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Subject: Study into the replacement of the gasifying  
section at Brennstoff Kombinat Vresova A.G.

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## Summary

A preliminary process design was performed to investigate the possibility of integrating new sustainable and more efficient gasification technology into the existing gasification site of the Coal and Gas Fuel Company Brennstoff Kombinat Vresova A.G. situated in the coal basin of North Bohemia, Czech Republic.

The gasification technology needs to be updated in order to generate electricity and comply with future environmental standards.

After comparison of various types of coal gasifiers and commercial processes the pressurized High Temperature Winkler gasification system turned out to be the most suitable for North Bohemian Lignite. Replacing the 26 existing Lurgi moving bed reactors by one HTW fluidised bed gasifier will eliminate the production of undesirable compounds such as tars and phenols and provide the required flexibility for the production of electricity in peak hours. As the Brennstoff Kombinat A.G. has a lack of space for new construction activities, an additional bonus of the HTW system is the large area which will become available after the replacement.

The gasifier was designed to produce approximately 280,000 m<sup>3</sup>n/hr of raw gas at 2,4 MPa. This resulted in a capacity of 340 MW, considerably more than the 200 MW the Brennstoff Kombinat had expected.

The oxygen blown HTW-gasification unit is designed to operate at a pressure of 24 bar, a temperature of 1100 °C, and a coal throughput of 127 t/hr.

It is expected that there will be no problems integrating the new gasification system and the current Rectisol process from both a process engineering and environmental point of view. The latter aspect will be checked by the Brennstoff Kombinat A.G.

Using Lang's and Taylor's method the fixed capital costs were estimated to be between 121 and 151 million DM, which agrees with the Lurgi estimate of 144 million DM. Although not all costs are known it seems that the production costs will be well below the earnings with the current market value of the generated electricity.

The results of the preliminary process design performed for the Brennstoff Kombinat A.G. compare well with the specifications of the KoBra project, that involves a 300 MW plant to be built in Germany.

The report ends with some recommendations for a better performance.

Based on the results of this report replacing the Lurgi gasifiers by a High Temperature Winkler gasification system is an obvious choice.

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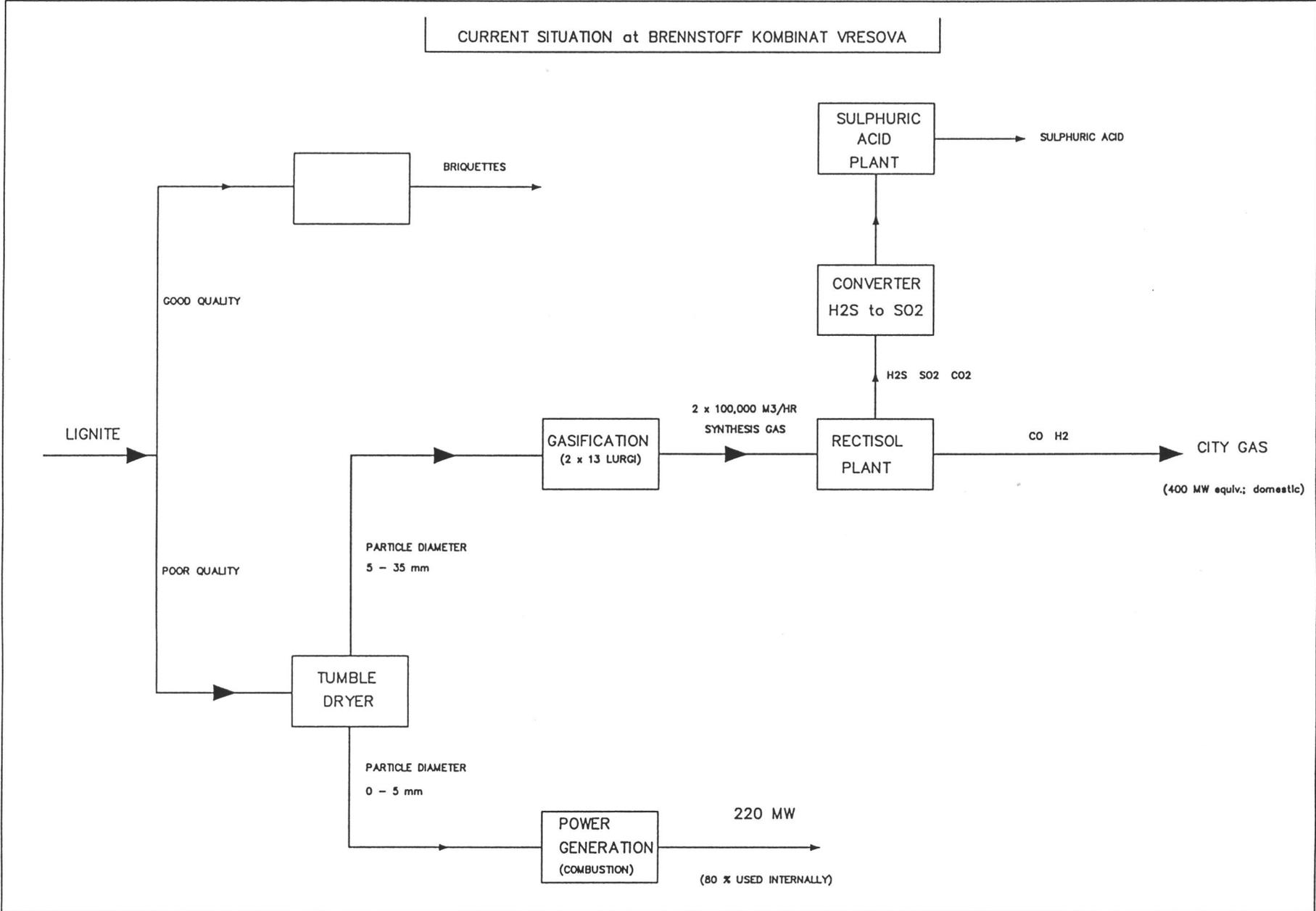


Figure 1: The current process at Brennstoff Kombinat A.G., Vresova.

## 1. Background of the project [1]

### 1.1 The Coal and Gas Fuel Company Vresova

The Coal and Gas Fuel Company Vresova is situated in the Sokolov coal basin between the cities Karlovy Vary and Sokolov. This company carries out lignite mining in this part of the coal basin and produces briquettes, city gas, power, heat, tar, crude benzene, phenol concentrate, liquid ammonia and many other products.

The original processing part of the company was put into operation between the years 1965 and 1970. In 1975 the processing part of the company merged with the open cast mine on Nove Sedlo and thus formed the basis of the current structure of the company.

On July 1, 1990, due to reorganization carried out by the Ministry of Energy the company was established as an independent state enterprise.

On the 18th of October 1993 the company was officially released from government control, and is now a private company, Brennstoff Kombinat Vresova A.G.

The financial turnover of this company with 6400 employees is 5 billion crowns<sup>1</sup> yearly, and gross profit exceeds 1.3 billion crowns.

### 1.2 Challenges and innovations

The current process (figure 1) is focused on the production of synthesis gas which is mixed with methane to form city gas for domestic consumption.

There are two reasons for innovating this set-up:

- . it is official government policy to have the use and generation of city-gas eliminated by 1997.
- . with the present set-up it will be very difficult to comply with the future environmental standards.

As a result the following steps are taken by the company to adapt to the new circumstances:

- . the construction of an electricity generating plant based on the combined use of steam and gas turbines in which the synthesis gas from coal gasification is combusted.
- . performing a preliminary design to investigate the effects and benefits of integrating new gasification technology (replacement of the current Lurgi fixed bed reactors) with existing gas cleaning facilities.

The latter task was undertaken by the authors of this report. A block diagram illustrating the future process is given in figure 2.

<sup>1</sup> In October 1993 one Czech crown equalled 1/15 Dutch guilders.

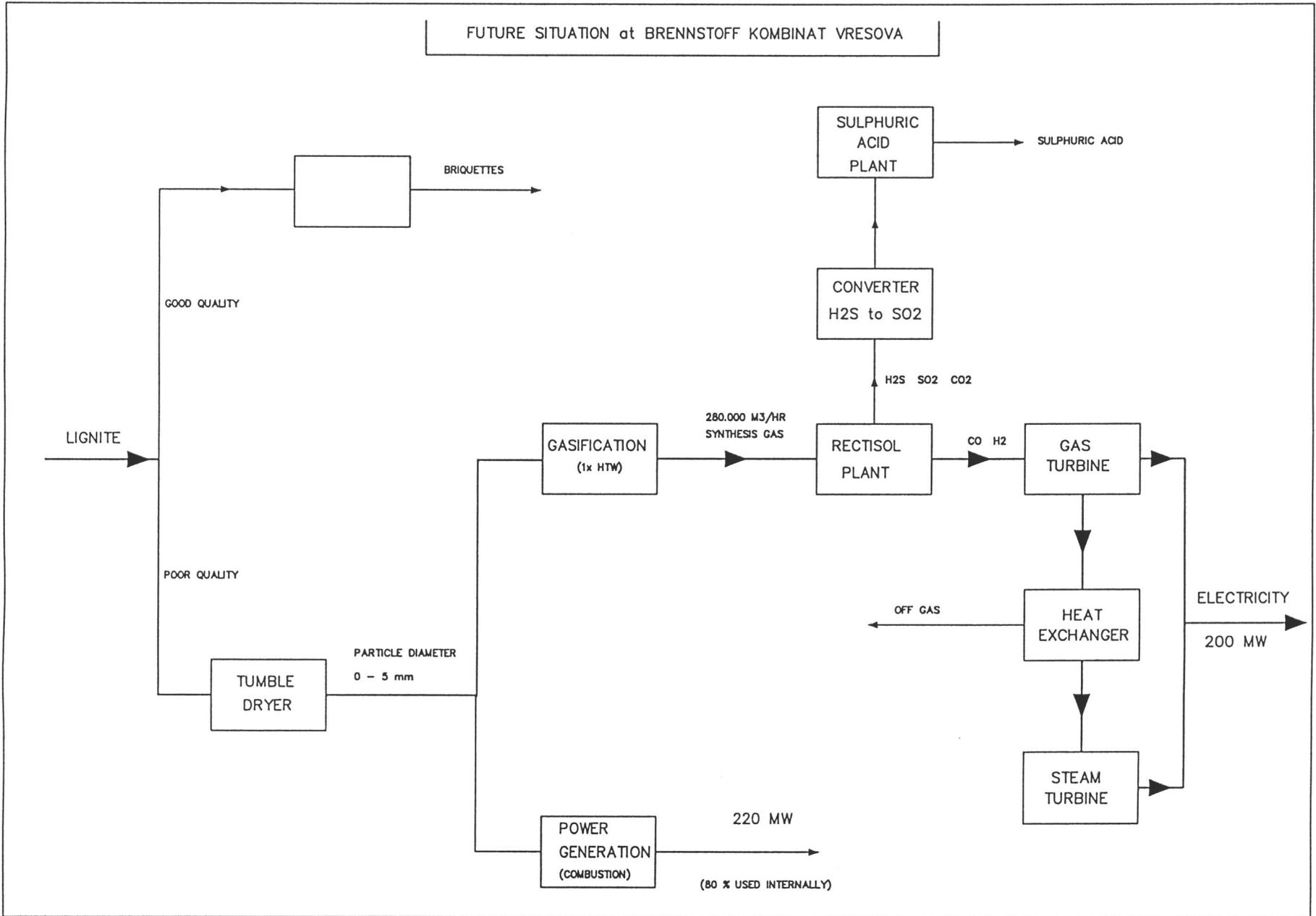


Figure 2: The future process at Brennstoff Kombinat A.G., Vresova.

## 2 Introduction

In this chapter a short introduction to the project, lignite composition, gasification and gas cleaning technologies will be given.

### 2.1 Specification of the project

The current gasification step (26 Lurgi fixed bed reactors incorporated in two trains) produces a large amount of undesirable compounds which are difficult to remove, such as tars and phenols. Therefore it needs to be replaced by a cleaner gasification technology.

The new reactor should be able to produce approximately 280,000 m<sup>3</sup>n/hr of raw gas at 2,4 MPa, equivalent to the desired capacity of approximately 200 MW of electricity [1].

For the cleaning of the raw gas the present Rectisol process has sufficient capacity to comply with future environmental standards. A block diagram illustrating the future process is given in figure 2.

### 2.2 Lignite composition [1]

The lignite found in North Bohemia is of a very high quality (tables 1-3): A high carbon-hydrogen ratio and a low sulphur content.

### 2.3 A comparison of various types of coal gasifiers [2]

Coal is expected to perform an increasing role in the future. Conventional coal-fired electricity generation has resulted in numerous environmental problems, notably emissions of sulfur and nitrogen compounds, both of which have been linked to acid rain, and emissions of particulates. Conventional coal firing technologies only partially solve these problems.

Modern coal gasification combined cycle (CGCC) power technologies, also known as integrated gasification combined cycle systems (IGCC), present electric power producers with important options and opportunities to improve efficiency, environmental performance, and overall cost effectiveness.

In this paper a comparison will be made of various methods of coal gasification.

There are essentially three types of coal gasifiers:

- moving bed (or countercurrent)
- fluidised bed (or back-mixed)
- entrained-flow (or plug-flow) gasifiers

Table 1: Composition of dry North-Bohemian lignite.

Component	Weight percentage
Ashes	16.77
H	4.53
C	61.4
S	0.45
N	0.94
O	15.91

Table 2: Composition of the ashes of North-Bohemian lignite.

Component	Weight percentage
SiO <sub>2</sub>	52.17
Al <sub>2</sub> O <sub>3</sub>	31.94
Fe <sub>2</sub> O <sub>3</sub>	6.17
TiO <sub>2</sub>	3.55
CaO	2.32
MgO	0.82
Na <sub>2</sub> O	1.03
K <sub>2</sub> O	0.34
P <sub>2</sub> O <sub>5</sub>	0.74
SO <sub>3</sub>	0.91
Heavy metals	0.01

Table 3: Composition of the heavy metals fraction.

