 115,660 tpa CO$_2$ product from the existing refinery H$_2$ plant is also captured as product

The two areas of CO$_2$ emission in this option are the preheater furnace and the refinery furnaces where the flue gases resulting from hydrogen rich gas combustion are not amine treated. The CAPO hydrogen rich gas leaving the reactor effluent amine treatment, which removes 99.95% of the CO$_2$, contains a small quantity of carbon based gases.

This option yields an overall CO$_2$ reduction of 84% before the shadow CO$_2$ is accounted for (see Attachment I). The shadow CO$_2$ lowers this reduction to 53%.

This option has a higher cost in terms of MM\$ capital cost per % CO$_2$ reduction than the amine treatment or oxygen rich cases and is similar to option 4. The CO$_2$ input into the refinery is about 36% more than the base case.

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As this option is not particularly attractive in terms of overall emission reduction and is no better than Steam Methane Reforming in terms of capital cost per % CO_2 reduction, it has not been further developed in task 3.

3.4.6 Option 6 (CAPO – 85% RFG as H_2, with fired heater for preheat)

This is similar to option 7 in terms of capital cost per % CO_2 reduction, with a worse overall CO_2 reduction which does not meet the target, it was therefore not further developed in task 3.

3.4.7 Option 8 (CAPO – 100% RFG as H_2, without fired heater for preheat)

The CO_2 flows associated with this option are shown in Figure 3-6 below:

Figure 3-6: CAPO, 100% RFG as H_2, without Fired Heater for Preheat (Option 8)
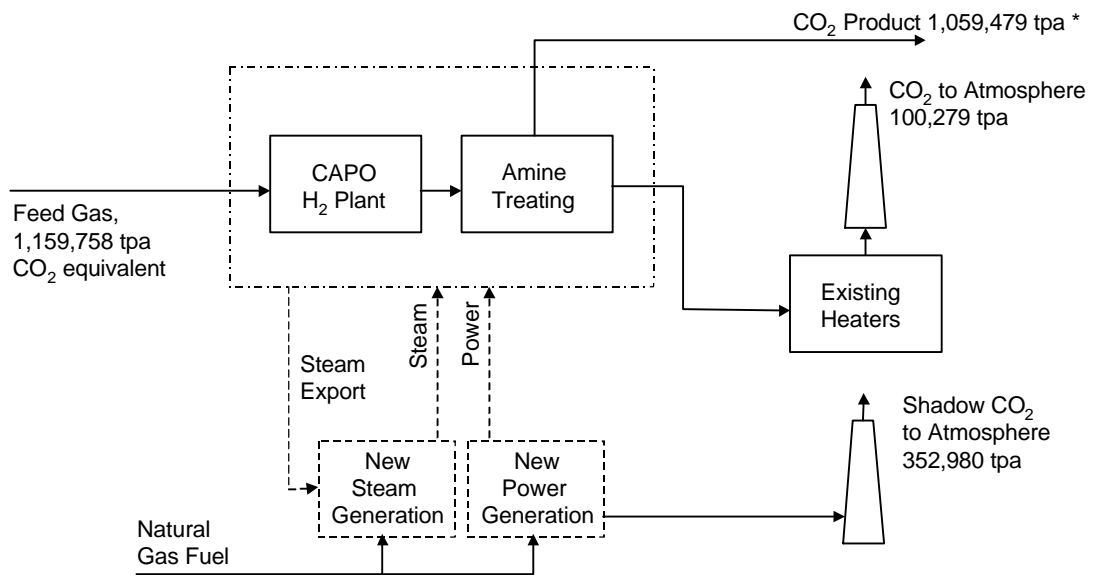
* An additional 115,660 tpa CO_2 product from the existing refinery H_2 plant is also captured as product

Option 8 differs from options 6 and 7 in that it does not have a fired preheater. In option 8 feed preheat is provided by process heat integration with hot reactor products. Heat integration is limited as reactor products at temperatures above 450 °C cannot be used due to the potential of metal dusting corrosion. See attachment 7 for further discussion of the metal dusting phenomena.

The CO_2 emissions in this option are from the refinery furnaces where the flue gases resulting from the hydrogen rich gas combustion are not amine treated. The CAPO hydrogen rich gas leaving the reactor effluent amine treatment, which removes 99.95% of the CO_2, contains a small quantity of carbon based gases.

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This option yields an overall CO$_2$ reduction of 91% before the shadow CO$_2$ is accounted for (see Attachment I). The shadow CO$_2$ lowers this reduction to 57%. The CO$_2$ input into the refinery is 22% more than the base case.

This option gives the fourth best overall CO$_2$ reduction and together with option 4 yields the best capital cost per % CO$_2$ reduction of all the hydrogen rich gas options. However, unlike option 4 (which gives the best overall CO$_2$ reduction) this option does not include a large natural gas fired heater which is not considered appropriate in the context of CO$_2$ abatement. It was therefore decided to develop option 8 in task 3.

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4.0   TASK 3 DISCUSSION

4.1   OVERVIEW

In this task a more detailed assessment was made of the 3 options carried forward from task 2 ie:

- Amine Treatment (Option 1)

- Oxygen rich burning (Option 2)

- Hydrogen Rich Fuel – CAPO – without fired heater preheat (Option 8)

The activities undertaken during this task and associated references are summarised in Table 4-1 below:

Table 4-1: Activities

Task 3 activity	Report reference Section No
Carbon Flow Block Flow Diagrams	3
Process Flow Diagrams / Process Descriptions	5
Capital Costs	6
Levelised cost of CO$_2$ capture	8
Discounted Cash Flow (DCF) analyses of options	7
Specific energy consumption figures	8
Major results summary	Attachment 5

4.2   CO$_2$ CAPTURED / AVOIDED

CO$_2$ captured and CO$_2$ avoided are two ways of expressing the success of CO$_2$ removal from the atmosphere for each of the options examined by this study. These expressions are used in the performance and economic summary in section 8 of this report.

CO$_2$ captured is the amount of CO$_2$ product that is collected for compression.

CO$_2$ avoided is the difference between the base case refinery CO$_2$ emission and the modified plant CO$_2$ emission. CO$_2$ avoided may be expressed with and without shadow CO$_2$.

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The diagrams below illustrate the two quantities.

Fig 4.1 CO₂ Capture Neglecting Shadow
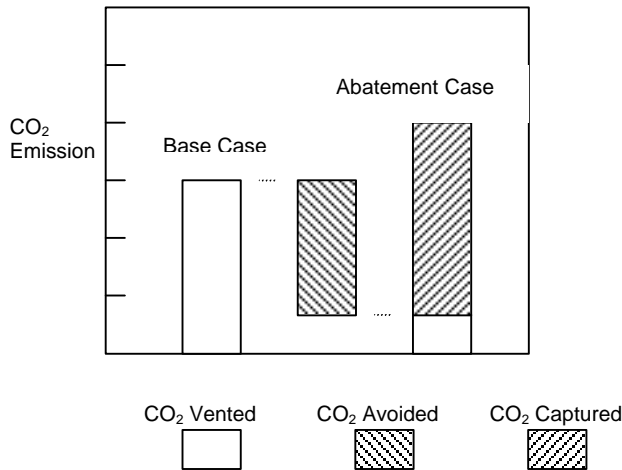
Fig 4.2 CO₂ Capture Including Shadow
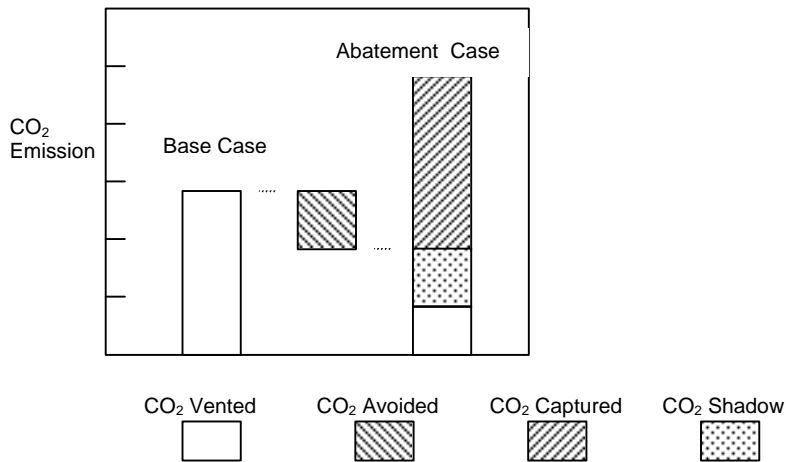
CO_2 captured and CO_2 avoided quantities have been calculated using the vented, captured and shadow quantities shown on the block flow diagrams for each option.

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Table 4-2: CO$_2$ Avoided and CO$_2$ Captured

	Option 1 Amine	Option 2 O2 Rich	Option 8 CAPO
Base case refinery emission tpa	1,062,780	1,062,780	1,062,780
From block flow diagram			
CO$_2$ compressed tpa**	1,013,942	939,915	1,175,139
CO$_2$ to atmosphere tpa	47,356	43,372	100,279
CO$_2$ shadow tpa	330,200	338,429	352,980
CO$_2$ Captured tpa	1,013,942	939,915	1,175,139
CO$_2$ Avoided tpa	1,013,942	1,019,408	962,501
CO$_2$ Avoided including shadow tpa	685,224	680,979	609,521
CO$_2$ Captured t/ plant lifetime *	22,813,695	21,148,088	26,440,628
CO$_2$ Avoided t/ plant lifetime *	22,813,695	22,936,680	21,656,273
CO$_2$ Avoided including shadow t/ plant lifetime *	15,417,540	15,322,028	13,714,223

* Plant lifetime is considered to be 25 years at 90% productivity.

** Including 115,660 tpa CO$_2$ product from existing H$_2$ plant which is also captured as product

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5.0   PROCESS DESCRIPTIONS

Please refer to the Process Flow Diagrams and Material Balances located at the end of this section.

5.1   AMINE TREATMENT (OPTION 1)

5.1.1 Overview

Flue gases from the refinery heaters are treated by scrubbing with amine solvent in an absorber. One amine absorber unit is required per stack to avoid having to redistribute scrubbed flue gases from one absorber back to several stacks.

Flue gas streams are named 1A, 1B, 1C etc. in the material balance corresponding to stacks A to G. Each of stacks A to G will have an associated absorber unit. Indicative equipment sizes were obtained through vendor enquiry. Small stacks, such as C, E & G, will require absorber columns of up to 2.5 m diameter x 26 m tall, whereas, the largest stacks (A & B) may require absorber columns of up to 6.1 m diameter x 27.0 m tall.

Rich amine solvent from the absorbers is routed to a single central regeneration facility, lean amine solvent from this facility is returned to the absorbers. The indicative size for the regenerator column is 6.1m diameter x 19 m tall.

5.1.2 Absorber / Regenerator

The flue gas enters the plant at slightly above atmospheric pressure and near the heater flue gas temperature (180°C for forced draft, 350°C for natural draft). A guard bed is used to remove trace amounts of SO$_x$ (<15ppm) from the gas prior to contact with the alkanolamine solution. It is important to remove the SOx as if it contacts with amine heat stable salts can form resulting in loss of solvent.

The flue gas is cooled in a Direct Contact Water Cooler to 46 – 62 $^{\circ}$C. Cooled flue gas is compressed to overcome the pressure drop in the Absorber and to enable the CO$_2$ free flue gas to reach atmosphere via the stack.   In the Absorber the CO$_2$ is removed by counter current contact with the alkanolamine solvent. Flue gas is vented from the top of the absorber and is directed to the existing stack. The solvent loaded with CO$_2$ termed "rich" is pumped from the bottom of the absorber through the Lean/ Rich Exchanger to the Solvent Regenerator.

In the regenerator the CO$_2$ is stripped from the solvent using steam generated in the column reboiler. The stripped "lean" solvent leaves the bottom of the Stripper for return to the Absorber. The solvent quality is maintained by passing a slip stream of lean solvent through a Carbon Filtration step prior to return to the Absorber. The steam and CO$_2$ leave as the Stripper overhead.

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The Condenser removes the water from the CO_2 and in the Reflux drum the CO_2 and water are separated. The water is returned to the column and the high purity CO_2 is available for capture at 1.4 bar and 44 °C.

This process is very similar in operation to other alkanolamine (MEA, DEA, DIPA, MDEA) units commonly found on refineries. However, due to the presence of Oxygen in the flue gas, there are special requirements for use of inhibitors in the solvent in order to minimise corrosion. The inhibitor and the maintenance of the amine solution are both critical to smooth and economical operation of this type of unit.

5.2   OXYGEN RICH BURNING (OPTION 2)

5.2.1   Overview

In this option the process heaters are converted to burn fuel gas in an oxygen rich environment yielding a reduced flue gas flow. Oxygen is supplied to the heaters from an Air Separation Unit (ASU).