Component	Parts per million (ppm)
Cd	1.7
Hg	0.484
Ni	325.0
Pb	33.1
Cr	665.0
Mn	78.5
V	150.0
Zn	130.0
Se	3.5
Be	6.4
As	101.2

### 2.3.1 Moving bed gasifier

The moving or fixed bed gasifier involves a series of countercurrent reactions in which large particles of coal slowly move down the bed and react with gases moving up through the bed. The Lurgi gasifier is an example.

At the top of the gasifier the entering coal is heated and dried in the drying zone while cooling the product gas as it exits the reactor. The gas exit temperature ranges from 315 °C for high moisture lignites to 550 °C for bituminous coals. The coal is further heated and devolatilized by hotter gas as it descends through the carbonization zone.

Below this zone, the devolatilized coal is gasified by reaction with steam and carbon dioxide in the gasification zone. The highest temperatures are reached in the combustion zone near the bottom of the gasifier, where the oxygen reacts with the char, which, together with ash, is all that remains of the original coal. Reaction of the char and steam, together with the presence of excess steam, moderates the temperature below the ash slugging temperature in this combustion zone. The whole bed is supported by a grate below the combustion zone where the ash is cooled by releasing heat to the entering steam and oxygen.

Characteristics of moving-bed gasifiers are low gasification temperatures, relatively low oxygen requirements, relatively high methane content in syngas produced, relatively low product gas temperature, production of hydrocarbon liquids such as tars and oils (!), combined with a limited ability to handle fines. A slugging version of the fixed-bed Lurgi gasifier is the British Gas/Lurgi gasifier. This gasifier offers many improvements, including the ability to handle caking coals and coal fines.

### 2.3.2 Fluidised bed gasifier

A fluidised-bed gasifier consists of a back-mixed gasifier where feed coal particles are well mixed with coal (or lignite) and char particles, which are already undergoing gasification. This gasifier is operated at a constant temperature below the initial ash fusion temperature in order to avoid molten slag formation. Some coal particles are reduced in size during gasification and are entrained with the hot raw gas as it leaves the reactor. These char particles are recovered and recycled to the reactor. Ash particles are removed from below the bed and are cooled by heating the incoming steam and recycle gas. Examples of fluidised-bed gasifiers are the high temperature Winkler (HTW) and the Kellogg Rust Westinghouse (KRW) gasifiers. Fluidised-bed gasifiers typically utilize significant fly-ash recycle to capture unconverted carbon. They have a uniform and moderate temperature throughout the whole of the gasifier. They are however limited in their ability to convert high rank coals, but are well suited for using other forms of coal as feedstock.

### 2.3.3 Entrained flow gasifier

The entrained-flow gasifier consists of a plug-flow system in which the fine coal particles concurrently react with steam and oxygen. Residence time is limited to a few seconds. Gasifiers of this type operate at high temperatures which are well above ash slugging conditions. This ensures good carbon conversion and provides a mechanism for removal of ash or molten slag. Entrained-flow gasifiers are used in the Shell coal gasification process, Texaco coal gasification process, Dow coal gasification process by Destec and a process known as Prenflo by Krupp-Koppers. The short residence time which is required in the entrained gasifiers can result in potentially high throughputs at elevated pressures. Gasifiers of the entrained-flow type are known for their high feedstock flexibility. Entrained gasifiers have a small coal inventory, which results in rapid start-up, shutdown and load-following characteristics. The product gases contain no tars and light oils, thus facilitating heat recovery and requiring less gas cleaning and purification. Furthermore, the product gas contains lower quantities of such impurities as mercaptans, ammonia, carbon disulfide, carbonyl sulfide and thiophene, than does that of other types of gasifiers.

Table 4: A comparison of the conditions of the various commercial processes based on the performance in gasifying South-African bituminous coal (comparable to Vresová coal).

	Shell SCGP	Texaco TCGP	Dow DESTEC	Winkler HTW	British Gas Lurgi
Gasifier type	entrained flow	entrained flow	entrained flow	fluidized bed	moving bed
Operating pressure [MPa]	2.1-2.8	3.9	0.1	0.5-2.5	2.5
Gasific. temp. [°C]	1300	>1250	1000	965-1100	
Gasifying medium	st/ox	ox	ox (water slurry)	st/ox	st/ox
Ratio steam/O <sub>2</sub>	0.080	1.0	0.0	0.94	1.0
Oxygen consumption [m <sup>3</sup> /m <sup>3</sup> STP]	0.297	0.321	0.3	0.19	0.20
CGE [%]	81	70.3	80	78-85	85.1
C-conversion [%]	99	94	95	95-96	100

## 2.4 A comparison of various commercial processes [2]

### 2.4.1 Shell Coal Gasification process (SCGP)

The SCGP is based on a dry feed, entrained bed, high pressure, high temperature slagging design. The process can handle a wide variety of coals, ranging from bituminous to lignite, in an environmentally acceptable way, and produces high purity, medium heating value gas that is attractive for use in power generation.

### 2.4.2 Texaco Coal Gasification Process (TCGP)

The TCGP incorporates a single stage, slagging pressurized, entrained-bed downflow gasifier. Rather than using a dry coal feed system, the Texaco gasifier uses a concentrated water slurry of coal ground to a carefully controlled size distribution.

The resulting gasification products in the syngas are predominantly carbon monoxide and hydrogen. Other components include carbon dioxide, nitrogen, hydrogen sulfide, and small amounts of ammonia and carbonyl sulfide.

The TCGP allows three possible configurations that differ in the amount of high level heat that is recovered from the gasifier gas.

### 2.4.3 Dow Coal Gasification Process (Destec)

The Dow coal gasification process (Destec) involves a two-stage, slurry feed, entrained-flow, slagging gasifier. The first stage assures high carbon conversion and optimum slag removal. The second stage reduces the raw product gas temperature to about 1000°C. This latter step helps to improve cold gas efficiency relative to other slurry fed processes and to lower waste heat recovery costs.

The subbituminous coal or lignite is almost totally gasified by partial combustion to CO, H<sub>2</sub>, CO<sub>2</sub> and H<sub>2</sub>O. The sulphur is almost exclusively converted to H<sub>2</sub>S.

### 2.4.4 High Temperature Winkler Process (HTW)

The HTW process developed by Rheinbraun is especially targeted for the gasification of brown and hard coals, peat and biomasses in a fluidised-bed gasifier.

The old atmospheric process has the following distinctive features compared to others which are commercially available:

- low oxygen consumption.
- simple coal preparation.
- good partial load behaviour over a wide range of performance.
- simple start-up and shut-down conditions of the gasifier.
- high operational reliability.
- no by-products in the raw gas, such as tars and liquid hydrocarbons.

The further development into the Rheinbraun HTW process added three major characteristics to the already mentioned advantages of the atmospheric Winkler gasifier:

- by increasing the pressure to 25 bar the reaction rate and thus the specific performance per gasifier cross-section unit was increased.
- by recirculating the dust fines entrained from the fluidised bed it was possible to essentially increase the C-conversion rate.
- by increasing the temperature the methane content in the raw gas was reduced and the carbon conversion rate and thus the gas yield increased. As a result favourable preconditions were obtained in regard of gas quality and specific gas yield for the production of synthesis gas.

#### 2.4.5 British Gas / Lurgi Slagging Gasifier

The technology developed by the British Gas Corporation and Lurgi started with the dry ash Lurgi gasifier and incorporated enhancements such as operation at a higher temperature that melts the coal ash to slag. Compared to the older Lurgi dry-ash gasifier a significant efficiency advantage is gained by reducing the steam requirement to only about 15% of the amount required by the older Lurgi process.

Compared with the dry-ash Lurgi gasifier the raw gas from the slagging gasifier has lower  $H_2O$ ,  $CO_2$  and  $CH_4$  and a higher  $CO$  content, primarily because of the lower steam consumption.

Recycle of the tar and oil in the slagging gasifier increases the gas yield by reducing the net hydrocarbon liquid production to only naphta and phenols.

The slagging gasifier offers additional advantages over the dry-bed gasifier in terms of feed flexibility because it can handle caking coals and a significant amount of fines.

## 2.5 Gasification mechanism [23]

During the coal gasification, the following overall reactions occur:

			$H_r(800K)$ in $\text{kJ}\cdot\text{mole}^{-1}$
$\text{C} + \text{H}_2\text{O}$	----->	$\text{CO} + \text{H}_2$	135.6
$\text{C} + \text{CO}_2$	----->	$2 \text{CO}$	172.5
$2 \text{C} + \text{O}_2$	----->	$2 \text{CO}$	- 222.0
$\text{C} + \text{O}_2$	----->	$\text{CO}_2$	- 394.2
$\text{C} + 2 \text{H}_2$	----->	$\text{CH}_4$	- 87.3
$2 \text{CO} + \text{O}_2$	----->	$2 \text{CO}_2$	- 172.2
$\text{CO} + \text{H}_2\text{O}$	----->	$\text{CO}_2 + \text{H}_2$	- 36.9

Table 5: High temperature adsorptive desulphurisation processes.

Process	Status of development (1985)	Acceptor	Acception		Regeneration	
			T (°C)	P (kPa)	method	T (°C)
Conoco	completed	MgO.CaCO <sub>3</sub>	900	15	H <sub>2</sub> O-CO <sub>2</sub>	600-760
U.S. Steel	in progress	MgO.CaO	> 816	1	proprietary	
Battelle Columbus	completed	Supported Fe <sub>2</sub> O <sub>3</sub>	538-816	1	air	593
IMMR	in progress	Gasifier ash (Fe <sub>2</sub> O <sub>3</sub> )	371-816	4-9	air	427-649
MERC	in progress	Supported Fe <sub>2</sub> O <sub>3</sub>	538-816	1-20	air	538-816
Kennecott	terminated	Cu/CuO	482-496	20-25	air	816
IFP	in progress	Supported ZnO	400-600	?	O <sub>2</sub> -containing gas	600-900
U.S. Steel	?	MnO <sub>x</sub> -Al <sub>2</sub> O <sub>3</sub>	700-1000	?	O <sub>2</sub> -containing gas	850-1150

## 2.6 Cleaning of the synthesis gas [2,3]

The process currently used in Vresova, the Rectisol process, will be maintained, and will therefore be integrated with the new gasification process. The process will however be converted from selective to non-selective. For review of the gas cleaning performance under the new circumstances software developed in Vresova will be used. Regarding the cleaning of the synthesis gas the aspects of effectivity and process integration will be of paramount importance.

The amount of synthesis gas produced by the new process will be approximately 280,000 m<sup>3</sup>n/hr (an increase of 40%). There will however still be sufficient gas cleaning capacity as currently only 50 percent of the Rectisol plant capacity is used.

We wish to state clearly that the best possibilities for increasing the efficiency in the future will be provided by high temperature gas cleaning techniques. Adsorptive desulphurisation processes are being developed for various types of acceptors.

Although many of these techniques are nearing completion and are considered commercially viable, so far the industry has been reluctant to integrate these technologies into new processes. An overview of some of these techniques is given in table 5.

### Rectisol Process

The Rectisol process developed by Lurgi and Linde, is particularly suitable for high pressure synthesis gas production. Low temperature operation enlarges the absorption capacity and minimizes the solvent losses. In practice two Rectisol processes are encountered:

- *The standard Rectisol process.*  
In this gas treating process, all acid gas constituents are absorbed simultaneously. A disadvantage of the standard Rectisol process is that the sulfur compounds are discharged with the entire volume of carbon dioxide and are thus diluted.
- *The selective Rectisol process.*  
The selective Rectisol is capable of producing an hydrogen rich off-gas stream that could be processed in a Claus plant for sulfur recovery. An almost sulfur-free off-gas stream of carbon dioxide is a second product. This can be vented or utilized, for example in urea plants which are often built in urea plants in conjunction with ammonia plants.

### 3 Process structure [4]

As can be concluded from the previous chapter, the Rheinbraun HTW-process is especially suitable for the gasification of brown coal in a fluidised-bed gasifier. In Germany, where more than 50% of the world's lignite reserves are located, the atmospheric Winkler process (1920s) has proven by far the most efficient gasification method. Over the last 15 years extensive research has been done to develop the pressurized Winkler-process (HTW) for commercial use, and since October 1993 the first commercial HTW-plant is in operation. Therefore the preliminary process design will be performed with the HTW-gasifier.

The flow sheet of the proposed gasification process (appendix I.1) will be clarified below.

The process uses a pressurized refractory-lined gasifier. The feedstock (crushed, predried lignite with a moisture content of 12% and diameter of 0-5 mm) is pressurized via a lock hopper system and mechanically transported into the gasifier via a screw feeder. Gasifying agent (oxygen of 95% purity at 3,3 MPa and steam at 3,5 MPa [1]) is fed into the gasifier at different levels. In the fluidised bed the feedstock is devolatilized and partly gasified. Ash containing various oxides (tables 2&3) is discharged at the bottom of the reactor. In the upper part of the gasifier, the post-gasification zone, the char particles entrained in the gas are gasified further and higher hydrocarbons are decomposed at temperatures up to 1100 °C. The gas virtually does not contain any tars. The raw gas (containing CO, CO<sub>2</sub>, H<sub>2</sub>, CH<sub>4</sub>, H<sub>2</sub>O, H<sub>2</sub>S and coal dust) leaving the gasifier passes through a cyclone where the coarser, carbon-containing particles are separated and directly recycled to the fluidised bed. In a second cyclone, finer dust particles are removed. The gas then passes a waste heat boiler to generate steam. The gas is cooled further in a quench and cleaned in a Venturi scrubber and a scrubbing tower. The wash water is recycled at an elevated temperature to saturate the gas with water vapour. The steam-to-gas ratio for a following shift conversion is established thereby without adding additional steam.

In the Rectisol unit feed gas is cooled by cold product streams. In order to prevent icing, methanol is injected. The condensed methanol-water mixture is separated in a methanol water column. In the upper section of the main absorber, CO<sub>2</sub> is absorbed by lean, cold methanol, in the lower section H<sub>2</sub>S and COS are removed with a part of the CO<sub>2</sub>-rich methanol.

After recovery of coabsorbed H<sub>2</sub> by pressure reduction, CO<sub>2</sub> is produced by flashing the loaded methanol. In order to increase the CO<sub>2</sub> product, the cold solvent is partly warmed up against lean and loaded methanol. The bulk of the still absorbed CO<sub>2</sub> is stripped off by N<sub>2</sub> in the H<sub>2</sub>S enrichment column. In order to produce a sulphur-free CO<sub>2</sub> and tail gas, H<sub>2</sub>S and COS are reabsorbed in the upper sections of two columns by only CO<sub>2</sub> loaded solvent. After heat exchange against lean methanol, the H<sub>2</sub>S enriched solvent is regenerated completely in the regeneration column with steam. After condensation of methanol vapour, the H<sub>2</sub>S fraction is delivered to the sulphuric acid unit. Refrigeration requirements of the plant are covered by a conventional refrigeration unit.

## 4 Design of equipment for HTW gasification

Design calculations were performed independently from the information provided by Lurgi to facilitate independent cost accounting and provide Brennstoff Kombinat with a second estimation of the costs.

### 4.1 Physical constants used in the design calculations

In the process great differences in temperature (20 - 1100°C) are encountered. The approximations made and relations used for calculation of densities, viscosities and specific heats as a function of temperature and pressure are listed in appendix II.1.

### 4.2 The gasifier

#### 4.2.1 The gasification simulation model

The gasifier will be simulated with Cycle-Tempo, a model that can be used to simulate all kinds of energy conversion processes, including ICGCC (Integrated Coal Gasification Combined Cycle). The model has been developed at the Laboratory for Thermal Power Engineering of the Delft University of Technology.

##### 4.2.1.1 Build-up of Cycle-Tempo programs

In order to simulate a coal gasifier with the Cycle-Tempo modelling system, little programs of various constituents will be made. These blocks forming the program are the following:

- Cycle definitions
- System definitions
- Definition of used aparates
- Mediums
- Topology of the system

##### 4.2.1.2 Cycle-Tempo run parameters

To run the Cycle-Tempo program simulating the fixed bed gasifier, various parameters have to be set. These parameters are built up of the following constituents:

- Gasifier inputs
- Gasifier outputs
- Inside gasifier
- Differences

##### 4.2.1.3 Cycle-Tempo output information

The Cycle-Tempo modelling system generates the following output information:

- Energy balance of the apparatus
- Data for all pipes
- Heating values of the media in the pipes which are gasmix or fuel (MJ·kg<sup>-1</sup>)
- Composition of the medium in the pipes which contain gasmix (mole fraction).

#### 4.2.1.4 *Fitting Cycle-Tempo to pilot plant results*

The objective is to manipulate the Cycle-Tempo run specifications in such a way that the output composition of the syngas resembles the output composition of Rheinbraun literature values as closely as possible, so that Cycle-Tempo can be used for the HTW-gasifier. This will be done by analysing the following different situations:

- Literature input
- Literature output
- Cycle-Tempo input
- Cycle-Tempo output

As a dependency between the capacity of Cycle-Tempo to simulate a HTW-Gasifier, and Cycle-Tempo run parameters has to be determined, analyses for two Rheinbraun pilot-scale tests will be performed.

The outcome of the above mentioned simulations indicated that the best results were obtained, when the HTW-gasifier was modelled with a Cycle-Tempo reaction temperature 250°C below the actual gasification temperature. Using this temperature difference, the calculated composition differed minimally from the experimental one.

#### 4.2.2 Optimization of the process conditions for the 200 MW HTW gasifier

To obtain the necessary insight in the influence of gasification temperature, specific oxygen consumption and steam/oxygen ratio on raw gas composition and required cooling capacity, the following simulations were done.

- Gasification temperature: 1000, 1025, 1050, 1075, 1100 °C
- Specific O<sub>2</sub> consumption: 0.7133, 0.7621, 0.8110 kg O<sub>2</sub>·(kg coal(daf))<sup>-1</sup> (1100°C)
- Steam/O<sub>2</sub> ratio: 0.9, 0.95, 1.0 mole steam·(mole O<sub>2</sub>)<sup>-1</sup> (1100°C)

The results of these simulations are labelled in appendix II.2.2. The optimal gasification temperature here from is 1100°C (even higher would be possible, but Rheinbraun A.G. mentions this as the maximum temperature [5]). At this high temperature the amount of undesirable tars, phenols, higher hydrocarbons and son on is minimal. The optimal specific oxygen consumption is 0.7133 and the optimal steam/O<sub>2</sub> ratio is 0.9 (optimal being defined from both an energetic and an economic point of view).

At these conditions both the energetic and economic efficiency (measured in terms of energy concentration in the raw gas stream, preferably in the form of combustion heat, required oxygen and steam as gasifying agents, released amount of steam from cooling etc.) are maximal. Graphical presentations of the above can be found in II.2.2. The only missing heat flows are those present in the ash and coal dust stream, but these are constant. These heat flows are released in the screw dischargers described in paragraph 4.4.

From the tables and graphs it would seem logical to use even lower amounts of oxygen and steam to reach even higher concentrations of CO, H<sub>2</sub> and CH<sub>4</sub>, but this cannot be done without a highly undesirable decrease of the carbon conversion (which would more than undo the advantages aimed at). Rheinbraun A.G. mentions the values above as the minimum at which the high carbon conversion of 96% can still be achieved. This has to do with a certain 'excess' of oxygen and steam, which is necessary to reach the high reaction rates needed because of the extremely low residence time. Furthermore, the total gas flow must be kept relatively high to maintain a good fluidisation.

Assumptions hereby were:

- Net calorific value (NCV) of the dried lignite equals 23,000 kJ·kg<sup>-1</sup> [6].
- Carbon conversion is 96% [5,7].
- Raw gas composition equals Cycle-Tempo equilibrium composition at 250°C below the actual gasification temperature.
- Ash temperature lies 300°C below gasification temperature [5].
- There is no carbon present in the ash stream.
- Gas outlet temperature lies 100°C below gasification temperature.
- Overall heat losses equal 10% of released reaction heat (Rheinbraun calculation [5]).
- Complete compensation for these heat losses is done by subtracting them from the cooling.
- The energy conversion efficiency of the whole process equals 43.8 %.

Because of the limited programming possibilities in Cycle-Tempo, it was inevitable to mix the raw gas stream with the unconverted coal dust. Therefore the results of the Cycle-Tempo calculations needed recalculation to compensate for this dust. An example of such a recalculation can be found in Appendix II.2.1.

### 4.2.3 Dimensions of the fluidised bed

The HTW-fluidised bed gasifier is dimensioned for the optimal modelling conditions described in paragraph 4.2.2 and appendix II.2.3. The model used to describe the hydrodynamic behaviour is based on a simplified model for a fast fluidised bed circulation system described by Kunii & Levenspiel [8]. The model is straightforward and includes the following assumptions:

- The porosity of the solid fraction at the end of the gasifier is  $1.0 \cdot 10^{-4}$ .
- The porosity of the solid fraction above the reactor in the pipe leading to the cyclone is  $5.0 \cdot 10^{-5}$ .
- The porosity of the solid fraction just at the end of the fluidisation zone is 0.15.
- The fluidisation zone has a conical geometry and a smaller mean inlet diameter than the entrainment zone.
- From the assumption that particles leaving the gasifier have the same terminal velocity as the superficial gas velocity at the outlet of the gasifier it follows that these particles have a diameter  $d_p$  smaller than or equal to 1.2 mm.
- The gasification reactions will partially take place in the fluidised zone, but mainly in the gasification zone.

Dimensioning a HTW-gasifier fed by Vresova lignite on the basis of this simplified model leads to a gasifier with an internal diameter of 2.75 m, a fluidisation height of 2 m, a total bed height of 12 m and an absolute solid recirculation flow of 1.52 kg/s [appendix II.2.3].

The construction material used for the gasifier must be resistant to temperatures of 1100°C or more, pressures up to 30 bar and resistant to both reductive and oxidative environments. Choosing the construction material and calculating the wall thickness using the method described by Coulson & Richardson [9] resulted in titanium stabilised 18Cr/8Ni stainless steel (321) with a minimal wall thickness of 75 mm.

### 4.3 Bunkers

The flow sheet (appendix I.1) contains nine bunkers, which are used for pressurization of coal and depressurization of ash and coal dust. The three largest bunkers L1,2&3 were designed using the procedure for the design of pressure vessels [9].

#### 4.3.1 Configuration and size

Various height-over-diameter ratios and configurations are possible for the bunkers. The most conventional ratio would be  $H : D = 2 : 1$  for each of the chambers.

Practical experience at the Delft University of Technology's coal burning pilot plant suggests that the optimal configuration is two parallel trains of three pressure chambers. This allows for easier and smoother loading than just one train. Furthermore it results in an efficient use of high pressure gas, since the gas can be used in one train after another to pressurize the chambers.

126.5 tons/h of lignite is to be fed to the gasifying reactor which is 63.25 tons/h per train.

The lignite in each chamber is loaded and discharged in cycles of 20 minutes (5 minutes for filling, 10 minutes for pressurizing, 5 minutes for discharging).

Each of the chambers will therefore contain a maximum of 21.1 tons of lignite at any given time, which equals about  $21 \text{ m}^3$  of lignite. (The density of lignite containing 12 % water equals approximately  $1000 \text{ kg/m}^3$ ). About  $4 \text{ m}^3$  of gas will be added to bring the pressure to 24 bar, resulting in a total volume of each individual chamber of  $25 \text{ m}^3$ . [10]

In order to remain on the safe side, the volume of the conical discharge of the chambers will be ignored.

With an aspect ratio of 2 (height : diameter = 2 : 1) one finds that each chamber will have a height of 5 meters and a diameter of 2.5 meters.

#### 4.3.2 Wall thickness

The wall thickness will be determined according to the procedures stated in Coulson & Richardson [9]. Where possible ample margins have been allowed for.

The wall thickness is determined on the basis of the maximum internal pressure the vessel is likely to have to withstand, and the results will be checked by performing an analysis of the stresses induced by pressure and weight.

The following assumptions are made:

- the construction material stainless steel 18Cr/8Ni (unstabilised) with a maximum allowable design stress of  $165 \text{ N/mm}^2$  (0 - 50 °C)
- spot radiography is applied to the welds which results in a welding joint factor of 0.85
- the corrosion allowance in the wall thickness is the maximum allowed value (4 mm) because of the corrosive and abrasive nature of the coal
- the chambers do not support each other, but are supported by a dedicated structure. They will therefore only support their own weight and the weight of the contents of the chamber, but not the chambers above or below
- the chambers consist of a cylindrical section, with an ellisoidal top and a conical bottom with a half apex of 60°.

The calculations result in the following wall thicknesses:

- ellipsoidal head : 27 mm
- cylindrical section : 27 mm
- conical section : 50 mm (at point of maximum thickness, just below the cylindrical section)

These values include a very generous corrosion allowance and a large number of safety factors, in particular for the conical section.

No wind and excentric loads are taken into account, since these will be absorbed by the supporting structure.

Table 6: Requirements for the screw feeders and dischargers. The throughput given in this table is the throughput of the optimized run. For design of the screw feeders an ample margin will be necessary for the throughput.

Number in flow scheme	Material to be transported	Specific heat (kJ kg <sup>-1</sup> °C <sup>-1</sup> )	Throughput (kg s <sup>-1</sup> )	T <sub>screw, in</sub> (°C)	T <sub>screw, out</sub> (°C)	Heat load (MW)	Steam (35 bar, 400°C) production (kg s <sup>-1</sup> ) due to cooling
C2	coal	1.3	35.14	25	800	-	-
C3	ash	1.2	4.93	800	200	3.548	1.13
C4	ash	1.2	4.93	25	25	-	-
C5	coal dust	1.3	1.24	1000	200	1.286	0.41
C6	coal dust	1.3	1.24	25	25	-	-

#### 4.4 Screw feeders and dischargers

The design of screw feeders and dischargers is a very specialised and highly empirical commercial field. Therefore very little information about design procedures is available. Thus only the requirements for the screw feeders and dischargers are presented (table 6).

#### 4.5 Cyclones

For dimensioning a cyclone a choice has to be made between a high efficiency cyclone and a high gas rate cyclone. As the cyclones are not the final particle separation step a high gas rate is the more important design factor.

The first cyclone  $Y_1$  was designed with a cut size of  $34 \mu\text{m}$  and the second cyclone  $Y_2$  with a cut size of  $11 \mu\text{m}$ . The cyclones were designed with a square inlet. The calculations and resulting specifications of the cyclones are presented in appendix II.4. The pressure drop over the cyclone section is 0.047 bar.

#### 4.6 Heat exchangers

The raw gas is cooled in two heat exchangers which are operated in series. The raw gas enters with a temperature of  $1000^\circ\text{C}$  and is cooled to  $250^\circ\text{C}$ . Further cooling would result in condensation of salts [11].

The heat exchangers were dimensioned with one shell pass and two tube passes. In both heat exchangers the raw gas was allocated to the tubes as the raw gas with the highest fluid rate, highest temperature and greatest tendency to foul.

The design calculations can be found in appendix II.5.

In the first heat exchanger  $18.7 \text{ kg}\cdot\text{s}^{-1}$  steam of 35 bar and  $400^\circ\text{C}$  is produced. The second heat exchanger produces  $7.2 \text{ kg}\cdot\text{s}^{-1}$  steam of 5 bar and  $200^\circ\text{C}$ . The pressure drop over the two heat exchangers is 0.28 bar.