The combined flue gas from the heaters is routed to a CO_2 recovery unit where CO_2 is recovered by ammonia refrigerated distillation. The small volume of tail gas from this process is vented to atmosphere via an adjacent existing stack.

5.2.2   Process Description

Oxygen for the heaters is produced at 20°C and 3 barg by a conventional ASU at about 95mol% purity. At each heater the oxygen line is tied into the recycled flue gas ducting, the combined flow is then routed into the burner flue gas plenum. The flue gas is recycled to cool the combustion temperature of the fuel gas and oxygen to acceptable limits.

The flue gas leaves the heater at about 370 °C, the oxygen rich combustion ensures this gas contains less than 0.2 mol% nitrogen (35.6%CO_2, 4%O_2, 60.1%H_2O). About two thirds of this flow is recycled to the burners via a new recycle flue gas blower, the remainder combines with the net flue gas from the other heaters and flows to the CO_2 Recovery Plant.

The flue gas is cooled in a direct water cooler, compressed to around 18 barg and dried in a molecular sieve system. High drying performance is required to avoid ice formation which would foul the cryogenic exchangers.

The gas is then chilled and refrigerated using an ammonia circuit before distillation. The column condenser is also ammonia refrigerated and is linked to the reboiler in a heat pump arrangement.

The top product containing the flue gas oxygen, nitrogen and argon and about 5 t/h of carbon dioxide is routed to atmosphere. The carbon dioxide leaves as the bottom product at 16 barg which is then pumped to the required 110 barg battery limit pressure.

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5.3   HYDROGEN RICH BURNING (SMR) (OPTIONS 3, 4 & 5)

5.3.1   Overview

For these options a new Steam Methane Reformer (SMR) is added to the generic refinery to supplement / replace the process heater fuel gas with a hydrogen rich gas. The displaced fuel gas is used as feed to the SMR made up with natural gas to meet the hydrogen product requirement. The fuel gas distribution system and heater burners are modified to accommodate the lower molecular weight fuel.

The three SMR options considered are described in Table 5-1 below:

Table 5-1

Option No	Option Title	Configuration / Basis
3	Hydrogen Rich Fuel – SMR - 85% of refinery fuel gas as H_2 ex SMR	SMR +Pressure Swing Adsorption (PSA) PSA tail gas routed to SMR heater Remainder of fuel for SMR heater made up with natural gas
4	Hydrogen Rich Fuel – SMR – 100% of refinery fuel gas as H_2 ex SMR	As above
5	Hydrogen Rich Fuel – SMR – 100% of refinery fuel gas as H_2 ex SMR	SMR w/o PSA Fuel for reformer heater entirely provided by hydrogen product

Option 4 gives superior CO_2 abatement (see section 3) therefore the process description and material balance data given here (5.3.2) is for this arrangement. The material balance differences with option 3 are shown on the PFD's. The key differences with option 5 are discussed in section 5.3.3.

5.3.2   SMR plus PSA (Option 3 & 4)

The total refinery gas make is compressed to 20 barg, combined with natural gas and fed to the reformer furnace. The feed is preheated to 371°C in a furnace convective coil before entering the desulphurisation system where any sulphur and hydrogen sulphide are removed. The desulphurised feed is then mixed with steam produced on the unit and further heated in the convective section to 540 ^{o}C. The mixed feed then enters the radiant section of the reformer and flows down through catalyst filled tubes, where it undergoes the reforming reactions to produce H_2 and CO.

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The reformer effluent leaves the furnace at 780°C, is cooled to 350°C to produce HP steam and is then routed to the Shift Conversion reactors (High and Low temperature shift). The shift conversion reaction converts most of the CO to CO$_2$ over a bed of catalyst. The gas leaves the shift converters at 228 ^{o}C and is used to preheat boiler feedwater where it is cooled to around 110°C.

Most (99.5%) of the CO$_2$ in the gas leaving the BFW heater is recovered by amine absorption. The CO$_2$ product is combined with CO$_2$ from the reformer furnace amine unit and compressed to the required 110 barg battery limit pressure.

The product gas leaving the CO$_2$ absorber is routed to a Pressure Swing Adsorption (PSA) unit where 85% of the hydrogen is recovered in a 99.9+% pure product stream which is routed to the refinery heaters. The tail gas (75mol% hydrogen and 19% methane) is routed to the reformer furnace to provide about 50% of the reformer duty requirement. The remaining fuel is made up from natural gas.

The flue gas from the reformer furnace contains about 7 mol% CO$_2$ and requires amine treatment. The treated flue gas is routed to atmosphere and the CO$_2$ product from the regenerator is combined with the reformer effluent CO$_2$ and compressed to product pressure (see above).

5.3.3 SMR without PSA (Option 5)

The flowscheme for this alternative differs from the SMR plus PSA scheme described above in the following areas :

- The hydrogen rich gas (95mol% hydrogen) leaving the reformer effluent amine absorber is routed directly to the refinery heaters and SMR heater.

- All of the reformer duty is met by the hydrogen rich gas.

- The reformer heater flue gas is not amine treated.

- About twice as much natural gas is required in the reformer feed, however the overall natural gas consumption increase is about 15% (cf alternative 2, feed plus fuel)

5.4 HYDROGEN RICH BURNING (CAPO)(OPTIONS 6,7 & 8)

5.4.1 Overview

For these options a new Catalytic Air Partial Oxidation (CAPO) unit is added to the generic refinery to supplement / replace the process heater fuel gas with a hydrogen rich gas. The displaced fuel gas is used as feed to the CAPO made up with natural gas to meet the hydrogen product requirement. The fuel gas distribution system and heater burners are modified to accommodate the lower molecular weight fuel.

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The three CAPO options considered are described in Table 5-2 below:

Table 5-2

Option No	Option Title	Configuration / Basis
6	Hydrogen Rich Fuel – CAPO - 85% of refinery fuel gas as H_2 ex SMR	Feed preheated in fired heater. Hydrogen product used to fire heater.
7	Hydrogen Rich Fuel – CAPO - 100% of refinery fuel gas as H_2 ex SMR	As above
8	Hydrogen Rich Fuel – CAPO - 100% of refinery fuel gas as H_2 ex SMR	Reactor effluent (<450°C) used to preheat feed, no fired heater.

Option 8 was selected for further development in task 3 (see section 3) therefore the process description and material balance data given here (5.4.2) is for this arrangement. The key differences with options 7 and 8 are discussed in section 5.4.3.

5.4.2 CAPO (Reactor effluent for preheat, no feed heater)(Option 8)

All the refinery fuel gas and natural gas are compressed to about 35 barg, preheated to 380°C and mixed with a small recycle hydrogen stream prior to desulphurisation where any sulphur and hydrogen sulphide are removed.

The desulphurised feed is mixed with steam and further preheated against reactor effluent prior to entering the CAPO reactor. Air for the partial oxidation is compressed to about 34.5 barg, preheated against reactor effluent and is also routed to the reactor. The reactor inlet temperature is limited by the 450°C maximum temperature of the reactor effluent which provides the preheat. This limit is set to avoid metal dusting which occurs in the feed / effluent exchanger above this temperature. (Refer to attachment 7 for further explanation of metal dusting.)

The reactor contains a nickel based catalyst over which the feed gas, steam and air react to form a hydrogen rich reformed gas. The product gas leaves the reactor at 850 ^{o}C and is cooled to raise HP steam and preheat air and feed gas before being routed to two High Temperature (HT) Shift Reactors and a LT shift reactor. The shift reaction converts most of the CO to CO_2 over a bed of catalyst. MP steam is raised between the two HT reactors and LP steam is raised between the second HT and LT reactor and downstream of the LT reactor.

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Condensed water from the shifted product gas leaving the final LP steam generator is removed in a gas separator vessel. The product gas from this vessel is routed for CO$_2$ removal in the amine unit where 99.5% of the CO$_2$ is recovered. The CO$_2$ product is compressed to the required 110 barg battery limit pressure.

Product gas leaving the CO$_2$ recovery is 50% hydrogen and 48% nitrogen, the remainder made up of methane with small quantities of argon, CO and CO$_2$. A small recycle stream is routed to the feed desulphuriser section and the remainder is fed to the refinery heaters.

5.4.3 CAPO with feed heater (Options 6and 7)

The flowscheme for these options differ from option 8 in the following areas :

- Feed is preheated to 650°C in a fired heater fuelled by hydrogen product from the unit

- About 20% more hydrogen is required in the product gas to fuel the heater (option 7 cf option 8)

- The overall hydrocarbon feed is increased to provide this product

- Less air is required per mol of feed for partial oxidation as the reactor inlet temperature is higher

- The overall molar throughput in the reactor and downstream equipment is about the same (option 7 cf option 8)

RELUX CONDENSER

CW

STRIPPER

3

CO₂ COMPRESSION

CO₂ Product 110 bar

Note 3

CARBON FILTER

2A

Flue gas to atmosphere via existing stack

REFLUX DRUM

ABSORBER

LP Steam

LEAN SOLVENT COOLER

REBOILER

WATER WASH COOLER

Heater Flue Gas

DIRECT CONTACT COOLER

Note 2

BLOWER

LEAN/RICH EXCHANGER

1A

RECLAIMER

SOx GUARD BED

DCC WATER COOLER

SOLVENT STORAGE

DCC CIRCULATION PUMP

WASH COOLER PUMP

RICH SOLVENT PUMP

LEAN SOLVENT PUMP

SOLVENT MAKE-UP PUMP

REFLUX PUMP

RICH SOLVENT FROM OTHER ABSORBERS	B
	C
	D
	E
	F
	G

	B	LEAN SOLVENT TO OTHER ABSORBERS
	C	
	D	
	E	
	F	
	G	

Notes

1. Facilities comprise of seven individual amine absorbers (one for each of stacks A to G) and a single amine regeneration unit common to all.

2. Only one absorber is shown. Heater flue gas stream 1A is diverted from stack A. Similar streams 1B, 1C, 1D, 1E, 1F and 1G are from stacks B to G respectively to other absorbtion facilities.

3. Treated flue gas stream 2A returns to stack A. Similar streams 2B, 2C, 2D, 2E, 2F and 2G return to stacks B to G respectively.

PROCESS FLOW DIAGRAM

PROJECT: IEA

CO₂ ABATEMENT IN REFINERIES: FIRED HEATERS

CASE: OPTION 1

AMINE TREATMENT

REV.:   2

O₂ to other heaters

Flue gas from other heaters

KO Drum

Direct Water Cooler

CWI

CWO

Compressor

KO Drum

Dryer

③

⑥

ASU

Oxygen ⑤

NH3

NH3

⑧

Tail Gas (note 1)

Heater

Distillation

TYPICAL PROCESS HEATER

④

NH3

Natural Gas ⑨

②

Refinery Gas ⑩

①

RECYCLE FLUE GAS BLOWER

⑦

CO₂ PRODUCT PUMPING

CO₂ Product 110 bar

Fuel to Other Heaters
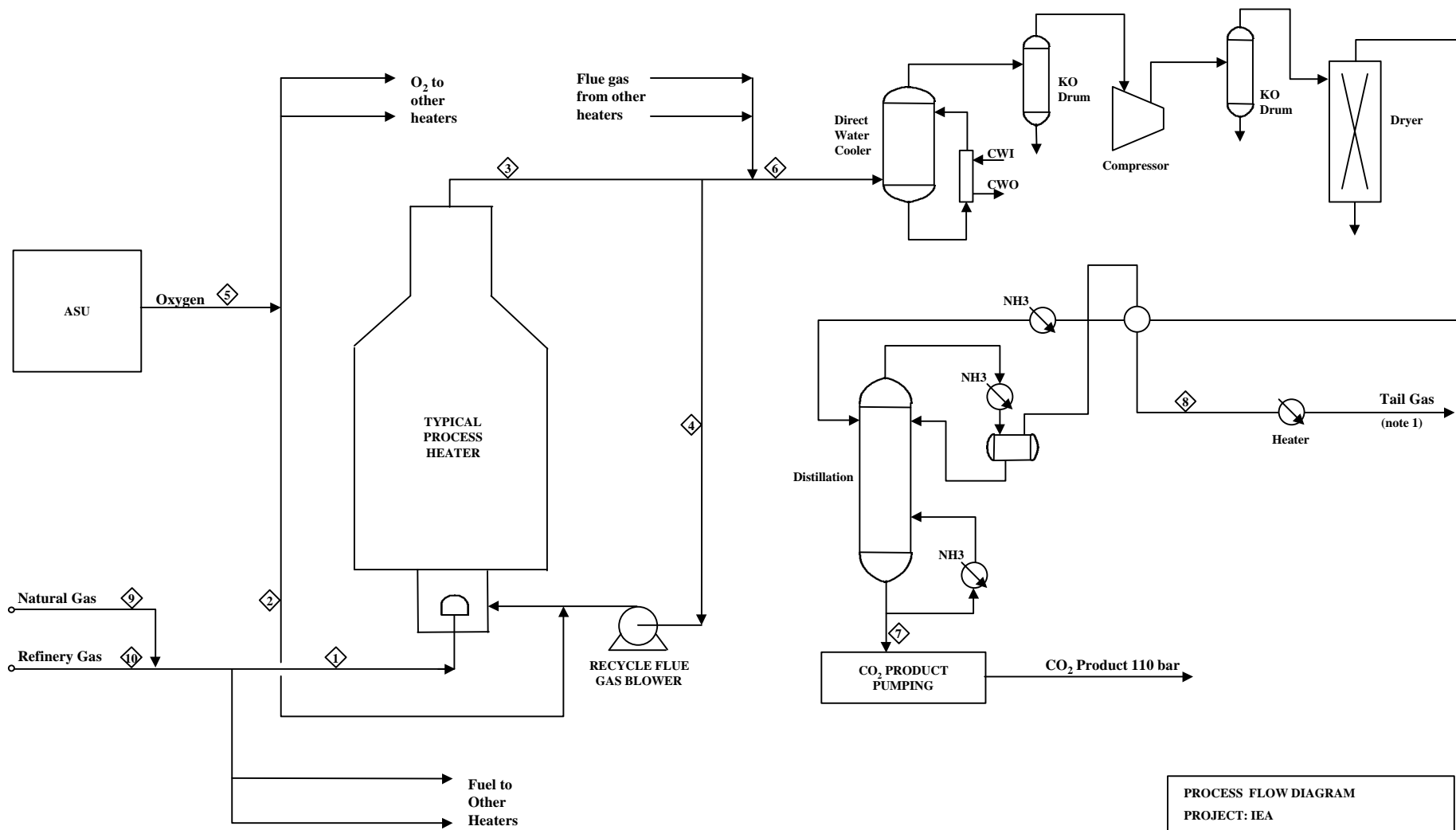
Notes

1. Process optimisation may result in the design of an oxygen recycle from the tail gas. The present scheme assumes 100% of stream 8 as tail gas vented to atmosphere.

PROCESS FLOW DIAGRAM

PROJECT: IEA

CO₂ ABATEMENT IN REFINERIES: FIRED HEATERS

CASE: OPTION 2

OXYGEN RICH GAS

REV.:   3

GAS COMPRESSION

Refinery Gas

Natural Gas ◇2

DE-SULPHURISATION

◇1

◇8

Flue to Atmosphere

C0₂ REMOVAL BY AMINE

CO₂ Product

◇7

CO₂ COMPRESSION

CO SHIFT CONVERTER (HT & LT SHIFT)

CO 2 REMOVAL BY AMINE

◇9

PSA

◇6

Hydrogen Rich Product

Natural Gas ◇3

HEAT RECOVERY

HEAT RECOVERY

◇10

BFW

◇5

Steam Export

PSA Tail Gas Recycle
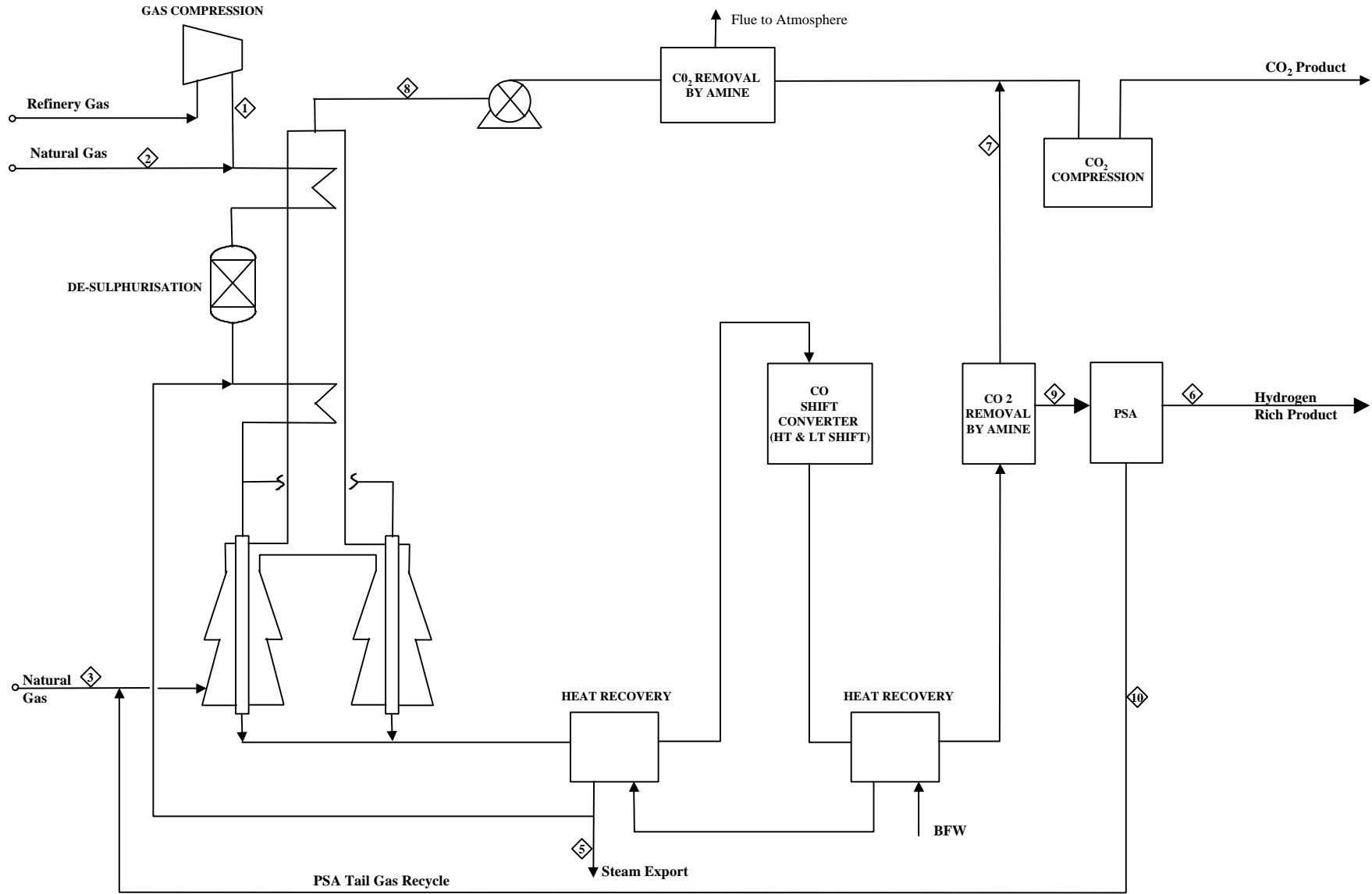
PROCESS FLOW DIAGRAM

PROJECT: IEA

CO₂ ABATEMENT IN REFINERIES: FIRED HEATERS

CASE: OPTIONS 3 & 4

STEAM/METHANE REFORMING

REV.: 2

GAS COMPRESSION

Flue to Atmosphere

Refinery Gas

⟨1⟩

⟨8⟩

Natural Gas

⟨2⟩

DE-SULPHURISATION

CO SHIFT CONVERTER (HT & LT SHIFT)

CO 2 REMOVAL BY AMINE

CO₂ COMPRESSION

CO₂ Product

⟨7⟩

⟨6⟩ Hydrogen Rich Product

⟨3⟩

HEAT RECOVERY

HEAT RECOVERY

Process Steam

BFW

⟨5⟩ Steam Export

Hydrogen Rich Fuel
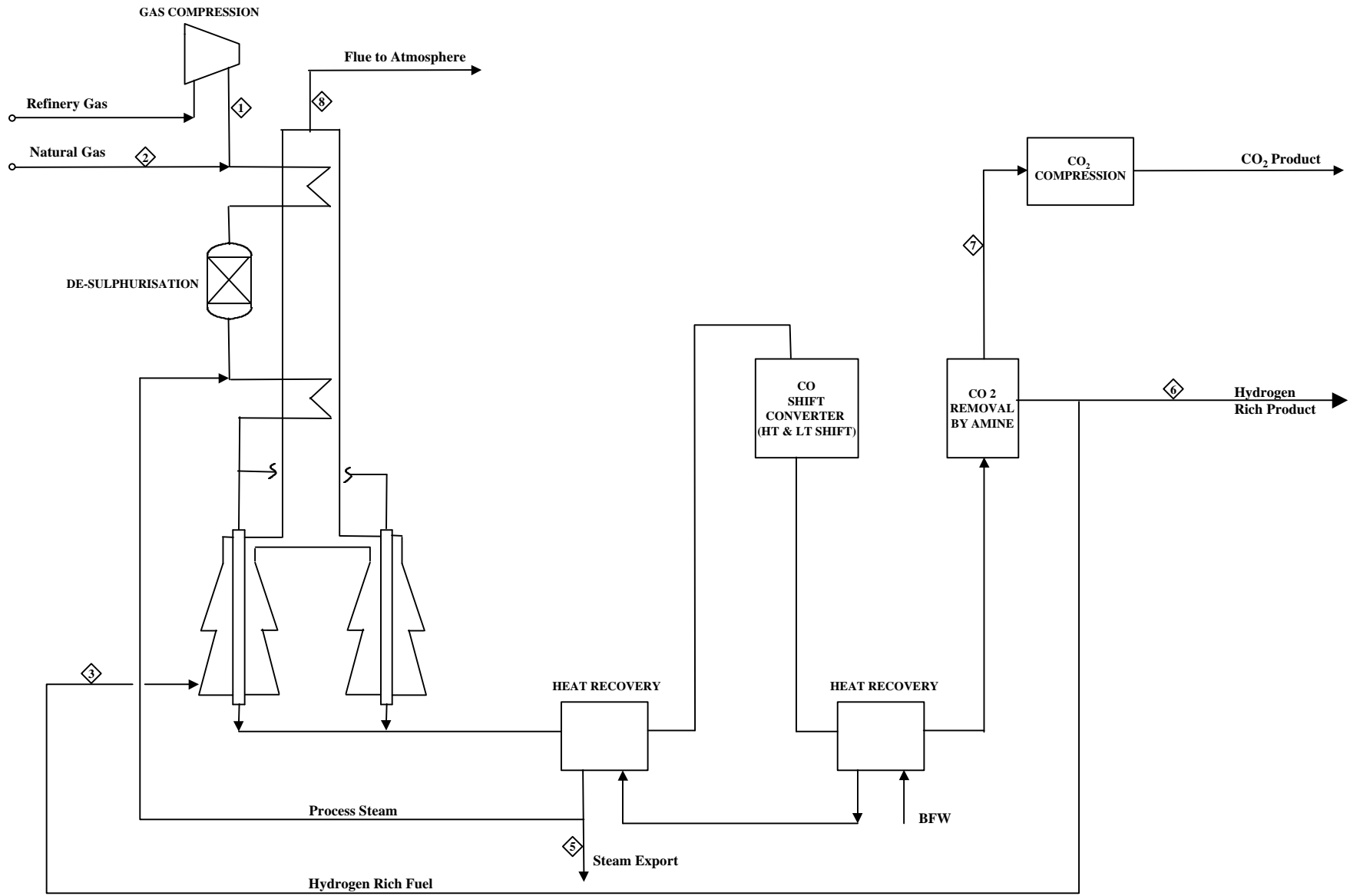
PROCESS FLOW DIAGRAM
PROJECT: IEA
CO₂ ABATEMENT IN
REFINERIES: FIRED HEATERS
CASE: OPTION 5
STEAM/METHANE REFORMING
REV.: 1

Refinery Gas Feed ◇2

Natural Gas Make-up ◇1

FEED GAS
KO DRUM

FEED GAS
COMPRESSOR

HYDROGEN
COMPRESSOR

Hydrogen for Desulphurisation

CO₂
COMPRESSION

CO₂ Product

HT CO
CONVERTER

HT CO
CONVERTER

LT CO
CONVERTER

◇9

Air

◇3

PRE-
REFORMER

DESULPHURISER

DESULPHURISER

◇4

MP Steam

CO₂
REMOVAL

MEMBRANE

◇8

Hydrogen
Rich Fuel Gas

◇10

Membrane
Reject

BFW

BFW

BFW

BFW

GAS
SEPERATOR

◇11 Condensate

◇5

HP Steam

◇6

MP Steam

◇7

LP Steam

FEED HEATER

CATALYTIC
PARTIAL
OXIDATION
REACTOR

PROCESS FLOW DIAGRAM

PROJECT: IEA

CO2 ABATEMENT IN
REFINERIES: FIRED HEATERS

CASE: OPTIONS 6 & 7

HYDROGEN RICH GAS (CAPO)