#### 4.7 Scrubbing section

This section of the plant includes the quench, the Venturi scrubber, the scrubbing tower and the pump. To dimension this section the following assumptions were done [11]:

- In both the quench and the scrubbing tower the gas is cooled to its adiabatic saturation temperature [12];
- The pressure drop from gasifier till quench and over the Venturi scrubber are both 1 bar, thus resulting in a pressure of 23 bar in the quench and 22 bar in the scrubbing tower;
- The water used in the scrubbing tower enters at  $90^\circ\text{C}$ , the waste water from the quench leaves it at  $115^\circ\text{C}$  and is then cleaned, cooled and recycled to the scrubbing tower;
- As appears from the flow sheet, the water from the scrubbing tower is divided into two flows, of which one goes to the quench (the major part) and the other to Venturi scrubber (this is really a negligible amount of water).

The dimensioning calculations for the scrubbing section can be found in appendix II.6.

## 5 Integration of HTW and Rectisol

The Brennstoff Kombinat Vresova A.G. is currently performing simulations of the behaviour of the Rectisol gas cleaning process with their dedicated software package. It is clear that the integration of the HTW-gasifying installation with the Rectisol process will not be problematic from a process engineering point of view.

The main question however, is the degree with which environmental standards are met when using the Rectisol process to clean the raw gas produced by the HTW-gasifier.

The limit which will have to be met is a maximum emission to the atmosphere of 2000 mg/m<sup>3</sup> of SO<sub>2</sub> which is equivalent with 1062.5 mg/m<sup>3</sup> of H<sub>2</sub>S.

This is the current value which is accepted in Germany for power generation stations with a capacity of less than 300 MW [1]. However, in 1994 new emission standards will be drawn up for the European Community, which will have to be met.

## 6 Molar and heat balances

The molar and heat streams in the HTW gasification process (appendix I.1) are incorporated in appendix III.

The following remarks have to be made concerning these streams:

- the oxygen molar stream includes the oxygen bound in the lignite.
- the size of stream 3 is unknown because the size of the recirculation stream of carbon particles from the first cyclone to the reactor is not known precisely.

## 7 Process control

The aim of the preliminary process control design is to develop a control structure and - if necessary - to estimate time constants, with which the product gas throughput and quality can be kept within acceptable margins [13].

### 7.1 Controlling the product gas throughput

Besides the gasification temperature (and pressure) and the carbon conversion rate, the coal, oxygen and steam feed determine the raw gas production rate. The best manner in which this rate can be controlled is by using the real manipulated variables (i.e. the variables that can be manipulated directly), that is the coal, oxygen and steam feed.

If one chooses to use the gasification temperature to control the production rate, this has an intolerable disadvantage: changing the gasification temperature (by changing the coolant flow) also changes the gas composition and this you don't want. Moreover, neither the gasification temperature nor the carbon conversion rate is a real manipulated variable.

Therefore, the gas production rate is controlled by manipulating the coal, oxygen and steam feed, thereby observing the fixed ratio between these three. Since the rate is not known until measured, a feedback flow (PID-)controller has to be applied. To maintain the fixed ratio between coal, oxygen and steam, two ratio controllers have to be used as well.

### 7.2 Controlling the product gas quality

The quality of the raw gas produced is determined by its heat of combustion, its sensible heat and its pressure energy. To keep this quality as high as possible, both the gasification temperature and the carbon conversion rate, which influence it directly, have to be kept very close to their setpoint.

The gasification and gas outlet temperature can be regulated by adjusting the flow of coolant (water). Here also a feedback flow (PID-)controller can be applied.

The carbon conversion rate has to be kept as high as possible. When this rate drops, the correct countermeasure is to use relatively more oxygen and relatively less steam. The gasification rate and consequently the carbon conversion rate will increase. The total oxygen and steam flow will be kept constant, but the ratio oxygen/steam is raised. It is unwise to change the total gas flow, because this will influence the fluidisation characteristics.

The mass flow of coal dust is measured and compared with the coal feed. The ratio between dust and feed is proportional with the fraction of unconverted carbon (constant coal composition assumed). Only in the case of a decrease of the carbon conversion (an increase of the dust/feed ratio), it is desirable to eliminate this in the way just described. In the case of an increase, there will be no need for controller action.

### 7.3 Additional disturbances

The composition of coal and oxygen (95% O<sub>2</sub>; 5% N<sub>2</sub>) are assumed constant. Therefore the fixed ratio between the three reactants, coal, oxygen and steam will not have to be changed.

The inlet temperatures of reactants and coolant water could vary, but these eventual disturbances are automatically eliminated by the temperature control of the gasifier.

### 7.4 Time constants of the controllers

The estimated residence or space time in the reactor plus cyclones is estimated to be 15 s (Gasifier 10 s, 1<sup>st</sup> cyclone 4.1 s, 2<sup>nd</sup> cyclone 0.5 s). Between 5 and 10 % of this value can be used as an approximation for the time constant for all controllers described above, approximately 1 second.

## **8 Safety, health and environment**

The concentration of industrial activity in the relatively small area of North-Bohemia is very high. This has a big impact on the region, most notably on the safety and health of the population.

The production site of Brennstof Kombinat Vresova is situated in the mountains of the North-Bohemian coal basin, about 60 kilometers away from the capital city of the Czech Republic, Prague. The lay-out of the facilities of the Brennstoff Kombinat has a very high density: the chemical process equipment and power generating installations are situated close to each other on a relatively small surface. This creates a situation with an inherent risk, and leaves little room for construction activities.

Because of this highly concentrated lay-out of the power generation facilities it is of great importance to obtain an insight in the risks of the process and possible environmental problems. To gain some insight in the risk of the designed process, it is important to distinguish between the risk of an explosion and toxicological or health risks. These will both be discussed briefly.

### **8.1 Safety, health and environment at the existing Vresova coal gasifying and power generation plant**

#### **8.1.1 Safety risks at the existing Vresova coal gasifying and power generation plant**

The prevention of an explosion occurring in the gasifying plant is of great importance. The presence of small flammable particles (coal dust), nearly pure oxygen and flammable product gases as well as the high temperatures and pressures at which the process is operated create a risk of devastating (dust) explosions. These could be caused by mechanical malfunctioning or accumulations of dust in the plant amongst other possibilities.

Besides the risk of explosion within the gasifying plant itself, there is the risk of serious damage to the other installations on the site since these are built in close proximity. The Rectisol gas cleaning unit for example is situated close to the existing gasifier installations containing 26 Lurgi moving bed reactors. There exists therefore a real possibility of a chain reaction occurring if an explosion occurs in any given plant on the site.

It is very important to use the available space for building the HTW plant in such a way, that the plant will be as far away from other hazardous installations as possible.

Furthermore it must be stressed that a proper concern for hygiene in the plant can help in preventing dust explosions. This will be especially true when the plant is being started after maintenance and cleaning, because these activities can result in large amounts of free dust in the plant. Installing dust filters in appropriate places can help in minimizing the risk of dust explosions.

It is strongly recommended to perform detailed safety studies, both for the existing installations and for the new gasifier which is to be built. It may be advisable to hire foreign, independent expertise to perform this task. Performing the studies will not only greatly contribute to safety in the plant, but may also have a positive effect on the Brennstoff Kombinat and its surrounding environment as a whole.

### 8.1.2 Health in relation to the existing Vresova gasifying and power generation plant

The toxicological risks of the process plants are mainly caused by the product gas containing about 38% carbon monoxide and 0.15% hydrogen sulfide. These highly toxic substances form a threat to human and animal health when emission takes place due to an accident (appendix IV.1). This risk will not disappear when the old gasification system is replaced by the HTW-gasification process, because it originates in other parts of the facilities. It should therefore be an important point in any safety policy which is to be developed to minimize the risk of exposure to these substances and to promote general awareness. This is especially valid for the existing Rectisol gas cleaning process, where streams with a high concentration of hydrogen sulfide are present.

Of specific interest should be the protection of the long-term health of the employees. Although this is a complicated problem, it should be one to which the highest priority is given.

In particular, it should be a matter of prime importance to create as healthy a work environment for the employees as possible.

The current situation may have an undesirable effect on the health and wellbeing of the employees. Hygiene in the facilities is poor, due to the frequent emissions of dust and other pollutants. This situation is mainly caused by the fact that the existing plants are partially outdated. Furthermore, no real work seems to have been done in the past to bring the hygiene of the plant upto a level which would be acceptable in (for example) Germany or The Netherlands. This is partially due to what seemed to be a lack of interest by the regulating bodies in the past, an attitude which is now rapidly changing.

As was the case with safety, it may be advisable to have foreign independent experts perform studies into the health effects of the current installations and take corrective action wherever needed. It is felt that the expenses for these studies will easily repay themselves in future.

### 8.1.3 Environmental effects of the existing Vresova coal gasifying plant

The existing gasifying system in Vresova was built in the late sixties and has been in operation for about 25 years. This created an environmental problem because of the emission of large amounts of  $\text{NO}_x$ ,  $\text{SO}_2$ ,  $\text{H}_2\text{S}$  and other pollutants.

Over the years Brennstoff Kombinat Vresova paid more and more attention to environmental issues and implemented cleaner technology in the existing power plant such as:

- building the Rectisol gas cleaning unit, which resulted in a drastic reduction of the emission of gaseous pollutants.
- installing systems which remove about 10,000 tons of carbohydrates and sulfur oxides from the plant's smokestack.
- reconstruction of the electrofilters of the power generating and coal treatment plants, which realising a considerable reduction in the emissions of fly ash und coal dust.
- taking measures to reducing the ammonia content in the effluents from the gasification plant.
- building a waste water purifying installation for internal and external use.

A major problem of the existing power plant which has not been solved yet is the production of tars. This is caused by the fact that the existing trains of moving bed Lurgi gasifiers operate at relatively low temperatures. As a result many of the hydrocarbons and aromates which are formed during the pyrolyses are not cracked. They leave the reactor as a tar, thereby creating the problem of toxic waste.

Upto this moment these tars are treated chemically, physically and biologically to remove the most damaging substances as much as possible, after which the residue is dumped in a nearby lake. The dumping of these tar residues is seen by the Brennstoff Kombinat as a serious environmental problem, which can not be solved easily if the Lurgi gasifiers continue to be used.

## 8.2 Positive effects on safety, health and environment of implementing the HTW-gasifier into the Vresova process site.

Replacing the Lurgi gasifiers with the HTW-gasifier will greatly benefit the SHE performance of the Brennstoff Kombinat.

Because the HTW-gasifier operates at high pressures and temperatures, only solid waste in the form of ash containing SiO<sub>2</sub>-like materials and unconverted coal dust is produced. This is a major improvement compared to the old Lurgi system which produced large amounts of tars, aromates, contaminated waste water and other pollutants. The practice of discharging environmentally dangerous tars in the nearby lake will therefore vanish when implementing the new HTW-gasifier. The next problem, how to clean the contaminated lake in the most efficient way is a matter of further investigation.

If the HTW-gasifier is integrated with an efficient gas treating system, an efficient predrying system, a combined cycle electric power plant and optimal use of waste heat is made, then the integrated power plant may, in our opinion, be seen as highly sustainable technology.

The site in Vresova will first be equipped with a HTW-gasifier and a new gas- and steam-turbine power plant. Combined with the already existing Rectisol gas treating-unit this adaption evolves to an integrated gasifier combined cycle system resembling the KoBra 300 MW IGCC power plant developed by Lurgi, Rheinbraun e.a.. Not all health, environmental and safety problems will immediately disappear, but this new sustainable technology can serve as a good basis for solving these problems and simultaneously be an example to other eastern european countries.

### 8.2.1 Hazard and operability studies of the HTW-gasifier

This Hazard and operability study is performed for the gasifying section of an ICGCC power plant [Appendix IV.2]. Performing a HAZOP-study of the whole plant is too much work in comparison to the length of this project, and requires great detail of design. The reason for choosing the gasifier for a HAZOP is an obvious because of the presence of high toxic and explosive gases while processing the gasifier. The risk analyses of the gasifier are of great importance. The checklist of guide words and HAZOPS sequence listings which were used are based on Coulson & Richardson standards [9].

After performing the HAZOPS it could be concluded that it is very important to have some form of permanent monitoring and control of the gasifier conditions, and the possibility to shut the process down if the desired conditions are lost to a serious extent. There is however not enough design detail available to specify this any further.

## 9 Cost estimation

### 9.1 Fixed capital costs

Three different methods were followed to determine the fixed capital costs of the gasifying section. Two of these were modular methods (Taylor's and Wilson's methods) and one was factorial (Lang's method).

The results yielded by these three methods were compared with the estimate provided by Lurgi A.G. (which, as an engineering firm, is involved in Rheinbraun's HTW-project).

#### 9.1.1 Lurgi estimate for the fixed capital costs

Lurgi A.G. estimates that a complete HTW combined cycle (i.e. including gas turbines) costs 3000 DM per kW of produced power [14].

If the turbine section already exists (as will be the case in Vresova) 1200 DM/kW can be subtracted from this figure, leaving 1800 DM/kW for the gasifier and gas cleaning facilities. It is assumed that about 40 % of this will be spent on the gasifying section (including drying of the coal) and 60 % on the gas cleaning section.

This means that, at a desired capacity of 200 MW, the fixed capital costs according to Lurgi A.G. will be 144 million DM, or considerably lower if the drying section already exists.

#### 9.1.2 Taylor's estimate for the fixed capital costs

Taylor's method is a "step" method. It makes use of a costliness index, which is found by using a scoring system. Each item of process equipment is attributed a certain score, in which throughput, material of construction, reaction time as well as pressure and temperature extremes are taken into account.