REV.:   0

Refinery Gas Feed ◇2

FEED GAS KO DRUM

Natural Gas Make-up ◇1

FEED GAS COMPRESSOR

HYDROGEN COMPRESSOR

Hydrogen for Desulphurisation

CO₂ COMPRESSION → **CO₂ Product**

HT CO CONVERTER

HT CO CONVERTER

LT CO CONVERTER

◇9

DESULPHURISER

Air ◇3

DESULPHURISER

BFW **BFW** **BFW** **BFW**

CO₂ REMOVAL → **Hydrogen Rich Fuel Gas** ◇8

◇4
MP Steam

◇5
HP Steam

◇6
MP Steam

◇7
LP Steam

GAS SEPERATOR

◇10 **Condensate**

FEED HEATER

CATALYTIC PARTIAL OXIDATION REACTOR

PROCESS FLOW DIAGRAM
PROJECT: IEA
  CO2 ABATEMENT IN
  REFINERIES: FIRED HEATERS
CASE: OPTION 8
  H2 RICH GAS (CAPO) NO HEATER
REV.:   0

STREAM NUMBER		1A	1B	1C	1D	1E	1F	1G				
DESCRIPTION		Stack A Flue Gas	Stack B Flue Gas	Stack C Flue Gas	Stack D Flue Gas	Stack E Flue Gas	Stack F Flue Gas	Stack G Flue Gas				
COMPONENTS		kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h				
Hydrogen		0.00	0.00	0.00	0.00	0.00	0.00	0.00				
Carbon Monoxide		0.00	0.00	0.00	0.00	0.00	0.00	0.00				
Carbon Dioxide		747.87	772.33	79.72	444.29	149.44	390.52	25.04				
Nitrogen		6134.19	6579.32	696.57	3644.16	1305.79	3412.37	205.42				
Oxygen		252.09	323.81	37.96	149.76	71.16	185.96	8.44				
Methane		0.00	0.00	0.00	0.00	0.00	0.00	0.00				
Ethane		0.00	0.00	0.00	0.00	0.00	0.00	0.00				
Propane		0.00	0.00	0.00	0.00	0.00	0.00	0.00				
Butane		0.00	0.00	0.00	0.00	0.00	0.00	0.00				
Pentane		0.00	0.00	0.00	0.00	0.00	0.00	0.00				
Hexane		0.00	0.00	0.00	0.00	0.00	0.00	0.00				
Argon		0.00	0.00	0.00	0.00	0.00	0.00	0.00				
Water		1268.85	1307.55	134.76	753.79	252.62	660.16	42.49				
FLOWRATE TOTAL	kmol/h	8403.0	8983.0	949.0	4992.0	1779.0	4649.0	281.4				
MASS FLOWRATE TOTAL	kg/h	235617	252121	26646	139990	49962	130567	7887				
LIQUID FLOWRATE TOTAL	Am³/h											
VAPOUR FLOWRATE TOTAL	Am³/h											
MOLE FRACTION VAP	frac											
TEMPERATURE	°C	180	283	350	180	350	350	180				
PRESSURE	bara	1.0	1.0	1.0	1.0	1.0	1.0	1.0				
MOLECULAR WEIGHT	kg/kmol	28.0	28.1	28.1	28.0	28.1	28.1	28.0				

NOTES	FW FOSTER WHEELER ENERGY LIMITED	MASS BALANCE PROCESS SPECIFICATION
	CUSTOMERS NAME: IEA	
	LOCATION: GENERIC REFINERY	PROJECT No: 1-17-11815
	SERVICE: AMINE - OPTION 1	UNIT No: OPTION 1
	REV 0	DOCUMENT CAT. -
	DATE 08/05/00	DOCUMENT No. of
	ORIG. BY RAC	SHEET
	APP. BY	DOCUMENT SEQUENCE No.

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STREAM NUMBER		2A	2B	2C	2D	2E	2F	2G	3			
DESCRIPTION		Flue Gas to Atm	Flue Gas to Atm	Flue Gas to Atm	Flue Gas to Atm	Flue Gas to Atm	Flue Gas to Atm	Flue Gas to Atm	CO2 Product			
COMPONENTS		kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h			
Hydrogen		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00			
Carbon Monoxide		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00			
Carbon Dioxide		37.39	38.62	3.99	22.21	7.47	19.53	1.25	2478.74			
Nitrogen		6134.19	6579.32	696.57	3644.16	1305.79	3412.37	205.42	0.00			
Oxygen		252.09	323.81	37.96	149.76	71.16	185.96	8.44	0.00			
Methane		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00			
Ethane		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00			
Propane		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00			
Butane		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00			
Pentane		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00			
Hexane		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00			
Argon		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00			
Water		418.72	431.49	44.47	248.75	83.36	217.85	14.02	0.00			
FLOWRATE TOTAL	kmol/h	6842.4	7373.2	783.0	4064.9	1467.8	3835.7	229.1	2478.7			
MASS FLOWRATE TOTAL	kg/h	189047	204061	21688	112324	40668	106278	6327	109089			
LIQUID FLOWRATE TOTAL	Am³/h											
VAPOUR FLOWRATE TOTAL	Am³/h											
MOLE FRACTION VAP	frac											
TEMPERATURE	°C	37	37	37	37	37	37	37	49			
PRESSURE	bara	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.4			
MOLECULAR WEIGHT	kg/kmol	27.6	27.7	27.7	27.6	27.7	27.7	27.6	44.0			

NOTES			
	FW **FOSTER WHEELER ENERGY LIMITED**		**MASS BALANCE PROCESS SPECIFICATION**
	CUSTOMERS NAME: IEA		
	LOCATION: GENERIC REFINERY		PROJECT No: 1-17-11815
	SERVICE: AMINE - OPTION 1		UNIT No: OPTION 1
	REV 0		DOCUMENT CAT. -
	DATE 08/05/00		DOCUMENT No. of
	ORIG. BY RAC		SHEET
	APP. BY		DOCUMENT SEQUENCE No.

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STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	
DESCRIPTION		Heater Fuel	Oxygen to Heater	Heater Flue	Recycle Flue	Oxygen Total	Total Flue to Recover	CO2	Tail Gas	Natural Gas	Refinery Gas	
COMPONENTS		kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	
Hydrogen		59.85	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	190.36	
Carbon Monoxide		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Carbon Dioxide		2.53	0.00	2320.48	1567.57	0.00	2394.17	2274.46	119.71	2.59	5.47	
Nitrogen		1.11	30.82	98.39	66.46	98.01	101.51	0.00	101.51	3.62	0.00	
Oxygen		0.00	1464.13	267.72	180.86	4655.71	276.22	0.00	276.22	0.00	0.00	
Methane		307.08	0.00	0.00	0.00	0.00	0.00	0.00	0.00	490.69	485.74	
Ethane		38.16	0.00	0.00	0.00	0.00	0.00	0.00	0.00	19.67	101.74	
Propane		34.41	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.04	108.31	
Butane		55.39	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	176.13	
Pentane		7.25	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	22.97	
Hexane		1.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	3.28	
Argon		0.00	46.24	142.56	96.31	147.02	147.02	0.00	147.02	0.00	0.00	
Water		0.00	0.00	3863.90	2610.21	0.00	3986.60	0.00	0.00	0.00	0.00	
FLOWRATE TOTAL	kmol/h	506.8	1541.2	6693.0	4521.4	4900.7	6905.5	2274.5	644.5	517.6	1094.0	
MASS FLOWRATE TOTAL	kg/h	11682	49562	188744	127504	157601	194738	100099	94639	8722	28433	
LIQUID FLOWRATE TOTAL	Am³/h											
VAPOUR FLOWRATE TOTAL	Am³/h											
MOLE FRACTION VAP	frac											
TEMPERATURE	°C	20	20	371	371	20	371	-23	-21	20	20	
PRESSURE	bara	3.0	3.0	1.0	1.0	3.0	1.0	17.0	15.0	3.0	3.0	
MOLECULAR WEIGHT	kg/kmol	23.1	32.2	28.2	28.2	32.2	28.2	44.0	146.8	16.9	26.0	

NOTES			FW FOSTER WHEELER ENERGY LIMITED		MASS BALANCE PROCESS SPECIFICATION	
			CUSTOMERS NAME: IEA			
			LOCATION: GENERIC REFINERY		PROJECT No: 1-17-11815	
			SERVICE: OXYGEN RICH GAS - OPTION 2		UNIT No: OPTION 2	
			REV	0	1			DOCUMENT CAT. -
			DATE	09/05/00	01/06/00			DOCUMENT No. of
			ORIG. BY	RAC	RAC			SHEET
			APP. BY					DOCUMENT SEQUENCE No.

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STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	
DESCRIPTION		RFG Feed	NG Feed	NG Fuel	Demin Make-up	Steam Prod	Hydrogen Prod	CO2 Prod	Flue Gas	PSA Feed	Tail Gas	
COMPONENTS		kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h
Hydrogen		190.36	0.00	0.00	0.00	0.00	5065.00	0.00	0.00	5964.71	894.71	
Carbon Monoxide		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	18.06	18.06	
Carbon Dioxide		5.47	0.24	2.38	0.00	0.00	0.00	1665.14	738.29	8.37	8.37	
Nitrogen		0.00	0.33	3.33	0.00	0.00	0.00	0.00	7605.56	0.33	0.33	
Oxygen		0.00	0.00	0.00	0.00	0.00	0.00	0.00	180.38	0.00	0.00	
Methane		485.74	44.91	451.30	0.00	0.00	0.00	0.00	0.00	215.89	215.89	
Ethane		101.74	1.80	18.09	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Propane		108.31	0.09	0.95	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Butane		176.13	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Pentane		22.97	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Hexane		3.28	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Argon		0.00	0.00	0.00	0.00	0.00	0.00	0.00	90.96	0.00	0.00	
Water		0.00	0.00	0.00	5154.31	1774.31	0.00	0.00	2415.57	30.05	30.05	
FLOWRATE TOTAL	kmol/h	1094.0	47.4	476.1	5154.3	1774.3	5065.0	1665.1	11030.8	6237.4	1167.4	
MASS FLOWRATE TOTAL	kg/h	28430	799	8024	92856	31965	10210	73282	298473	16912.51	6692.001	
LIQUID FLOWRATE TOTAL	Am³/h	0.00	0.00	0.00	93	0.00	0.00	0.00	0.00	0.00	0.00	
VAPOUR FLOWRATE TOTAL	Am³/h	1445	57	11389	0.00	2202	15397	36785	385278	16610.64	22796.49	
MOLE FRACTION VAP	frac	1.00	1.00	1.00	0.00	1.00	1.00	1.00	1.00	1.00	1.00	
TEMPERATURE	°C	65.00	25.00	25.00	30.00	400.00	40.02	100.00	160.00	35.00	32.00	
PRESSURE	bara	21.00	21.00	3.00	6.00	42.45	8.61	1.40	1.03	9.65	1.30	
MOLECULAR WEIGHT	kg/kmol	25.99	16.86	16.86	18.02	18.02	2.02	44.01	27.06	2.71	5.73	

NOTES	FOSTER WHEELER ENERGY LIMITED	MASS BALANCE PROCESS SPECIFICATION
1. Mass balance based on firing reformer with natural gas.		

CUSTOMERS NAME:	IEA				
LOCATION:	GENERIC REFINERY			PROJECT No:	1-17-11815
SERVICE:	SMR 85% CASE - OPTION 3			UNIT No:	OPTION 3
REV	0	1		DOCUMENT CAT. -	
DATE	10/04/00	19/05/00		DOCUMENT No.	of
ORIG. BY	NI	NI		SHEET	
APP. BY				DOCUMENT SEQUENCE No.	