It was assumed that the gasifier section will be constructed from stainless steel, which is believed to be realistic because of the corrosive nature of the lignite fuel and the produced gases.

The method does not take the specific dimensions of the designed process equipment into consideration.

For the total costliness index a value of 31.15 was found (appendix V.1.1). This resulted in an estimated fixed capital cost of 151 million DM for the gasifying section, closely approaching the Lurgi estimate.

#### 9.1.3 Wilson's estimate for the fixed capital costs

Wilson's method is also a "step" method, which is based on one average unit cost. This average unit cost is assumed to be valid for every set of equipment of roughly equal size.

This method resulted in an estimated investment of £ 12.8 million in base year 1971 (appendix V.1.2). This was corrected using the CPE price index [9] to being approximately £ 79 million in 1993. At the current exchange rate (November 1993) 1 DM equals £ 0.41, resulting in the fixed capital being estimated at 192 million DM.

It is clear that the large uncertainties which were introduced in achieving this estimate (e.g. the extrapolation from the base year 1971) make this result a thoroughly unreliable one.

### 9.1.4 Lang's estimate for the fixed capital costs

The method proposed by Lang [15] is a factorial method which takes the specific details of the equipment which has been designed into account. According to Coulson & Richardson [9] it can be used to make a quick estimate of capital cost in the early stages of project design, when the preliminary flow sheets have been drawn up and the main items of equipment are roughly sized. This made Lang's method ideal for our purposes. The calculations resulted in an estimated fixed capital cost for the HTW installation of DM 121 million (Appendix V.1.3).

## 9.2 Costs dependent on production volume

The costs dependent on production volume are those costs which originate in the raw materials needed for producing a fixed amount of product, in our case the synthesis gas used to generate electricity.

The costs were calculated based on the price data supplied by Brennstoff Kombinat Vresova A.G. The net production of steam by the gasifier plant was not taken into account as a financial benefit.

The calculations (appendix V.2) result in a cost of 624 Czech Crowns per ton of product gas, which equals DM 37.10 per ton of product gas.

If the plant were to run continuously (365 days per year, 24 hours per day) it would produce a maximum of 1.8 million tons of synthesis gas. However, it is envisaged to operate the HTW gasifier only during peak hours (approximately 8 hours per day).

Then, on an annual basis, the costs dependent on the production volume will be a maximum of 378 million Czech Crowns, which equals DM 22.5 million.

During operation of the plant, a maximum of 354 MW of electricity can be produced when the net produced steam from the gasifier section is also used to drive the steam turbine, and the coal dust is burnt in the existing coal combustion plant.

This electricity has a market value of 3000 Czech crowns per MWh, resulting in a turnover of 3099 million Crowns (or DM 184 million) annually.

## 9.3 Rentability, return on investment (ROI) and internal rate of return (IRR)

These were considered irrelevant, since the design concerns the replacement of part of an existing facility (which has to remain within a given budget) rather than a completely new venture.

Furthermore, the data which would be needed for such calculations (labour costs, taxes etc.) are not available and are set to change rapidly in the current economic climate.

## 9.4 Evaluation of the cost estimation

### 9.4.1 Fixed capital costs

Fixed capital costs are estimated to be between 121 and 151 million Deutsch Marks, the lower estimate of which seems to be the more reliable since it was found using Lang's method. This factorial method takes the specific details of the designed equipment into account.

The Taylor method with which the higher value was found is known to produce relatively high costs, and is a step method which does not make use of the dimensions of the equipment.

The Wilson method finally, yields an excessively high result (DM 192 million) but is considered to be outdated and unreliable.

The estimates also seem to agree with the Lurgi estimate of DM 144 million, which includes a new system for reducing the moisture content of the lignite to 12 wt%.

For correctness it has to be stated that investments will also be needed for matters such as auxiliary equipment, start-up, supplies and many other things which are not incorporated in the above estimates.

However, based on the crude estimates made, it seems likely that the replacement of the existing Lurgi gasifiers by a HTW is feasible.

More detailed cost studies using more reliable and accurate measurements are however recommended.

### 9.4.2 Costs dependent on production volume

The costs dependent on the production volume which were determined show clearly that these costs are well below the current market value of the generated electricity. However the unknown production costs caused by operation of the Rectisol plant still have to be added.

The calculated production costs do not incorporate such things as wages, overhead etc. It is virtually impossible to make a reliable estimate for these costs in the current Czech economy, which is in rapid transition from centralised to market-oriented.

Furthermore, the situation faced by the Brennstoff Kombinat is quite exceptional. Amongst others it has to be considered that the company has its own secure supply of cheap lignite as well as a secured market.

The replacement of the current Lurgi gasifiers was brought about by technical and environmental necessity. The economy of the project therefore mainly depends on the budget available within the company for the replacement.

Common sense dictates that the modern HTW gasifier will be cheaper to staff, maintain and operate than the existing ageing Lurgi gasifiers. The real running costs can only be provided by Rheinbraun and Lurgi A.G.

## 10 The Vresova project compared to the KoBra project.

The Vresova project as described in this report shows a considerable similarity with the KoBra project. The KoBra project involves a 300 MW Integrated Coal Gasification Combined Cycle power plant to be built in Goldenberg, Germany [16].

Table 7: The Vresova project compared to the KoBra project.

Characteristics	Vresova <sup>1</sup>	KoBra
Power output gasifier (MW)	342 (193)	312
Additional output attributable to the combustion of the bottom product dust in the boilers (MW)	12 (7)	27
Coal throughput (t/h)	127 (72)	160
Fuel energy (NCV) (MJ/s)	808 (457)	748
Efficiency (NCV) (%)	43.8	45.3
Gasification pressure (bar)	24 (27)	27
Gasification temperature (°C)	1100	1020
Max. diameter crushed coal (mm)	5	6
Gasifying medium	steam/O <sub>2</sub>	steam/air
Raw gas throughput (kg/s)	58 (33)	140
Carbon conversion rate (%)	96	91
Steam pressures used (bar)	35/24/5	128/37

<sup>1</sup> The values between parentheses have been standardised to a power output of 200 MW (193 MW of the gasifier and 7 MW due to combustion of the coal dust).

## 11 Conclusions

After comparison of various types of coal gasifiers and commercial processes the pressurized High Temperature Winkler gasification system turned out to be the most suitable for North Bohemian Lignite. Replacing the 26 existing Lurgi moving bed reactors by one HTW fluidised bed gasifier will eliminate the production of undesirable compounds such as tars and phenols and provide the required flexibility for the production of electricity in peak hours. As the Brennstoff Kombinat A.G. has a lack of space for new construction activities, an additional bonus of the HTW system is the large area which will become available after the replacement.

The gasifier was designed to produce approximately 280,000 m<sup>3</sup>/hr of raw gas at 2,4 MPa. This resulted in a capacity of 340 MW, considerably more than the 200 MW the Brennstoff Kombinat had expected.

It is expected that there will be no problems integrating the new gasification system and the current Rectisol process from both a process engineering and environmental point of view. The latter aspect will be checked by the Brennstoff Kombinat A.G.

Using Lang's and Taylor's method the fixed capital costs were estimated to be between 121 and 151 million DM, which agrees with the Lurgi estimate of 144 million DM. Although not all costs are known it seems that the production costs will be well below the earnings with the current market value of the generated electricity.

Based on the results of this report replacing the Lurgi gasifiers by a High Temperature Winkler gasification system is an obvious choice.

## 12 Recommendations

- A gasification pressure of 27 bar (like in the KoBra project) instead of 24 bar, because then the syngas can enter the gas turbines at the ideal pressure of 23 bar (this is technically possible).
- An overpressure of 3 bar for the steam and oxygen feed compared to the gasification pressure (this means 30 bar with the first recommendation followed).
- Burn the coal dust in the nearby coal-fired boiler.
- Shut one Rectisol train down and don't expect two trains anywhere in the process to be economically feasible: "Moreover, the estimated investment costs also showed that a commercial plant can only be operated economically (using a gas turbine of the 200 MW class) if units such as coal drying, gasification, gas cooling and treatment are designed as single train units" [16].

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## Appendix II: Design of equipment for HTW gasification

### II.1: Physical constants used in the design calculations

The relations used and approximations used for calculation of densities, viscosities and specific heats for gases at different pressures and temperatures are listed in this appendix.

#### Densities

Considering the high pressures and temperatures in the system the ideal-gas law is not adequate. Therefore a two parameter equation of state, the van der Waals equation [17] was used:

$$P = \frac{R \cdot T}{V - b} - \frac{a}{T^{0.5} \cdot V \cdot (V + b)}$$

where	P:	Pressure	(Pa)
	R:	Gas constant = 8314.33	(Pa·m <sup>3</sup> ·kmole <sup>-1</sup> ·K <sup>-1</sup> )
	T:	Temperature	(K)
	V:	Volume	(m <sup>3</sup> ·kmole <sup>-1</sup> )
	b:	Constant parameter for a gas [17]	(m <sup>3</sup> ·kmole <sup>-1</sup> )
	a:	Constant parameter for a gas [17]	(K <sup>0.5</sup> ·m <sup>6</sup> ·kmole <sup>-2</sup> ·Pa)

The density is calculated as follows:

$$\rho = \frac{M}{V}$$

where	ρ:	Density	(kg·m <sup>-3</sup> )
	M:	Molecular mass	(kg·kmole <sup>-1</sup> )
	V:	Volume	(m <sup>3</sup> ·kmole <sup>-1</sup> )

#### Viscosities

Due to the lack of data and relations for calculating the temperature dependency of the viscosity of gases the viscosity of air at 20 atm. which is labelled from 300 - 2800 K [18] is used as an approximation.

**Specific heats**

The specific heat of coal varies between 1.09 and 1.55  $\text{kJ}\cdot\text{kg}^{-1}\cdot\text{K}^{-1}$  [17]. For the design calculations  $C_p = 1.3 \text{ kJ}\cdot\text{kg}^{-1}\cdot\text{K}^{-1}$  was used.

The specific heat of the ash was calculated as the weight average of the specific heats of the ash components, which were calculated with the following relations [17]:

$$C_{p, SiO_2} = 10.87 + 0.008712 \cdot T - \frac{241200}{T^2}$$

$$C_{p, Al_2O_3} = 22.08 + 0.008971 \cdot T - \frac{522500}{T^2}$$

$$C_{p, Fe_2O_3} = 24.72 + 0.01604 \cdot T - \frac{423400}{T^2}$$

where  $C_p$ : Specific heat (kcal $\cdot\text{K}^{-1}\cdot\text{kmole}^{-1}$ )

For the calculation of the mean heat capacity of gases the following equation was used [19]:

$$\frac{C_{p,mh}}{R} = A + B \cdot T_{am} + \frac{C}{3} \cdot (4 \cdot T_{am}^2 - T_1 \cdot T_2) + \frac{D}{T_1 \cdot T_2}$$

where  $C_{p,mh}$ : Specific heat (kJ $\cdot\text{K}^{-1}\cdot\text{kmole}^{-1}$ )  
 $R$ : Gas constant = 8.3144 (kJ $\cdot\text{K}^{-1}\cdot\text{kmole}^{-1}$ )  
 $T_{am}$ : Arithmetic-mean temperature =  $(T_1+T_2)/2$  (K)  
 A, B, C & D are constants listed in the literature [19]

## II.2: Design of the gasifier

### II.2.1: Recalculation of the Cycle-Tempo results

The programmed assumption is: 4 percent of the coal (namely ungasified dust) goes along with the raw gas in pipe 6. The only difference in composition with the coal in the feed is the absence of the 12% moisture. The problem now is that Cycle-Tempo does not discriminate between: 1) hydrogen gas in the raw gas and bound hydrogen in the coal dust; 2) oxygen gas in the raw gas (fortunately this is, of course, absent) and bound oxygen in the coal dust; 3) nitrogen gas and bound nitrogen in the coal dust.

Therefore the following recalculation is necessary:

#### Example (Best Run):

$(100-8.52) = 91.48$  mole% of the coal is moisture-free.

Consequently: mole fraction C(s) in coal dust =  $57.52/91.48$ .

In pipe 6 2.10 mole% is C(s), so  $(91.48/57.52)*2.10 = 3.34$  mole% in pipe 6 is coal dust and  $(100-3.34) = 96.66$  mole% is raw gas.

Bound H in coal dust:  $(25.29/57.52)*2.10 = 0.9233$  mole% (pipe 6).

Hydrogen gas in pipe 6:  $(33.73-0.9233) = 32.8067$  mole% .

Hydrogen gas in raw gas:  $32.8067/0.9666 = 33.95$  mole% .

Bound N in coal dust:  $(0.38/57.52)*2.10 = 0.0139$  mole% (pipe 6).

Nitrogen gas in pipe 6:  $(1.35-0.0139) = 1.3361$  mole% .

Nitrogen gas in raw gas:  $1.3361/0.9666 = 1.38$  mole% .

CO in raw gas:  $37.17/0.9666 = 38.46$  mole%

CO<sub>2</sub> in raw gas:  $12.06/0.9666 = 12.48$  mole%

CH<sub>4</sub> in raw gas:  $1.24/0.9666 = 1.28$  mole%

H<sub>2</sub>O in raw gas:  $11.87/0.9666 = 12.28$  mole%

H<sub>2</sub>S in raw gas:  $0.15/0.9666 = 0.16$  mole%

COS in raw gas:  $0.01/0.9666 = 0.01$  mole%

Mole flow in pipe 6:  $2*1.503 \text{ kmole}\cdot\text{s}^{-1} = 3.006 \text{ kmole}\cdot\text{s}^{-1}$ .

Mole flow coal dust:  $0.0334*3.006 = 0.100 \text{ kmole}\cdot\text{s}^{-1}$ .

Mole flow raw gas:  $0.9666*3.006 = 2.906 \text{ kmole}\cdot\text{s}^{-1}$ .