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STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	
DESCRIPTION		RFG Feed	NG Feed	NG Fuel	Demin Make-up	Steam Prod	Hydrogen Prod	CO2 Prod	Flue Gas	PSA Feed	Tail Gas	
COMPONENTS		kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h
Hydrogen		190.36	0.00	0.00	0.00	0.00	8096.03	0.00	0.00	9525.61	1429.58	
Carbon Monoxide		0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	28.52	28.52	
Carbon Dioxide		5.47	5.32	3.69	0.00	0.00	0.00	2559.91	1168.37	12.86	12.86	
Nitrogen		0.00	7.45	5.16	0.00	0.00	0.00	0.00	12076.98	7.45	7.45	
Oxygen		0.00	0.00	0.00	0.00	0.00	0.00	0.00	286.44	0.00	0.00	
Methane		485.74	1008.64	699.47	0.00	0.00	0.00	0.00	0.00	358.22	358.22	
Ethane		101.74	40.43	28.04	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Propane		108.31	2.13	1.48	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Butane		176.13	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Pentane		22.97	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Hexane		3.28	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Argon		0.00	0.00	0.00	0.00	0.00	0.00	0.00	144.36	0.00	0.00	
Water		0.00	0.00	0.00	7954.83	2737.83	0.00	0.00	3855.32	64.36	64.36	
FLOWRATE TOTAL	kmol/h	1094.0	1064.0	737.8	7954.8	2737.8	8096.0	2559.9	17531.5	9997.0	1901.0	
MASS FLOWRATE TOTAL	kg/h	28430	17934	12437	143309	49323	16321	112661	474052	27392	11159	
LIQUID FLOWRATE TOTAL	Am³/h	0.00	0.00	0.00	144	0.00	0.00	0.00	0.00	0.00	0.00	
VAPOUR FLOWRATE TOTAL	Am³/h	1445	1271	17651	0.00	3397	24588	56552	612333	26622	37121	
MOLE FRACTION VAP	frac	1.00	1.00	1.00	0.00	1.00	1.00	1.00	1.00	1.00	1.00	
TEMPERATURE	°C	65.00	25.00	25.00	30.00	400.00	39.73	100.00	160.00	35.00	32.00	
PRESSURE	bara	21.00	21.00	3.00	6.00	42.45	8.61	1.40	1.03	9.65	1.3	
MOLECULAR WEIGHT	kg/kmol	25.99	16.86	16.86	18.02	18.02	2.02	44.01	27.04	2.74	5.87	

NOTES

1. Mass balance based on firing reformer with natural gas.

FW

FOSTER WHEELER ENERGY LIMITED

MASS BALANCE PROCESS SPECIFICATION

CUSTOMERS NAME:	IEA				
LOCATION:	GENERIC REFINERY			PROJECT No:	1-17-11815
SERVICE:	SMR 100% CASE - OPTION 4			UNIT No:	OPTION 4
REV	0	1	2	DOCUMENT CAT. -	
DATE	18/05/00	19/05/00	21/06/00	DOCUMENT No.	of
ORIG. BY	NI	NI	RAC	SHEET	
APP. BY				DOCUMENT SEQUENCE No.	

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STREAM NUMBER		1	2	3	4	5	6	7	8			
DESCRIPTION		RFG Feed	NG Feed	Hydrogen Fuel	Demin Make-up	Steam Prod	Hydrogen Prod	CO2 Prod	Flue Gas			
COMPONENTS		kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h
Hydrogen		190.36	0.00	5928.38	0.00	0.00	7159.99	0.00	0.00			
Carbon Monoxide		0.00	0.00	17.63	0.00	0.00	21.30	0.00	0.00			
Carbon Dioxide		5.47	10.40	7.86	0.00	0.00	9.49	3453.84	258.10			
Nitrogen		0.00	14.56	6.59	0.00	0.00	7.96	0.00	14054.32			
Oxygen		0.00	0.00	0.00	0.00	0.00	0.00	0.00	342.12			
Methane		485.74	1971.54	226.61	0.00	0.00	273.69	0.00	0.00			
Ethane		101.74	79.03	0.00	0.00	0.00	0.00	0.00	0.00			
Propane		108.31	4.16	0.00	0.00	0.00	0.00	0.00	0.00			
Butane		176.13	0.00	0.00	0.00	0.00	0.00	0.00	0.00			
Pentane		22.97	0.00	0.00	0.00	0.00	0.00	0.00	0.00			
Hexane		3.28	0.00	0.00	0.00	0.00	0.00	0.00	0.00			
Argon		0.00	0.00	0.00	0.00	0.00	0.00	0.00	168.09			
Water		0.00	0.00	29.96	9291.65	2275.65	36.18	0.00	6593.27			
FLOWRATE TOTAL	kmol/h	1094.0	2079.7	6217.0	9291.7	2275.7	7508.6	3453.8	21415.9			
MASS FLOWRATE TOTAL	kg/h	28430	35054	17150	167392	40997	20713	152003	541511			
LIQUID FLOWRATE TOTAL	Am³/h	0.00	0.00	0.00	168.07	0.00	0.00	0.00	0.00			
VAPOUR FLOWRATE TOTAL	Am³/h	1445	2484	16578	0.00	2824	20760	76300	747785			
MOLE FRACTION VAP	frac	1.00	1.00	1.00	0.00	1.00	1.00	1.00	1.00			
TEMPERATURE	°C	65.00	25.00	34.70	30.00	400.00	34.71	100.00	160.00			
PRESSURE	bara	20.00	20.00	9.65	6.00	42.45	9.31	1.40	1.03			
MOLECULAR WEIGHT	kg/kmol	25.99	16.86	2.76	18.02	18.02	2.76	44.01	25.29			

NOTES			
1. Alternative SMR case based on firing reformer with H2 rich gas. Process does not include a PSA or flue gas amine treating.	FW **FOSTER WHEELER ENERGY LIMITED**		**MASS BALANCE PROCESS SPECIFICATION**

CUSTOMERS NAME:	IEA	
LOCATION:	GENERIC REFINERY	PROJECT No: 1-17-11815
SERVICE:	SMR - OPTION 5	UNIT No: OPTION 5
REV	0	DOCUMENT CAT. -
DATE	10/05/00	DOCUMENT No. of
ORIG. BY	NI	SHEET
APP. BY		DOCUMENT SEQUENCE No.

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STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	
DESCRIPTION		NG Feed	Refinery Feed	Air Fuel	Steam Feed	HP Steam Raised	MP Steam Raised	LP Steam Raised	Product Fuel Gas	CO2 Product	Conden-sate	
COMPONENTS		kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	
Hydrogen		0.00	190.36	0.00	0.00	0.00	0.00	0.00	5441.66	17.01	0.00	
Carbon Monoxide		0.00	0.00	0.00	0.00	0.00	0.00	0.00	30.42	0.00	0.00	
Carbon Dioxide		2.59	5.47	0.00	0.00	0.00	0.00	0.00	9.87	1963.56	1.74	
Nitrogen		3.63	0.00	3410.76	0.00	0.00	0.00	0.00	3415.78	6.46	0.17	
Oxygen		0.00	0.00	915.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Methane		491.44	485.74	0.00	0.00	0.00	0.00	0.00	283.88	1.94	0.00	
Ethane		19.70	101.74	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Propane		1.04	108.31	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Butane		0.00	176.13	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Pentane		0.00	22.97	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Hexane		0.00	3.28	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Argon		0.00	0.00	40.95	0.00	0.00	0.00	0.00	41.48	0.00	0.00	
Water		0.00	0.00	36.55	3992.85	2472.92	953.67	1025.49	0.00	158.67	1736.30	
FLOWRATE TOTAL	kmol/h	518	1094	4403	3993	2473	954	1025	9223	2148	1738	
MASS FLOWRATE TOTAL	kg/h	8740	28430	127139	71940	44512	17166	18459	114199	86663	31361	
LIQUID FLOWRATE TOTAL	Am³/h											
VAPOUR FLOWRATE TOTAL	Am³/h											
MOLE FRACTION VAP	frac											
TEMPERATURE	°C	15	15	15	390	400	300	200	30	49	70	
PRESSURE	bara	21.0	3.0	1.0	35.0	40.0	21.0	4.0	4.0	1.4	26.5	
MOLECULAR WEIGHT	kg/kmol	16.9	26.0	28.9	18.0	18.0	18.0	18.0	12.4	40.4	18.0	

NOTES		FW FOSTER WHEELER ENERGY LIMITED		MASS BALANCE PROCESS SPECIFICATION	
		CUSTOMERS NAME: IEA			
		LOCATION: GENERIC REFINERY		PROJECT No: 1-17-11815	
		SERVICE: CAPO 85% CASE - OPTION 6		UNIT No: OPTION 6	
		REV	0	1			DOCUMENT CAT. -	
		DATE	02/05/00	14/06/00			DOCUMENT No. of	
		ORIG. BY	RAC	RAC			SHEET	
		APP. BY					DOCUMENT SEQUENCE No.	

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STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	
DESCRIPTION		NG Feed	Refinery Feed	Air Fuel	Steam Feed	HP Steam Raised	MP Steam Raised	LP Steam Raised	Product Fuel Gas	CO2 Product	Conden- sate	
COMPONENTS		kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	
Hydrogen		0.00	190.36	0.00	0.00	0.00	0.00	0.00	8688.53	27.29	0.00	
Carbon Monoxide		0.00	0.00	0.00	0.00	0.00	0.00	0.00	47.33	0.00	0.00	
Carbon Dioxide		8.68	5.47	0.00	0.00	0.00	0.00	0.00	15.41	3066.60	2.51	
Nitrogen		12.16	0.00	5557.22	0.00	0.00	0.00	0.00	5564.56	10.44	0.28	
Oxygen		0.00	0.00	1490.82	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Methane		1646.58	485.74	0.00	0.00	0.00	0.00	0.00	417.12	2.69	0.00	
Ethane		66.00	101.74	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Propane		3.47	108.31	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Butane		0.00	176.13	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Pentane		0.00	22.97	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Hexane		0.00	3.28	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Argon		0.00	0.00	66.72	0.00	0.00	0.00	0.00	66.56	0.00	0.00	
Water		0.00	0.00	59.55	6190.20	5875.90	1514.46	1628.01	0.00	253.66	2789.64	
FLOWRATE TOTAL	kmol/h	1737	1094	7174	6190	5876	1514	1628	14800	3361	2792	
MASS FLOWRATE TOTAL	kg/h	29277	28430	207081	111492	105767	27261	29303	184823	135352	50377	
LIQUID FLOWRATE TOTAL	Am³/h											
VAPOUR FLOWRATE TOTAL	Am³/h											
MOLE FRACTION VAP	frac											
TEMPERATURE	°C	15	15	15	390	400	300	200	30	49	70	
PRESSURE	bara	21.0	3.0	1.0	35.0	40.0	21.0	4.0	4.0	1.4	26.5	
MOLECULAR WEIGHT	kg/kmol	16.9	26.0	28.9	18.0	18.0	18.0	18.0	12.5	40.3	18.0	

NOTES

 FOSTER WHEELER ENERGY LIMITED

MASS BALANCE PROCESS SPECIFICATION

CUSTOMERS NAME:	IEA					
LOCATION:	GENERIC REFINERY				PROJECT No:	1-17-11815
SERVICE:	CAPO 100% CASE - OPTION 7				UNIT No:	OPTION 7
REV	0	1			DOCUMENT CAT. -	
DATE	02/05/00	14/06/00			DOCUMENT No.	of
ORIG. BY	RAC	RAC			SHEET	
APP. BY					DOCUMENT SEQUENCE No.	

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STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	
DESCRIPTION		NG Feed	Refinery Feed	Air Fuel	Steam Feed	HP Steam Raised	MP Steam Raised	LP Steam Raised	Product Fuel Gas	CO2 Product	Conden-sate	
COMPONENTS		kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	kmol/h	
Hydrogen		0.00	190.36	0.00	0.00	0.00	0.00	0.00	7311.25	22.87	0.00	
Carbon Monoxide		0.00	0.00	0.00	0.00	0.00	0.00	0.00	32.21	0.00	0.00	
Carbon Dioxide		6.86	5.47	0.00	0.00	0.00	0.00	0.00	14.69	2924.15	2.94	
Nitrogen		9.60	0.00	6978.37	0.00	0.00	0.00	0.00	6978.85	13.21	0.33	
Oxygen		0.00	0.00	1872.07	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Methane		1300.66	485.74	0.00	0.00	0.00	0.00	0.00	229.90	1.61	0.00	
Ethane		52.14	101.74	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Propane		2.74	108.31	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Butane		0.00	176.13	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Pentane		0.00	22.97	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Hexane		0.00	3.28	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	
Argon		0.00	0.00	83.78	0.00	0.00	0.00	0.00	83.47	0.00	0.00	
Water		0.00	0.00	74.77	5588.10	5267.70	1359.90	1896.30	0.00	251.89	3262.83	
FLOWRATE TOTAL	kmol/h	1372	1094	9009	5588	5268	1360	1896	14650	3214	3266	
MASS FLOWRATE TOTAL	kg/h	23122	28430	260010	100710	94860	24507	34166	218886	129134	58914	
LIQUID FLOWRATE TOTAL	Am³/h											
VAPOUR FLOWRATE TOTAL	Am³/h											
MOLE FRACTION VAP	frac											
TEMPERATURE	°C	15	15	15	390	400	300	200	30	49	70	
PRESSURE	bara	21.0	3.0	1.0	35.0	40.0	21.0	4.0	4.0	1.4	26.5	
MOLECULAR WEIGHT	kg/kmol	16.9	26.0	28.9	18.0	18.0	18.0	18.0	14.9	40.2	18.0	

NOTES	FW FOSTER WHEELER ENERGY LIMITED	MASS BALANCE PROCESS SPECIFICATION	
	CUSTOMERS NAME: IEA		
	LOCATION: GENERIC REFINERY	PROJECT No: 1-17-11815	
	SERVICE: CAPO NO PREHEAT - OPTION 8	UNIT No: OPTION 8	
	REV 0	DOCUMENT CAT. -	
	DATE 12/06/00	DOCUMENT No. of	
	ORIG. BY RAC	SHEET	
	APP. BY	DOCUMENT SEQUENCE No.	