Mass flow in pipe 6:  $2*29.402 \text{ kg}\cdot\text{s}^{-1} = 58.804 \text{ kg}\cdot\text{s}^{-1}$ .

Mass flow coal dust (dry;  $12.30 \text{ kg}\cdot\text{kmole}^{-1}$ ) =  $1.237 \text{ kg}\cdot\text{s}^{-1}$ .

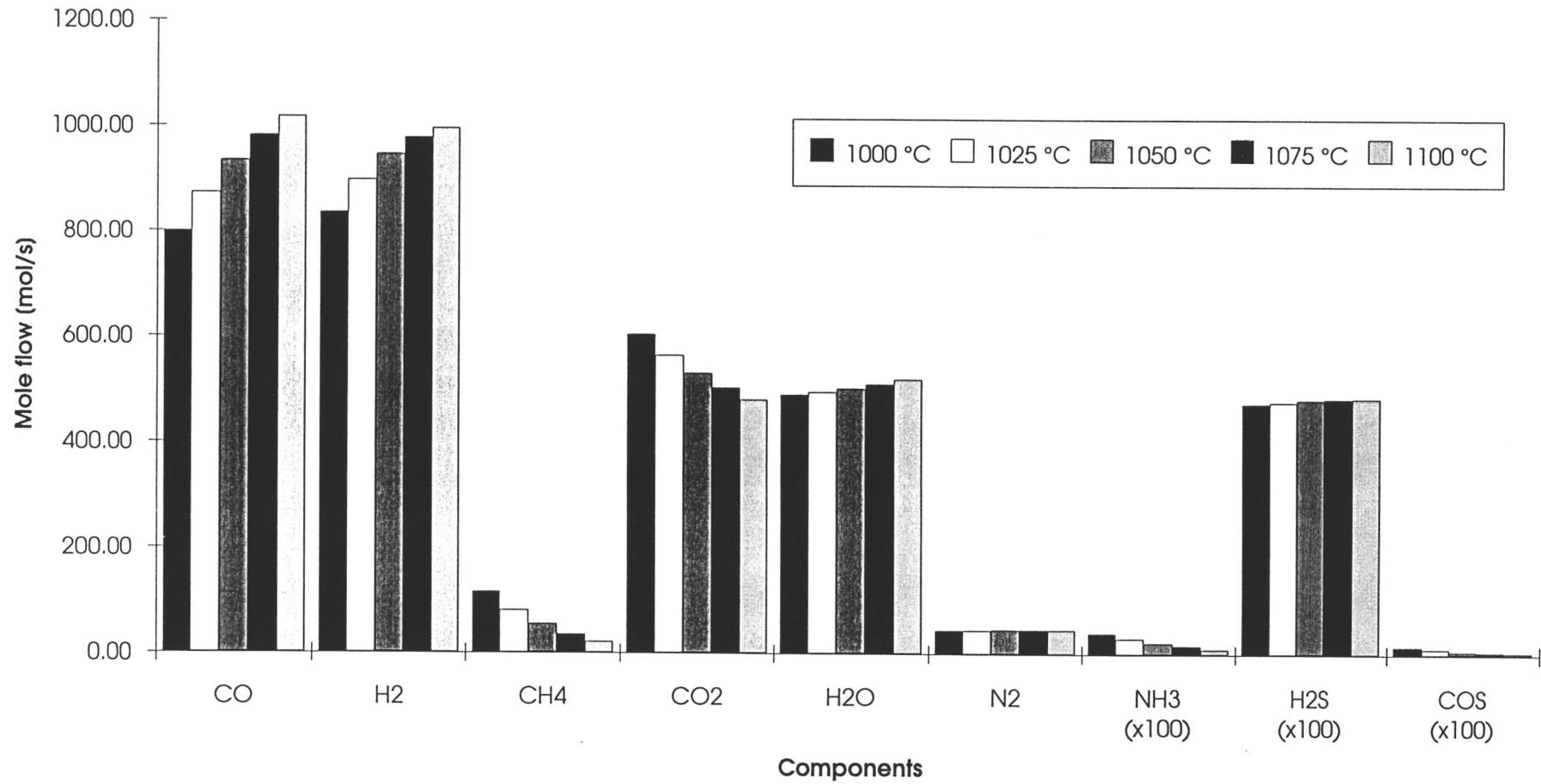
Mass flow raw gas ('wet';  $19.81 \text{ kg}\cdot\text{kmole}^{-1}$ ) =  $57.567 \text{ kg}\cdot\text{s}^{-1}$ .

**II.2.2: Optimization of the process conditions**

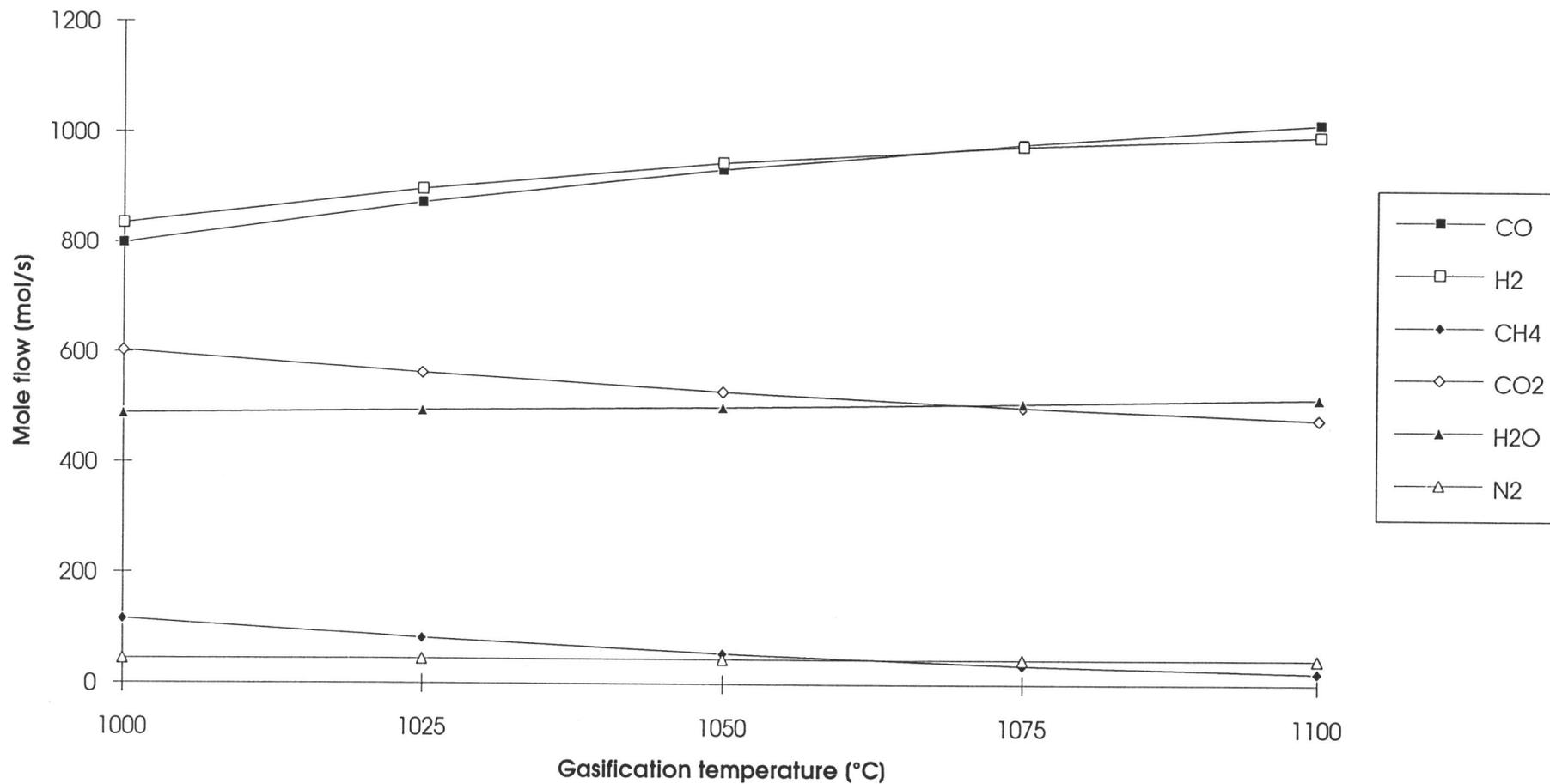
Table 8: Temperature optimization.

	Temp. (°C)	1000	1025	1050	1075	1100
Component	Total mole flow	2891.729	2959.733	3013.406	3053.028	3079.89
CO		798.67	873.08	933.69	981.20	1017.11
H2		835.01	897.45	945.49	976.75	994.32
CH4		115.59	81.43	54.18	34.46	20.94
CO2		603.45	563.42	529.45	501.64	479.46
H2O		489.21	494.56	500.81	509.19	518.26
N2		44.54	44.63	44.70	44.76	44.83
NH3 (x100)		38.66	29.01	21.29	15.49	9.68
H2S (x100)		472.77	476.63	480.50	482.43	483.40
COS (x100)		14.50	10.64	6.77	4.84	3.87
						(mol/s)
Pressure energy		22.67	23.22	23.64	23.95	24.16
Heat of combustion		567.88	576.33	583.01	587.83	590.97
Sensible heat		93.20	96.51	99.74	102.88	105.93
						(MW)
Theoretical cooling		87.96	81.84	77.1	72.78	69.28
Heat loss		21.76	20.54	19.86	19.38	19.06
Required cooling		66.2	61.3	57.24	53.4	50.22
						(MW)

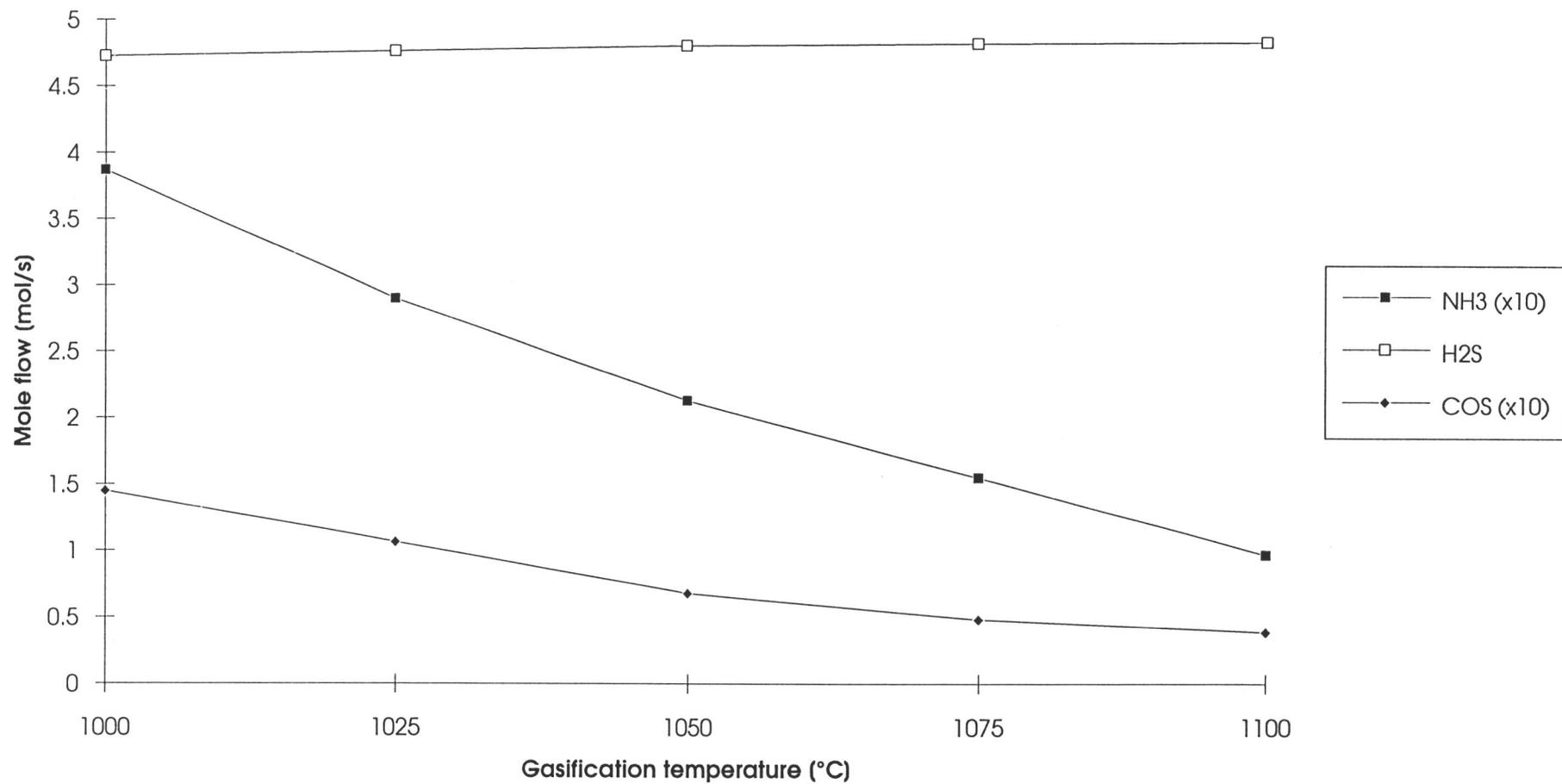
Effect of the gasification temperature on the synthesis gas composition