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CO₂ ABATEMENT IN OIL
REFINERIES: FIRED
HEATERS

FOSTER WHEELER ENERGY LTD

AUGUST 2000

6.0   COST ESTIMATES

A cost estimate summary for the eight options considered in task 2 is shown overleaf in tables 6-1 and 6-2. The accuracy of these estimates is considered to be +/- 30%.

It should be noted that the estimate does not allow for outside battery limit costs to support the incremental utility requirements of the options. This will depend on the spare capacity of those utility systems in the specific location considered and the extent of modifications required.

Client: IEA

Project: CO2 ABATEMENT IN OIL REFINERIES : FIRED HEATERS (Study Report)

FW Contract No: 1-17-11815

Table 6-1: Estimate Summary & Comparison: Total Cost Breakdown into Materials, Construction & Engineering

Option No	Description	Materials US $	Construction US $	Engineering US $	Total US $
1	Amine Treatment	76 246 075	43 605 541	26 308 384	**146 160 000**
2	Oxygen Rich Burning	74 166 000	37 824 000	24 583 000	**136 573 000**
3	Hydrogen Rich fuel - SMR - 85% of RFG as H2 ex SMR (w/PSA)	91 951 000	48 068 000	30 736 000	**170 755 000**
4	Hydrogen Rich fuel - SMR - 100% of RFG as H2 ex SMR (w/PSA)	127 402 000	61 311 980	41 425 020	**230 139 000**
5	Hydrogen Rich fuel - SMR - 100% of RFG as H2 ex SMR, H2 as Reformer fuel	149 028 550	73 159 470	48 772 980	**270 961 000**
6	Hydrogen Rich Gas - CAPO - 85% of RFG as H2 ex CAPO (w/heater)	85 982 000	35 952 000	26 766 000	**148 700 000**
7	Hydrogen Rich Gas - CAPO - 100% of RFG as H2 ex CAPO (w/heater)	121 208 000	48 436 000	37 239 000	**206 883 000**
8	Hydrogen Rich Gas - CAPO - 100% of RFG as H2 ex CAPO without fired heater preheat	119 529 280	46 596 160	36 466 560	**202 592 000**

Client: IEA
Project: CO2 ABATEMENT IN OIL REFINERIES : FIRED HEATERS (Study Report)
FW Contract No: 1-17-11815

Table 6-2: Estimate Summary & Comparison: Total Cost Breakdown by Plant Area

Option No	Description	Plant Area	Installed Cost of Plant Area US$
1	Amine Treatment	Duct work between absorbers & stacks	7 227 000
		Modification to existing stacks	1 006 000
		Absorbers	90 287 000
		Regenerator	17 703 000
		CO2 Compression	29 937 000
		Total installed cost	146 160 000
2	Oxygen Rich Burning	Air Separation Unit	65 594 000
		Heater Modifications	14 237 000
		CO2 Recovery	38 258 000
		CO2 Compression	18 484 000
		Total installed cost	136 573 000
3	Hydrogen Rich fuel - SMR - 85% of RFG as H2 ex SMR (w/PSA)	New fuel distribution system to heaters	1 140 000
		Modified burners	65 000
		New SMR (including PSA)	79 808 000
		CO2 Removal & Compression	78 595 000
		Feed Gas Compression	11 147 000
		Total installed cost	170 755 000
4	Hydrogen Rich fuel - SMR - 100% of RFG as H2 ex SMR (w/PSA)	New fuel distribution system to heaters	1 140 000
		Modified burners	65 000
		New SMR (including PSA)	115 782 000
		CO2 Removal & Compression	102 005 000
		Feed Gas Compression	11 147 000
		Total installed cost	230 139 000
5	Hydrogen Rich fuel - SMR - 100% of RFG as H2 ex SMR. H2 as Reformer fuel	New fuel distribution system to heaters	1 140 000
		Modified burners	65 000
		New SMR	175 680 000
		CO2 Removal & Compression	82 929 000
		Feed Gas Compression	11 147 000
		Total installed cost	270 961 000

Client: IEA
Project: CO2 ABATEMENT IN OIL REFINERIES : FIRED HEATERS (Study Report)
FW Contract No: 1-17-11815

Table 6-2: Estimate Summary & Comparison: Total Cost Breakdown by Plant Area

Option No	Description	Plant Area	Installed Cost of Plant Area US$
6	Hydrogen Rich Gas - CAPO - 85% of RFG as H2 ex CAPO (w/heater)	New fuel distribution system to heaters	1 140 000
		Modified burners	65 000
		New CAPO , CO2 Removal Unit & Compression	134 946 000
		Feed Gas Compression	6 702 000
		Air Compression	4 285 000
		Feed Heater	1 562 000
		Total installed cost	148 700 000
7	Hydrogen Rich Gas - CAPO - 100% of RFG as H2 ex CAPO (w/heater)	New fuel distribution system to heaters	1 140 000
		Modified burners	65 000
		New CAPO , CO2 Removal Unit & Compression	189 569 000
		Feed Gas Compression	8 125 000
		Air Compression	5 910 000
		Feed Heater	2 074 000
		Total installed cost	206 883 000
8	Hydrogen Rich Gas - CAPO - 100% of RFG as H2 ex CAPO without fired heater preheat	New fuel distribution system to heaters	1 140 000
		Modified burners	65 000
		New CAPO , CO2 Removal Unit & Compression	187 560 000
		Feed Gas Compression	7 255 000
		Air Compression	6 572 000
		Total installed cost	202 592 000

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**CO$_2$ ABATEMENT IN OIL
REFINERIES: FIRED
HEATERS**

FOSTER WHEELER ENERGY LTD

AUGUST 2000

7.0   ECONOMIC EVALUATION

7.1   DCF BASIS

The following basis was used in accordance with IEA Technical Specification: "Technical and Financial Assessment Criteria" Appendix II, Rev B : Nov 99.

Criteria	Basis
Design and Construction Period	2 years for CO$_2$ capture plant and chemical plants. Costs split 40% year 1 and 60% year 2.
Plant Life	25 years
Load Factor	For all plants fuelled by natural gas and plants solely processing gases 90% of rated capacity is used.
Cost of Debt	All capital requirements are treated as debt at the same discount rate used to derive capital charges
Capital Charges; Inflation	Discounted cash flow is expressed at a discount rate of 10%. All annual expenditure occurs at the end of the year. No allowance is made for inflation of fuel, labour or other costs.
Currency	US $
Commissioning and Working Capital	Commissioning period is 3 months. Commissioning cost is included in fixed capital cost. Working capital allows for 15 day storage at rated capacity of raw materials, products and consumables.
Decommissioning	Decommissioning cost assumed zero.
Location	5% of the installed plant cost will be assumed to cover land purchase, surveys, and general site preparation.
Taxation and Insurance	Tax at 1% of the installed plant cost. Taxation on profits not included. Insurance as 1% pa of the installed plant cost.
Fees	2% of installed plant cost for process/ patent fees, agents, consultants, legal and planning fees. Contractors fees included in installed plant cost.
Contingency	Included in plant fixed capital cost.
Maintenance	2% pa for plants handling gases and liquids and services plants
Labour	Maintenance labour included in above. Operating labour works 1960 hour/yr in 4 shift pattern. Supervision is 20% of operating labour direct costs Administration and overheads is a further 60 % of operating labour costs. Operators per shift are:- Amine Option 1 1.5 O$_2$ Rich Option 2 1.5 CAPO Option 8 2

The following data was alternative or additional to that given in IEA Technical Specification: "Technical and Financial Assessment Criteria" Appendix II, Rev B : Nov 99.

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**CO$_2$ ABATEMENT IN OIL
REFINERIES: FIRED
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FOSTER WHEELER ENERGY LTD

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- Product Costs – As discussed in meeting NOM03 4/5/00 the CO$_2$ product has no value.
- Labour Costs – Operator salary $30,500 pa based on previous study.

7.2   UTILITY COSTS

As agreed in correspondence with IEA

- Power $/kWh 0.05
- Natural Gas $/Gj 2
- HP Steam $/tonne 11
- MP Steam $/tonne 8
- LP Steam $/tonne 7

Other costs developed from historic data are:

- Demin. Water $/tonne 1.6
- Cooling Water $/m3 0.032

7.3   OPERATING COSTS

Operating costs are made up of the components listed below. The value of these components was calculated according to the notes in section 7.1.

- Labour
- Maintenance
- Insurance
- Administration

Catalyst and Chemical costs are also included in the operating costs. These were determined on a case by case basis. The following sums were included.

- Amine $MM 0.873 per annum
- O$_2$ Rich $MM 0.037 per annum
- CAPO $MM 2.302 per annum

The largest constituents of the operating costs are the utility running costs. These were calculated from power, steam and water consumptions for each option. Any utility production (such as steam generation) was considered as a credit offsetting other utility expenditure.

7.4   DCF TABULATION

Tabulations showing discounted cash flow for the 3 selected options are presented overleaf. Discussion of results is made in section 8 of this report.

IEA STUDY
CO2 ABATEMENT IN REFINERIES: FIRED HEATERS

DISCOUNTED CASH FLOW MODEL
CASE: **OPTION 1 - AMINE TREATING OF FLUE GASES**
Fixed Capital Cost MM$ -146.2
Fees MM$ -2.9
Land MM$ -7.3

CAPEX $MM: -156.4

YEAR	-1	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25
DCF Summary																											
Operating Factor	0	0	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0
Capital expenditure $MM	-62.574	-93.86																									
Delta Working Capital $MM	0.000	0.000	-0.036	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.036
operating Costs $MM	0.000	0.000	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	0.000
Feed Costs $MM	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Product Revenue $MM	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Net cash flow $MM	-62.574	-93.86	-28.107	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	-28.071	0.036
cummulative cash flow $MM	-62.574	-156.43	-184.54	-212.61	-240.68	-268.75	-296.82	-324.89	-352.97	-381.04	-409.11	-437.18	-465.25	-493.32	-521.39	-549.46	-577.53	-605.60	-633.67	-661.75	-689.82	-717.89	-745.96	-774.0	-802.1	-830.2	-830.13
Discounted cash flow factor	0.909	1.000	1.100	1.210	1.331	1.464	1.611	1.772	1.949	2.144	2.358	2.594	2.853	3.138	3.452	3.797	4.177	4.595	5.054	5.560	6.116	6.727	7.400	8.140	8.954	9.850	10.835
DCF Cost $MM	-68.83	-93.86	-25.552	-23.199	-21.090	-19.173	-17.430	-15.845	-14.405	-13.095	-11.905	-10.823	-9.839	-8.944	-8.131	-7.392	-6.720	-6.109	-5.554	-5.049	-4.590	-4.173	-3.793	-3.448	-3.135	-2.850	0.003

Net Present Value $MM   -414.93

IEA STUDY
CO2 ABATEMENT IN REFINERIES: FIRED HEATERS

DISCOUNTED CASH FLOW MODEL
CASE: **OPTION 2 - OXYGEN RICH BURNING**
Fixed Capital Cost MM$ -136.6
Fees MM$ -2.7
Land MM$ -6.8

CAPEX $MM: -146.2

YEAR	-1	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25
DCF Summary																											
Operating Factor	0	0	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0
Capital expenditure $MM	-58.465	-87.70																									
Delta Working Capital $MM	0.000	0.000	-0.002	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.002
operating Costs $MM	0.000	0.000	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	-32.595	0.000
Feed Costs $MM	0.000	0.000	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	2.573	0.000
Product Revenue $MM	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Net cash flow $MM	-58.465	-87.70	-30.023	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	-30.021	0.002
cummulative cash flow $MM	-58.465	-146.16	-176.18	-206.21	-236.23	-266.25	-296.27	-326.29	-356.31	-386.33	-416.36	-446.38	-476.40	-506.42	-536.44	-566.46	-596.48	-626.50	-656.53	-686.55	-716.57	-746.59	-776.61	-806.6	-836.7	-866.7	-866.67
Discounted cash flow factor	0.909	1.000	1.100	1.210	1.331	1.464	1.611	1.772	1.949	2.144	2.358	2.594	2.853	3.138	3.452	3.797	4.177	4.595	5.054	5.560	6.116	6.727	7.400	8.140	8.954	9.850	10.835
DCF Cost $MM	-64.31	-87.70	-27.294	-24.811	-22.555	-20.505	-18.641	-16.946	-15.406	-14.005	-12.732	-11.575	-10.522	-9.566	-8.696	-7.906	-7.187	-6.534	-5.940	-5.400	-4.909	-4.462	-4.057	-3.688	-3.353	-3.048	0.000

Net Present Value $MM   -421.74

IEA STUDY
CO2 ABATEMENT IN REFINERIES: FIRED HEATERS

DISCOUNTED CASH FLOW MODEL
CASE:	OPTION 8 - CAPO WITHOUT FIRED PREHEAT
Fixed Capital Cost MM$	-202.6
Fees MM$	-4.1
Land MM$	-10.1

CAPEX $MM: -216.8

YEAR	-1	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25
DCF Summary																											
Operating Factor	0	0	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0.9	0
Capital expenditure $MM	-86.713	-130.07																									
Delta Working Capital $MM	0.000	0.000	-0.095	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.095
operating Costs $MM	0.000	0.000	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	-31.483	0.000
Feed Costs $MM	0.000	0.000	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	-7.800	0.000
Product Revenue $MM	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Net cash flow $MM	-86.713	-130.07	-39.378	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	-39.284	0.095
cummulative cash flow $MM	-86.713	-216.78	-256.16	-295.44	-334.73	-374.01	-413.29	-452.58	-491.86	-531.15	-570.43	-609.71	-649.00	-688.28	-727.56	-766.85	-806.13	-845.42	-884.70	-923.98	-963.27	-1002.6	-1041.8	-1081.1	-1120.4	-1159.7	-1159.6
Discounted cash flow factor	0.909	1.000	1.100	1.210	1.331	1.464	1.611	1.772	1.949	2.144	2.358	2.594	2.853	3.138	3.452	3.797	4.177	4.595	5.054	5.560	6.116	6.727	7.400	8.140	8.954	9.850	10.835
DCF Cost $MM	-95.38	-130.07	-35.798	-32.466	-29.514	-26.831	-24.392	-22.175	-20.159	-18.326	-16.660	-15.146	-13.769	-12.517	-11.379	-10.345	-9.404	-8.549	-7.772	-7.066	-6.423	-5.839	-5.308	-4.826	-4.387	-3.988	0.009

Net Present Value $MM	-578.48

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8.0   PERFORMANCE AND ECONOMIC SUMMARY

8.1   DCF SUMMARY AND DISCUSSION

Table 8-1: DCF Summary

	Option 1 Amine Treatment of Flue Gases	Option 2 Oxygen Rich Burning	Option 8 CAPO without Fired Preheat
Total Installed Cost MM$	-146.2	-136.6	-202.6
Operating Cost per year MM$	-28.071	-32.595	-31.483
Feed Cost per year MM$ (natural gas)	0	2.573	-7.8
DCF Net Present Value (NPV) MM$	-414.93	-421.74	-578.48

Comparison of the Net Present Values for the three options shows that Option 1, Amine Treatment is the least loss making option with Option 2 Oxygen Rich Burning a close second place and Option 8 CAPO as the most expensive by a considerable margin.

NPV increases generally in line with the increasing Fixed Capital Costs for each option, however, due to the high operating costs of option 2 over option 1 the NPV of option 2 is greater.

Option 2, Oxygen Rich Burning, includes a natural gas credit in it's feed costs arising from the fact that this option uses less natural gas than the generic refinery base case owing to improved fuel efficiency. However, this is not enough to outweigh the cost of high power requirements associated with the Air Separation Unit.

A large proportion of the operating costs for option 1 are associated with provision of large quantities of LP steam for amine regeneration and cooling water for direct water cooling of flue gases.

In option 8, CAPO, the significant operating costs are associated with power for compression and LP steam for amine regeneration. Feed costs for additional natural gas (quantities in excess of generic refinery basis consumption) are also large. However, this option generates, HP, MP and LP steam for which credit is taken which offsets the high costs for other utilities.

It should be noted that as discussed in section 6 of this report the fixed capital costs for each option that were used to develop the DCF models do not allow for outside battery limit costs to support the incremental utility requirements of the options. For example, considerable new cooling water demand is required in Amine option 1, Option 2 has high power and cooling water demand whilst Option 8 also has a high power demand and may require expansion to

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existing demineralised water facilities for steam generation. No attempt has been made to evaluate these outside battery limit costs as they are refinery specific.

8.2   LEVELISED COST OF CO$_2$ REMOVED

The table below presents the levelised cost of CO$_2$ removal expressed both as CO$_2$ avoided and CO$_2$ captured. For definition of CO$_2$ captured and avoided refer to section 4 of this report.

Table 8-2: Summary of Cost of CO$_2$ removal

	Option 1 Amine Treatment of Flue Gases	Option 2 Oxygen Rich Burning	Option 8 CAPO without Fired Preheat
DCF Net Present Value (NPV) MM$	-414.93	-421.74	-578.48
CO$_2$ Captured, t/ plant lifetime*	22,813,695	21,148,088	26,440,628
CO$_2$ Avoided, t/ plant lifetime*	22,813,695	22,936,680	21,656,273
CO$_2$ Avoided including shadow t/ plant lifetime*	15,417,540	15,322,028	13,714,223
Levelised Cost of CO$_2$ removal NPV/ CO$_2$:			
$/t CO$_2$ captured	-18.2	-19.9	-21.9
$/t CO$_2$ avoided	-18.2	-18.4	-26.7
$/t CO$_2$ avoided including shadow	-26.9	-27.5	-42.2

* Plant lifetime is 25 years

Option 1 amine treating is shown to be most cost effective per unit of CO$_2$ removed whether expressed as captured, avoided or avoided including shadow. Option 2 appears slightly costlier but CO$_2$ avoided is very similar to Option 1. Option 8 is the most expensive in all respects particularly when expressed as CO$_2$ avoided.

8.3   DCF SENSITIVITIES

8.3.1 Utility Cost Sensitivity

An analysis of DCF NPV for Utility price sensitivity was made. Results are in table 8-3 below.

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Table 8-3: DCF Model Sensitivity to Utility Costs

	Option 1 Amine treatment of Flue Gases	Option 2: Oxygen Rich Burning	Option 8 CAPO without Fired Preheat
DCF Net Present Value (NPV) MM$	-414.93	-421.74	-578.48
DCF Net Present Value (NPV) MM$ for Utility cost sensitivities:			
Power +20%	-429.37	-469.6	-613.47
Power -20%	-400.49	-373.89	-543.49
Natural Gas +20%	-414.93	-417.12	-592.5
Natural Gas -20%	-414.93	-426.37	-564.47
Steam +20%	-433.65	-421.74	-581.11
Steam -20%	-396.21	-421.74	-575.86
Demin +20%	-414.93	-421.74	-580.63
Demin -20%	-414.93	-421.74	-576.34
Cooling Water +20%	-422.17	-424.44	-579.46
Cooling Water -20%	-407.69	-419.05	-577.51

The DCF NPV is not sensitive to a utility prices variation of ±20% for any option. The most significant variations are power price reduction and steam price increase where Option 2 becomes a little more favourable than option 1. In all other analysis the order of costs remains unchanged with option 1 least expensive and Option 8 most expensive.

8.3.2 External Power Station Efficiency

As detailed in section 2.3 of this report, this study calculates shadow CO_2 emissions from a power station external to the refinery assuming natural gas firing with the following operating efficiencies.
- Electrical power 35%
- Steam 85%

Sensitivity to power station efficiencies are only shown in the shadow CO_2 numbers. An exercise was carried out to show the cost effect of varying the power and steam generation efficiencies to

- Electrical power 30%, steam 85%
- Electrical power 35%, steam 95%
- Combined heat and power station 55%

The results are given in the table 8-4 below which compares levelised costs for CO_2 avoided including shadow.

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Table 8-4: Cost of CO₂ Removed, Sensitivity to Power Station Efficiency

	Option 1 Amine Treatment of Flue Gases	Option 2 Oxygen Rich Burning	Option 8 CAPO without Fired Preheat
DCF Net Present Value (NPV) MM$	-414.93	-421.74	-578.48
CO₂ Avoided including shadow (study case) t/ plant lifetime	15,417,540	15,322,028	13,714,223
CO₂ Avoided shadow 30% power t/ plant lifetime	15,034,658	14,052,915	12,786,255
CO₂ Avoided shadow 95% steam t/ plant lifetime	15,957,788	15,322,028	13,964,153
CO₂ Avoided shadow 55% CHP t/ plant lifetime	13,453,493	18,090,990	14,443,830
Levelised Cost of CO₂ removal NPV/ CO₂:			
$/t CO₂ avoided including shadow study case	-26.9	-27.5	-42.2
$/t CO₂ avoided including shadow 30% power	-27.6	-30.0	-45.2
$/t CO₂ avoided including shadow 95% steam	-26.0	-27.5	-41.4
$/t CO₂ avoided including shadow 55% CHP	-30.8	-23.3	-40.1

An increase in power production efficiency favours high power consumption options. Option 2 has the largest benefit followed by Option 8 and lastly Option 1. The converse is true for a decrease in power production efficiency.

A decrease in steam generation efficiency penalises high steam consumers such as Options 1 and 8 which use a large quantity of LP steam for amine regeneration. Option 2 is unaffected as there is no steam consumption in the Oxygen Rich Option.

8.3.3 Multiple Streams and Equipment Sparing

Reliability of equipment has not been considered as part of this study and there is no standby equipment, except for centrifugal pumps, filters etc. All CO₂ abatement is performed in single train units although some of these are at the upper capacity limit for which they are normally designed eg ASU (option 1) and Terraced Wall Reformer (Option 5) (also refer to section 3.4.1).

It is noted in that CO₂ abatement in some options is potentially more vulnerable than others to failure of specific equipment. For example, a shutdown of the Terraced Wall Reformer in options 3,4 and 5 will lead to a

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total loss of CO_2 abatement whereas the loss of a single amine absorber in option 1 will only result in a partial reduction in CO_2 capture.

8.3.4 Stack Ratio – Effect on Amine Regeneration

This subject is also discussed in section 3.2.2 of this report. The study assumes each stack has a dedicated amine scrubber but only one central regeneration facility for the refinery is required. The generic refinery has a stack ratio of approximately 2:1. The effect of a higher stack ratio 4:1 was examined, reducing the number of absorbers and also the cost by MM$13.7.

8.4 SPECIFIC ENERGY CONSUMPTION FIGURES

As requested in the IEA Technical Study Specification IEA/CON/99/61 specific energy consumption figures are given below for the task 3 options.

Table 8.5: Specific Energy Consumptions

	Option 1 Amine Treatment of Flue Gases	Option 2 Oxygen Rich Burning	Option 8 CAPO without Fired Preheat
CO_2 compression kWh/t CO_2	108	40	108
Inlet pressure for CO_2 Compression (barg)	0.4	16	0.4
O_2 production kWh/ tonne O_2	-	309	-

8.5 SINGLE PAGE SUMMARY

A single page summary of this section may be found in attachment 5.

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9.0   RECOMMENDATIONS

Areas for future work following on from this study are as follows:

- Examine CO$_2$ abatement options against a range of utility balance / cost scenarios for a specific refinery.

- Examine options for increasing the availability of low grade heat for LP steam production in order to make the options involving amine regeneration more favourable.

- Investigate integration with external power plant with respect to steam and power, heat balance and flue gas treatment.

- Investigate abatement of FCC / RCC coke burn CO$_2$ (excluded from this study scope) on a generic basis.

- Investigate generic refinery process configuration for opportunities for CO$_2$ abatement.

- Establish and agree (with all relevant parties) a comprehensive and standardised formula for deriving shadow CO$_2$ to allow for variations in power station fuels and efficiency.

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ATTACHMENT 1 – TASK 2 SUMMARY

Table overleaf

Attachment 1

Task 2 Summary   Rev 2

*Base Case CO$_2$ for Reduction =   **1 062 780** tpa

Ref: Generic Refinery Study - Heater and CO$_2$ Study Design Basis (Rev 1). Excludes Cat Cracker Coke burn CO$_2$.