Effect of the gasification temperature on the synthesis gas composition



Effect of the gasification temperature on the synthesis gas composition



### Effect of the gasification temperature on the heat flow

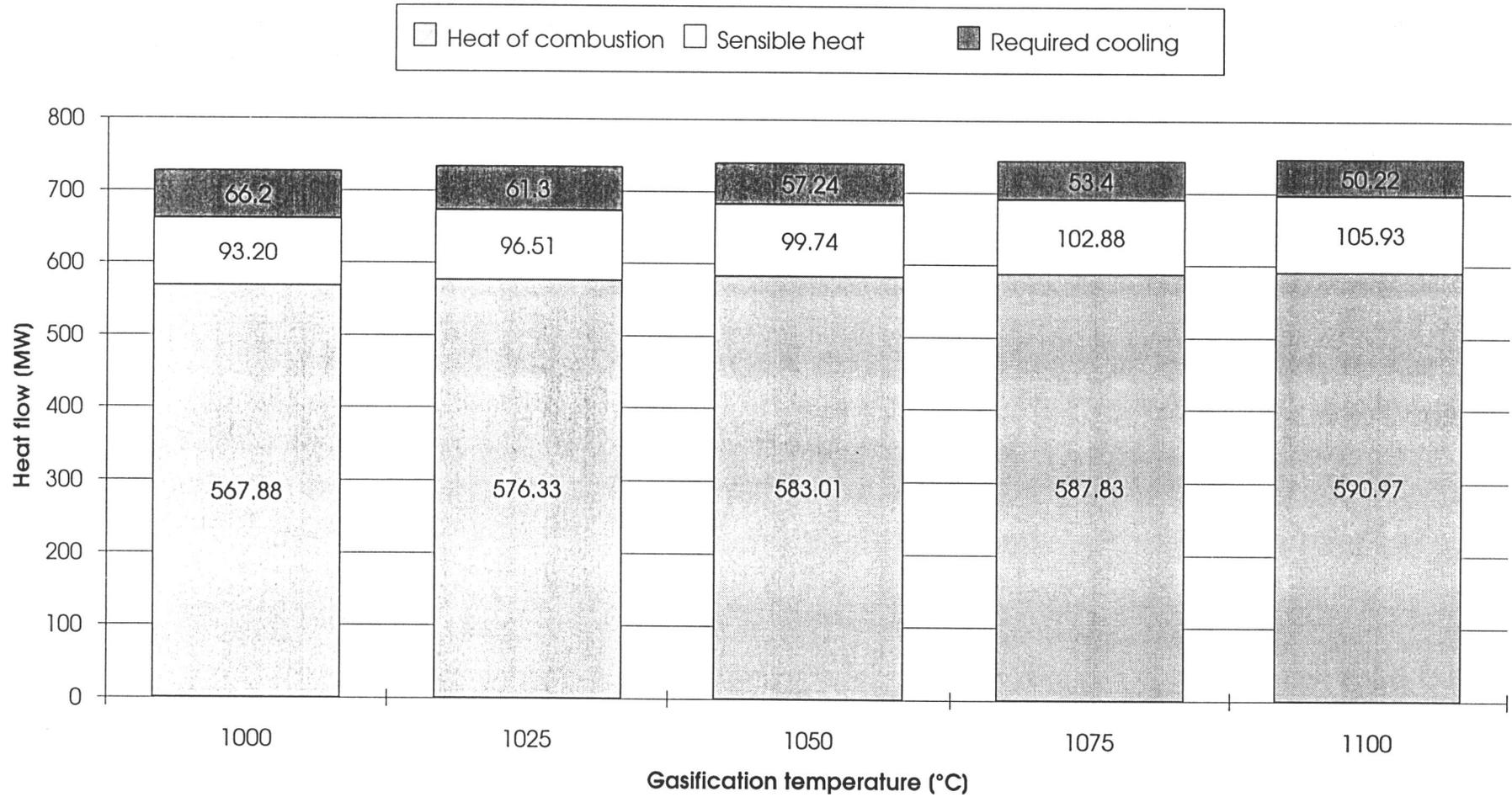
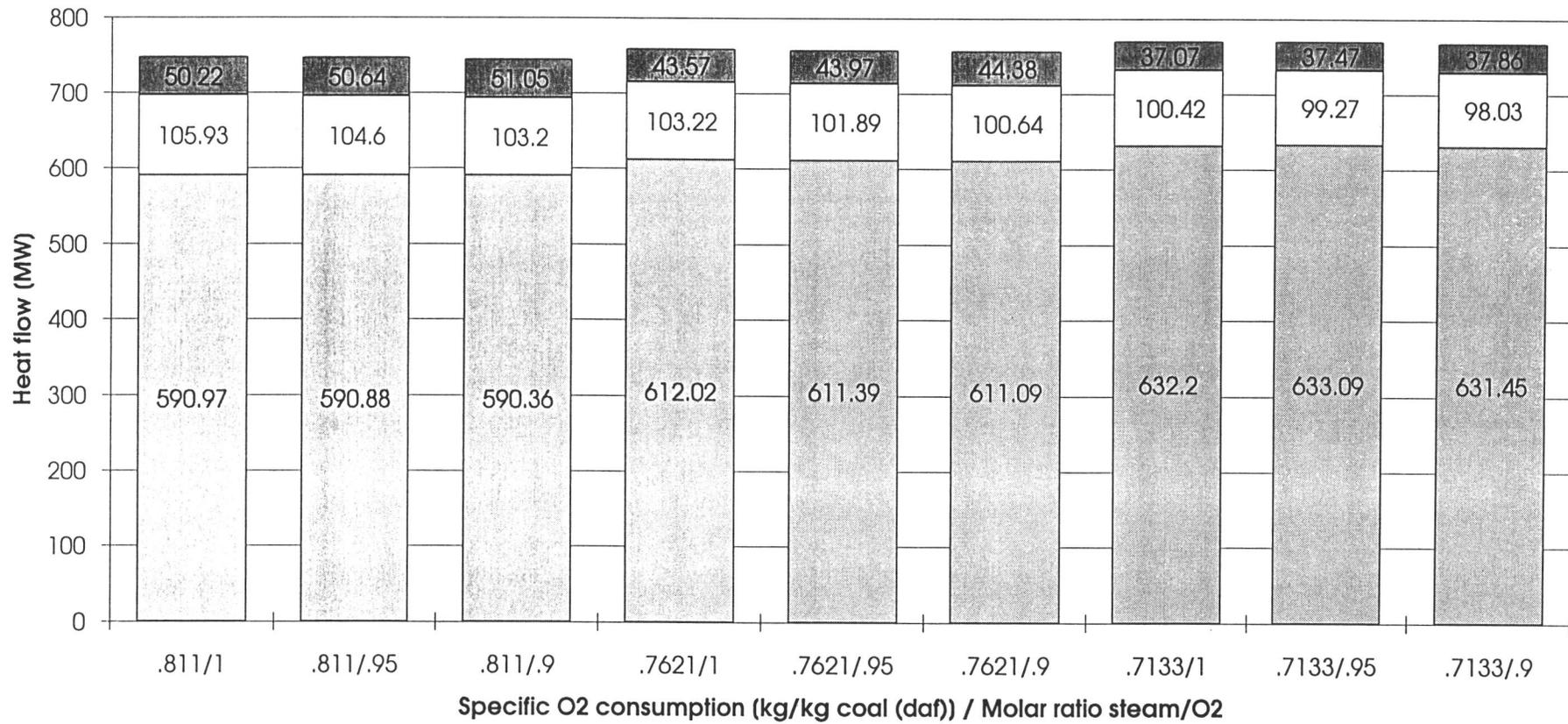
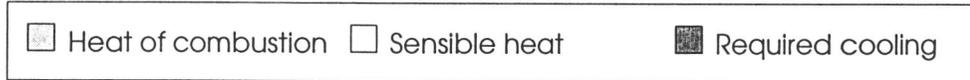


Table 9: Steam/oxygen ratio and rate optimization.

	(kg O <sub>2</sub> / kg coal(daf))									
	Spec. ox. cons.	.811/1	.811/.95	.811/.9	.7621/1	.7621/.95	.7621/.9	.7133/1	.7133/.95	.7133/.9
	Steam/O <sub>2</sub> ratio (mol st / mol O <sub>2</sub> )	1	0.95	0.9	1	0.95	0.9	1	0.95	0.9
Component	Total mole flow	3079.89	3045.898	3009.727	3028.169	2993.906	2961.723	2971.865	2942.061	2909.777
CO		1017.11	1026.92	1035.9	1058.95	1067.73	1077.17	1099.45	1108.61	1117.2
H <sub>2</sub>		994.32	982.15	968.49	1006.91	993.24	980.11	1013.1	1000.22	986.16
CH <sub>4</sub>		20.94	21.62	22.57	27.23	28.12	29	35.02	36.13	37.19
CO <sub>2</sub>		479.46	469	458.36	431.14	420.63	410.75	382.5	372.77	362.62
H <sub>2</sub> O		518.26	496.41	474.61	454.14	434.39	414.89	392	374.53	356.81
N <sub>2</sub>		44.83	44.82	44.81	44.81	44.8	44.8	44.79	44.79	44.78
NH <sub>3</sub> (x100)		9.68	10.54	11.39	11.62	12.33	13.04	13.27	13.84	14.41
H <sub>2</sub> S (x100)		483.4	477.29	471.18	469.51	464.42	459.33	457.48	453.4	449.33
COS (x100)		3.87	9.98	16.09	17.76	22.85	27.94	29.79	33.87	37.94
	(mol/s)									
Difference in steam&O <sub>2</sub>		-14.04	-11	-7.96	-9.7	-6.84	-3.98	-5.36	-2.68	0
Heat of combustion		590.97	590.88	590.36	612.02	611.39	611.09	632.2	633.09	631.45
Sensible heat		105.93	104.6	103.2	103.22	101.89	100.64	100.42	99.27	98.03
Pressure energy		24.06	23.79	23.51	23.64	23.39	23.12	23.19	22.96	22.72
	(MW)									
Theoretical cooling		69.28	69.72	70.15	60.55	60.98	61.41	52.01	52.44	52.87
Heat loss		19.06	19.08	19.1	16.98	17.01	17.03	14.94	14.97	15.01
Required cooling		50.22	50.64	51.05	43.57	43.97	44.38	37.07	37.47	37.86



### Effect of the oxygen and steam consumption on the heat flow



### II.2.3: Calculation of the gasifier dimensions

The height of the fluidisation zone is calculated as follows:

$$H_f = \frac{\ln \frac{\epsilon_{se} - \epsilon_s^*}{\epsilon_{sd} - \epsilon_s^*}}{a}$$

where	$H_f$ :	Height of the fluidisation zone	(m)
	$\epsilon_{se}$ :	Porosity at the the end of the reactor = $1.0 \cdot 10^{-4}$	(-)
	$\epsilon_{sc}$ :	Porosity in the pipe leading to the cyclone = $5.0 \cdot 10^{-5}$	(-)
	$\epsilon_{sd}$ :	Porosity at the end of the fluidised zone = 0.15	(-)
	$a$ :	Decay factor = 3.5	( $m^{-1}$ )

The determination of the total gasifier height is done as follows:

$$H_t = L_{mf} \cdot (1 - \epsilon_{mf}) - \frac{\epsilon_{sd} - \epsilon_{se}}{a} + H_f \cdot \frac{\epsilon_{sd} - \epsilon_s^*}{\epsilon_s^*}$$

where	$H_t$ :	Total height of the HTW fluidised bed gasifier	(m)
	$L_{mf}$ :	Minimal fluidisation height = 2.6	(m)

To calculate the absolute recirculation flow the following relation is used:

$$G_{s, absolute} = \rho_s \cdot u_t \cdot \epsilon_s^*$$

where	$G_{s, absolute}$ :	Absolute solid recirculation mass flow rate	( $kg \cdot s^{-1}$ )
	$\rho_s$ :	Density of the lignite = 1200	( $kg \cdot m^{-3}$ )
	$u_t$ :	Terminal particle velocity = 2.13	( $m \cdot s^{-1}$ )

To calculate the minimal wall thickness the following relation is used:

$$e = \frac{P_i \cdot D_i}{2 \cdot f - P_i}$$

where,	$e$ :	Minimal wall thickness	(m)
	$P_i$ :	Internal pressure = 30	(Pa)
	$D_i$ :	Internal diameter = 2.75	(m)
	$f$ :	Design stress = $60 \cdot 10^6$	(Pa)

Dimensioning a HTW-gasifier fed by Vresova lignite on the basis of this simplified model leads to a Gasifier with an internal diameter of 2.75 m, a fluidisation height of 2 m, a total bed height of 12 m and a absolute solid recirculation flow of  $1.52 \text{ kg} \cdot \text{s}^{-1}$ .

The construction material used in building the gasifier must be resistant to temperatures of  $1100^\circ\text{C}$  or more, pressures up to 30 bar and resistant to both reductive and oxidative environments. Choosing the construction material and calculating the wall thickness using the method described by Coulson & Richardson [9] resulted in titanium stabilised 18Cr/8Ni stainless steel (321) and a minimal wall thickness of 75 mm (4.5 mm corrosion allowance). Because the materials produced are toxic and potentially explosive the risk of leakage and technical malfunctioning must be as low as possible.

### II.3: Bunkers

#### Calculation of height and diameter

Assume H:D = 2:1

$$V = 25 \text{ m}^3 = \pi/4 \cdot D^2 \cdot H \quad \text{--->} \quad \begin{aligned} D &= 2.5 \text{ m} \\ H &= 5 \text{ m} \end{aligned}$$

#### Design on the basis of internal pressure

*Cylindrical section*

$$e = \frac{P_i \cdot D_i}{2 \cdot J \cdot f - P_i} = 22.75 \text{ mm} + 4 \text{ mm corrosion allowance} = 27 \text{ mm}$$

where	e:	Minimal thickness required	(mm)
	P <sub>i</sub> :	Internal design pressure + 10% = 1.1 · 2.3 = 2.53	(N·mm <sup>-2</sup> )
	D <sub>i</sub> :	Internal diameter	(mm)
	J:	Joint factor = 0.85	(-)
	f:	Design stress = 165	(N·mm <sup>-2</sup> )

*Ellipsoidal end (top)*

$$e = \frac{P_i \cdot D_i}{2 \cdot J \cdot f - 0.2 \cdot P_i} = 22.5 \text{ mm} + 4 \text{ mm corrosion allowance} = 27 \text{ mm}$$

*Conical end (bottom)*

$$e = \frac{P_i \cdot D_c}{2 \cdot J \cdot f - P_i} \cdot \frac{1}{\cos \alpha} = 45.5 \text{ mm} + 4 \text{ mm corrosion allowance} = 50 \text{ mm}$$

where	α:	Half the cone apex angle = 60°	(-)
	D <sub>c</sub> :	Internal diameter cone	(mm)

Assume D<sub>c</sub> = D<sub>i</sub> to obtain the maximum value of e.