Option	Option Description	Capital	CO$_2$ from refinery			Shadow CO$_2$					Emissions Summary					
			Refinery emissions TPA	% Reduction from Base Case*	MM$ / % Reduction	Shadow Produced TPA	Shadow Avoided TPA	Net Shadow TPA	Shadow From LP Steam TPA	Net Shadow excl.LP Steam TPA	Refinery emission + Net Shadow (TPA)	% Reduction from Base Case*	MM$ / % Reduction	Refinery emission + Net Shadow excl.LP Stm (TPA)	% Reduction from Base Case*	MM$ / % Reduction
1	Amine Treatment	146.2	47 356	96	1.5	330 200	0	330 200	228 101	102 099	377 556	64	2.3	149 455	86	1.7
2	Oxygen rich burning	136.6	43 372	96	1.4	338 429	0	338 429	0	338 429	381 801	64	2.1	381 801	64	2.1
3	Hydrogen Rich Fuel - SMR - 85% of RFG as H2 ex SMR	170.8	348 661	67	2.5	234 432	-53 893	180 540	154 777	25 763	529 201	50	3.4	374 424	65	2.6
4	Hydrogen Rich Fuel - SMR - 100% of RFG as H2 ex SMR	230.1	21 166	98	2.3	356 216	-83 157	273 059	238 054	35 005	294 225	72	3.2	56 171	95	2.4
5	Hydrogen Rich Fuel - SMR - 100% of RFG as H2 ex SMR H2 as Reformer Fuel	271.0	204 029	81	3.4	360 043	-69 120	290 922	263 985	26 938	494 951	53	5.1	230 967	78	3.5
6	Hydrogen Rich Fuel - CAPO - 85% of RFG as H2 ex CAPO	148.7	452 726	57	2.6	401 577	-192 980	208 597	185 645	22 952	661 323	38	3.9	475 678	55	2.7
7	Hydrogen Rich Fuel - CAPO - 100% of RFG as H2 ex CAPO	206.9	173 860	84	2.5	630 974	-308 849	322 125	290 032	32 092	495 985	53	3.9	205 952	81	2.6
8	Hydrogen Rich Fuel - CAPO - Without Fired Heater Preheat	202.6	100 279	91	2.2	604 807	-251 827	352 980	262 746	90 234	453 259	57	3.5	190 513	82	2.5

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ATTACHMENT 2 - UTILITY CONSUMPTION SUMMARIES

Table A2-1:Task 3 Options Utility Summaries

Utility	Amine Treatment	Oxygen Rich Burning	Hydrogen Rich Fuel CAPO (w/o fired heater preheat)
Electrical Power (MW)	21.7	71.9	52.5
HP Steam (MW)	0	0	(34.2)
MP steam (MW)	0	0	(20.2)
LP Steam (MW)	117.6	0	108.9
Cooling water (m3/h)	17,000	10,545	2,280

Notes:
1. ( ) indicates quantity produced
2. Electrical power includes additional power for cooling water circulation
3. Utility Breakdowns are given in Table A2-2 overleaf.

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Table A2-2:Task 3 Options Utility Breakdown

	Amine Treatment		Oxygen Rich Burning		Hydrogen Rich Fuel CAPO (w/o fired heater preheat)	
Plant Area / Utility	Absorbers	Regenerator & CO2 Compression	ASU	Remainder of Plant (note 1)	CAPO unit	CO$_2$ removal and compression
Electrical Power (MW)	5.9	15.8	46.9	25	42.2	10.3
HP Steam (MW)	0	0	0	0	(34.2)	
MP steam (MW)	0	0	0	0	(20.2)	
LP Steam (MW)	0	117.6	0	0	(26.7)	135.6
Cooling water (m3/h)	13,560	3,440	4,220	6,325	2,280	included in CAPO unit consumption

Notes
1. CO$_2$ Recovery, compression and flue gas recycle.

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ATTACHMENT 3 - PLOT PLANS

Plot plans in this attachment have been developed from the "Generic European Refinery Layout" provided by IEA in Invitation to Tender Document IEA/CON/99/61.

Drawings are not to scale and hence plot areas are indicative only. Approximate routing of ducting has been shown where applicable. Piping connections are not shown.

It is assumed that the heaters are located close to the existing stacks and therefore flue gas ducting from the heater is shown from the same location as the stack.

The following plot plans are included in this attachment.

Fig A3-1:   Generic Refinery plot plan showing available plot areas
Fig A3-2:   Option 1, Amine, approximate plot areas (stack ratio 2:1)
Fig A3-3:   Option 2, Oxygen Rich Burning, approximate plot areas
Fig A3-4:   Option 8, CAPO, approximate plot areas

FIG. A3-1: GENERIC REFINERY PLOT PLAN SHOWING AVAILABLE PLOT AREAS

FIG. A3-2: OPTION 1, AMINE, APPROXIMATE PLOT AREAS (STACK RATIO 2:1)

STACK G

FCCU

ALKYLATION UNIT

STACK H

CAT REFORMER

STACK D

D

STACK A

MEROX UNIT

A

G

C

CDU

STACK C

DISTILLATE HYDROTREATER

E

R

SWS2

GAS OIL HYDROTREATER

STACK E

SULPHUR UNIT

HYDROGEN UNIT

STACK F

ADIP UNIT

CENTRAL CONTROL BUILDING

F

B

HYDROCRACKER COMPLEX

VDU

STACK B

INDICATIVE SCALE IN METRES

0   100   200

FUEL OIL STORAGE

COOLING TOWER

MAIN PIPERACK LOCATION

REFINERY ROAD

A   ABSORBER PLOT AREA ABSORBERS A TO G

R   REGENERATOR PLOT AREA

STACK I

POWER GENERATION

NEW DUCTING ROUTE. INCLUDES DUCTING FROM HEATER TO ABSOBER AND FROM ABSORBER BACK TO STACK

FIG. A3-3: OPTION 2, OXYGEN RICH BURNING, APPROXIMATE PLOT AREAS

STACK G ●

ALKYLATION UNIT

FCCU

STACK H ●

CAT REFORMER

STACK D ●

CO₂ REMOVAL   AIR SEPARATION UNIT

STACK A ●

MEROX UNIT

CDU

STACK C ●

DISTILLATE HYDROTREATER

SWS2

GAS OIL HYDROTREATER

STACK E ●

SULPHUR UNIT   HYDROGEN UNIT

● STACK F

ADIP UNIT

CENTRAL CONTROL BUILDING

HYDROCRACKER COMPLEX

VDU

STACK B ●

INDICATIVE SCALE IN METRES

| 0 | 100 | 200 |

FUEL OIL STORAGE

COOLING TOWER

░░░░ MAIN PIPERACK LOCATION

▬▬ REFINERY ROAD

▦ NEW PLOT AREA

━━ NEW DUCTING ROUTE FROM CO₂ REMOVAL UNIT TO AVAILABLE EXISTING STACK. DUCTING FROM HEATERS TO CO₂ REMOVAL UNIT NOT SHOWN

● STACK I

POWER GENERATION

FIG. A3-4: OPTION 8, CAPO, APPROXIMATE PLOT AREA

STACK G

FCCU

STACK H

STACK D

ALKYLATION UNIT

CAT REFORMER

STACK A

MEROX UNIT

CDU

DISTILLATE
HYDROTREATER

STACK C

SWS2

GAS OIL HYDROTREATER

STACK E

SULPHUR
UNIT

HYDROGEN
UNIT

STACK F

CAPO
COMPLEX

ADIP
UNIT

CENTRAL
CONTROL
BUILDING

HYDROCRACKER COMPLEX

VDU

STACK B

INDICATIVE SCALE IN METRES

0   100   200

FUEL OIL
STORAGE

COOLING
TOWER

MAIN PIPERACK LOCATION

REFINERY ROAD

NEW PLOT AREA

STACK I

POWER
GENERATION

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AUGUST 2000

ATTACHMENT 4 - ACKNOWLEDGEMENTS

The following companies provided assistance in the development of this study report:

Name	Address	Area of study where assistance provided
GTC Technology Corporation	1001 S.Dairy Ashford, #200 Houston, TX77077 USA	Amine Treatment*
BOC Gases	Engineering centre 30 Priestly Road, Guildford GU2 5YH England	Oxygen Rich burning
Costain Oil, Gas & Process Limited	Costain House, Styal Road, Manchester M22 5WN England	Oxygen Rich burning

*Mitsubishi Heavy Industries were requested to provide information on Amine Treatment but declined.

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**ATTACHMENT 5
SUMMARY TABLES FOR PERFORMANCE AND ECONOMICS**

This appendix gives a single page summary of results given in section 8 of this report.

Table A5-1: DCF Summary

	Option 1 Amine Treatment of Flue Gases	**Option 2 Oxygen Rich Burning**	**Option 8 CAPO without Fired Preheat**
Total Installed Cost MM$	-146.2	-136.6	-202.6
DCF Net Present Value (NPV) MM$	-414.93	-421.74	-578.48

The table below presents the levelised cost of CO_2 removal expressed both as CO_2 avoided and CO_2 captured. For definition of CO_2 captured and avoided refer to section 4 of this report.

Table A5-2: Summary of Cost of CO_2 removal

	Option 1 Amine Treatment of Flue Gases	Option 2 Oxygen Rich Burning	Option 8 CAPO without Fired Preheat
DCF Net Present Value (NPV) MM$	-414.93	-421.74	-578.48
CO_2 Captured, t/ plant lifetime*	22,813,695	21,148,088	26,440,628
CO_2 Avoided, t/ plant lifetime*	22,813,695	22,936,680	21,656,273
CO_2 Avoided including shadow t/ plant lifetime*	15,417,540	15,322,028	13,714,223
Levelised Cost of CO_2 removal NPV/ CO_2:			
$/t CO_2 captured	-18.2	-19.9	-21.9
$/t CO_2 avoided	-18.2	-18.4	-26.7
$/t CO_2 avoided including shadow	-26.9	-27.5	-42.2

* Plant lifetime is 25 years

Option 1 amine treating is shown to be most cost effective per unit of CO_2 removed whether expressed as captured, avoided or avoided including shadow. Option 2 appears slightly costlier but CO_2 avoided is very similar to Option 1. Option 8 is the most expensive in all respects particularly when expressed as CO_2 avoided.

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ATTACHMENT 6 - GLOSSARY OF TERMS
For use in conjunction with Section 1.0

ASU
Air Separation Unit.

Base case CO$_2$ emissions
The CO$_2$ emission from the generic refinery basis before application of abatement techniques.

CAPO
Catalytic Air Partial Oxidation.

CO$_2$ Avoided
The difference between the base case refinery CO$_2$ emission and the modified plant CO$_2$ emission. CO$_2$ avoided may be expressed with and without shadow.

CO$_2$ Captured
The amount of CO$_2$ product that is collected for compression.

DCF Analysis
Discounted cash flow analysis.

Generic Refinery
The base case refinery configuration against which all study options were developed.

Levelised Cost of CO$_2$ Removal
The levelised cost of CO$_2$ removal has been defined here as the NPV divided by the tonnes of CO$_2$ either captured or avoided over the lifetime of the plant.

NPV
Net Present Value results from discounted cash flow (DCF) analysis. It is the sum of the annual DCF costs over the lifetime of the plant.

RFG
Refinery fuel gas.

Shadow CO$_2$
The additional CO$_2$ associated with providing electrical power and steam to the CO$_2$ abatement process. Shadow "produced" results from power and steam consumption in abatement options. Shadow "avoided" is a CO$_2$ credit arising from generation of steam and power within the abatement option. "Net" shadow is the sum of shadow produced and shadow avoided.

SMR
Steam Methane Reformer.

Target CO$_2$ Reduction
A nominal reduction target of 75% of base case CO$_2$ emissions was set as an objective for this study.

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ATTACHMENT 7

**COMPARISON OF CAPO OPTIONS TO PRECOMBUSTION
DECARBONISATION STUDIES**

The purpose of this appendix is to provide a comment on the relative merits of the CAPO option compared to other H$_2$ producing technologies. Previous IEA studies of precombustion decarbonisation for power stations have shown CAPO H$_2$ production to be more cost effective than equivalent SMR based options. This study does not show such a clear distinction and the following discussion sets out two key reasons for this.

In precombustion decarbonisation studies for power stations air compression and preheat is provided by the gas turbine air compressor. This is much more efficient and cost effective than providing a dedicated compressor and preheater as is required for this study. The synergistic air compression stage in precombustion decarbonisation equipment saves on equipment costs, running costs and hence, shadow CO$_2$ debits.

The CAPO process has lots of potential for process heat integration between hot reactor products and reactor feed streams. Previous precombustion decarbonisation studies have maximised heat integration. In this study heat integration has been more limited due to consideration of the effects of metal dusting corrosion. Awareness of the serious corrosion potential of metal dusting is relatively recent and has therefore not been considered in previous studies. Metal dusting is a high temperature corrosion phenomenon experienced when ferrous metals are exposed to CO rich gases at temperatures between 450 °C and 850 °C. The effects of metal dusting can be severe leading to total destruction of material within a few days.

In CAPO option 8, process heat integration has been used to give preheat but limited to reactor product temperatures of 450 °C and below. This results in a cooler reactor feed stream compared to CAPO preheat options 6 and 7 where a gas fired preheater is used. A cooler feed stream results in a raised air demand in order to develop the required temperature for partial oxidation in the reactor. Increased air demand results in increased air compressor capital and running costs.