#### Check on the basis of stress analysis

*Weight load*

For the weight of a steel vessel with two conical ends the weight is given by:

$$W_v = 240 \cdot C_v \cdot D_m \cdot (H_v + 0.8 \cdot D_m) \cdot t$$

where	t:	Wall thickness	(-)
	D <sub>m</sub> :	D <sub>i</sub> + t · 10 <sup>-3</sup>	(m)
	C <sub>v</sub> :	0.8 (few fittings)	(-)
	H <sub>v</sub> :	Height of vessel = 5	(m)

From which it follows that W<sub>v</sub> = 124 kN. Adding a generous safety factor of 20% for the weight of the conical section results in W<sub>v</sub> = 149 kN. The contribution of the weight of the contents (21 tons of lignite) is 21.1 · 10<sup>3</sup> · 9.81 = 207 kN. Therefore the total weight load is 356 kN.

Dead weight stress:

$$\sigma_w = \frac{W_v}{\pi (D_i + t) t} = \frac{356 \cdot 10^3}{\pi (2.5 \cdot 10^3 + 25) 25} = -1.80 \text{ N}\cdot\text{mm}^{-2}$$

compressive (at bottom of cylindrical section)

Pressure stresses:

$$\sigma_l = \frac{P \cdot D_i}{4 t} = 59 \text{ N}\cdot\text{mm}^{-2}$$

$$\sigma_h = \frac{P \cdot D_i}{2 t} = 118 \text{ N}\cdot\text{mm}^{-2}$$

The resultant longitudinal stress  $\sigma_z = \sigma_l + \sigma_w = 59 + (-1.80) = 57.2 \text{ N}\cdot\text{mm}^{-2}$  is well below the maximum allowable design stress of  $165 \text{ N}\cdot\text{mm}^{-2}$ .

Therefore the wall thicknesses are sufficient and probably somewhat larger than strictly necessary.

## II.4: Cyclones

The raw gas stream from the gasifier has a volumetric flow of  $12.7 \text{ m}^3 \cdot \text{s}^{-1}$  (24 bar,  $1000^\circ\text{C}$ ) and contains particles with a diameter smaller than or equal to  $1.2 \text{ mm} = 1240 \text{ }\mu\text{m}$ .

Gas cyclones are dimensioned with a gas inlet velocity between  $6$  and  $21 \text{ m}\cdot\text{s}^{-1}$  [17]. In practice the optimal gas velocity is  $15 \text{ m}\cdot\text{s}^{-1}$ .

The inlet area of the cyclone  $A_i$  is calculated as follows:

$$A_i = \frac{F_v}{v_i}$$

where  $F_v$ : Volumetric flow of raw gas through one cyclone (Pa·s)

In case of a square inlet the inlet diameter  $d_i = A_i^{0.5}$ .

The other dimensions of the cyclone are related to  $d_i$  as follows [20]:

- Diameter of the cyclone,  $D_c = 2 \cdot d_i$ .
- Height of the cyclone,  $H_c = 4 \cdot D_c$ .
- Outlet diameter of the cyclone,  $d_o = 0.5 \cdot D_c$ .

The cut size of the cyclone can be calculated from the following relation [20]:

$$d_{50} = \sqrt{\frac{0.14 \cdot \eta \cdot d_i}{(\rho_s - \rho_g) \cdot v_i}}$$

where  $\eta$ : Viscosity of the gas mixture =  $40 \cdot 10^{-6}$  ( $1000^\circ\text{C}$ ) (Pa·s)  
 $d_i$ : Diameter cyclone inlet (m)  
 $\rho_s$ : Density of the lignite particles = 1200 ( $\text{kg}\cdot\text{m}^{-3}$ )  
 $\rho_g$ : Density of the raw gas = 4.54 ( $\text{kg}\cdot\text{m}^{-3}$ )  
 $d_{50}$ : Cut size of the cyclone (m)

The pressure drop  $\Delta P$  over the cyclone is given by the following equation [20]:

$$\Delta P_{cyclone} = 4 \cdot \rho_f \cdot v_i^2$$

where  $\rho_f$ : Density of the gas mixture ( $\text{kg}\cdot\text{m}^{-3}$ )  
 $v_i$ : Inlet velocity of the gas mixture ( $\text{m}\cdot\text{s}^{-1}$ )

Table 10: Specifications of the cyclones in the HTW gasification installation.

	Cyclone Y <sub>1</sub>	Cyclone Y <sub>2</sub>
Number of cyclones	1	5
Volumetric flow per cyclone ( $\text{m}^3 \cdot \text{s}^{-1}$ )	12.67	2.53
Inlet velocity ( $\text{m}\cdot\text{s}^{-1}$ )	6	15
Inlet diameter (m)	1.45	0.41
Diameter cyclone (m)	2.91	0.82
Cut size ( $\mu\text{m}$ )	33.7	11.3
Pressure drop (mbar)	40.9	6.54

## II.5: Heat exchanger

The design was performed using data and figures provided by Coulson and Richardson [9]. The heat exchangers have one shell pass, two tube passes and a split ring floating head.

The heat load is calculated as follows:

$$HL = G_{gas} \cdot C_{p, gas, mean} \cdot (T_1 - T_2)$$

where	HL:	Heat load of the heat exchanger	(kW)
	$G_{gas}$ :	Mass flow rate raw gas = 57.6	(kg·s <sup>-1</sup> )
	$C_{p, gas, mean}$ :	Mean specific heat of the gas mixture	(kJ·kg <sup>-1</sup> ·K <sup>-1</sup> )
	$T_1$ :	Inlet temperature raw gas	(K)
	$T_2$ :	Outlet temperature raw gas	(K)

The cooling water mass flow  $G_{water}$  follows from the following relation:

$$G_{water} = \frac{HL}{H_{steam} - H_{water}}$$

where	$G_{water}$ :	Mass flow rate water	(kg·s <sup>-1</sup> )
	HL:	Heat load of the heat exchanger	(kW)
	$H_{steam}$ :	Enthalpy of the outgoing steam	(kJ·kg <sup>-1</sup> )
	$H_{water}$ :	Enthalpy of the incoming water	(kJ·kg <sup>-1</sup> )

The true temperature difference is given by:

$$\Delta T_m = F_t \cdot \Delta T_{lm}$$

$$R = \frac{T_1 - T_2}{t_2 - t_1}$$

$$S = \frac{t_2 - t_1}{T_1 - t_1}$$

$$\Delta T_{lm} = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln \frac{T_1 - t_2}{T_2 - t_1}}$$

where	$\Delta T_m$ :	True temperature difference	(K)
	$\Delta T_{lm}$ :	Logarithmic mean temperature difference	(K)
	$F_t$ :	Temperature correction factor (function of R and S)	
	$T_1$ :	Inlet temperature raw gas	(K)
	$T_2$ :	Outlet temperature raw gas	(K)
	$t_1$ :	Inlet temperature water	(K)
	$t_2$ :	Outlet temperature steam	(K)

An estimate for the overall heat transfer coefficient is provided by Coulson & Richardson [9],  
 $U = 250 \text{ W}\cdot\text{m}^2\cdot\text{K}^{-1}$ .

The area required for heat transfer is thus given by the following relation:

$$A = \frac{HL \cdot 10^3}{\Delta T_m \cdot U}$$

where	$\Delta T_m$ :	True temperature difference	(K)
	HL:	Heat load of the heat exchanger	(kW)
	U:	Overall heat transfer coefficient = 250	( $\text{W}\cdot\text{m}^2\cdot\text{K}^{-1}$ )
	A:	Heat-transfer area	( $\text{m}^2$ )

The number of tubes N is calculated as follows:

$$N = \frac{A}{L \cdot \pi \cdot d_o}$$

where	L:	Length of a pipe	(m)
	$d_o$ :	Outer diameter of the pipe	(m)

As the shell side fluid (water/steam) is clean a 1.25 triangular pitch is used. The bundle diameter follows from:

$$D_b = d_o \cdot (N/K_1)^{(1/n_1)}$$

where	$d_o$ :	Outer diameter of the pipe	(m)
	N:	Number of tubes	(-)
	$K_1$ :	Constant dependent on pitch [9] = 0.249	(-)
	$n_1$ :	Constant dependent on pitch [9] = 2.207	(-)

The shell bundle clearance is a function of the bundle diameter and type of heat exchanger.

The pressure drop over the heat exchanger is taken to be the average of the pressure drop calculated for inlet conditions and the pressure drop calculated for outlet conditions. These pressure drops are calculated as follows:

$$\Delta P = 4f \cdot 0.5 \cdot \rho_g \cdot v^2 \cdot \frac{L}{d_o}$$

$$4f = 0.316 \cdot Re^{-0.25}$$

$$Re = \frac{\rho_g \cdot v \cdot d_i}{\eta}$$

where	4f:	Fanning friction factor	(-)
	$\rho_g$ :	Density of the raw gas	( $\text{kg}\cdot\text{m}^{-3}$ )
	$\eta$ :	Viscosity of the raw gas	(Pa·s)
	v:	Superficial velocity of the raw gas in a pipe	( $\text{m}\cdot\text{s}^{-1}$ )
	$d_o$ :	Outer diameter of the pipe	(m)
	$d_i$ :	Inner diameter of the pipe	(m)
	L:	Length of a pipe	(m)

Table 11: Specifications and dimensions of the heat exchangers.

	Heat exchanger 1		Heat exchanger 2	
$T_{in} / T_{out}$ raw gas	1000 / 450		450 / 250	
Steam produced				
· Temperature (°C)	400		200	
· Pressure (bar)	35		5	
· Enthalpy (kJ·kg <sup>-1</sup> )	3224.2		2855.1	
$C_{p, gas, mean}$ (kJ·kg <sup>-1</sup> ·°C <sup>-1</sup> )	1.857		1.738	
Heat load (MW)	58.779		20.004	
Mass flow rate water (kg·s <sup>-1</sup> )	18.73		7.22	
$\Delta T_{lm}$ (°C)	510.3		239.9	
R	1.447		1.111	
S	0.388		0.4186	
$F_t$	0.85		0.89	
$\Delta T_m$ (°C)	433.7		213.5	
Heat transfer area (m <sup>2</sup> )	542.1		374.8	
Tubes				
· Inner diameter (mm)	16		16	
· Outer diameter (mm)	20		20	
· Length (m)	4.83		4.83	
· Number of tubes (-)	1786		1235	
Bundle diameter (m)	1.117		0.945	
Bundle diametrical clearance (mm)	75		71	
Shell diameter (m)	1.192		1.016	
Pressure drop calculation	1000 °C	450 °C	450 °C	250 °C
· Volumetric flow raw gas (m <sup>3</sup> ·s <sup>-1</sup> )	12.67	7.20	7.20	5.21
· Superficial velocity gas (m·s <sup>-1</sup> )	35.28	20.04	28.98	20.96
· Density raw gas (kg·m <sup>-3</sup> )	4.54	8.00	8.00	11.05
· Viscosity raw gas (Pa·s)	40e-6	33e-6	33e-6	27e-6
· Reynolds	64088	77683	112344	137309
· Pressure drop (bar)	0.169	0.092	0.175	0.120
Average pressure drop (bar)	0.131		0.148	

## II.6 Scrubbing section

### Pressure drop from gasifier until scrubbing section

Pressure drop in the gasifier (approx.):	0.20 bar
Pressure drop due to gas velocity increase:	0.23 bar
Pressure drop in pipes (l = 50m; 0.02% / m: 1%):	0.24 bar
Pressure drop in cyclones:	0.05 bar
Pressure drop in heat exchangers:	0.28 bar
-----+-----	
Total pressure drop from gasifier until quench:	1.00 bar

### Quench and scrubbing tower

Mass flow syngas:	57.567 kg·s <sup>-1</sup>
Molar mass syngas:	19.81 kg·kmole <sup>-1</sup>

Quench:	23 bar; 12.28% H <sub>2</sub> O ---> t <sub>sat</sub> = 131°C [19].
Scrubbing tower:	22 bar; 12.28% H <sub>2</sub> O ---> t <sub>sat</sub> = 129°C [19].

The goal of the calculation is to determine the mass flow of coolant water with which the design temperatures of the gas and water are reached.

$$G_{water} = \frac{G_{gas} \cdot C_{p, gas, mean} \cdot (T_{gas, inlet} - T_{gas, out})}{C_{p, water} \cdot (T_{water, out} - T_{water, inlet})}$$

where	$G_{water}$ :	Mass flow rate water	(kg·s <sup>-1</sup> )
	$G_{gas}$ :	Mass flow rate raw gas = 57.6	(kg·s <sup>-1</sup> )
	$C_{p, gas, mean}$ :	Mean specific heat of the gas mixture	(kJ·kg <sup>-1</sup> ·K <sup>-1</sup> )
	$C_{p, water}$ :	Specific heat of water	(kJ·kg <sup>-1</sup> ·K <sup>-1</sup> )

The maximum velocity in contact equipment in practice is given by the following relation [12]:

$$v_{max} = 0.1 \cdot \sqrt{\frac{\rho_l - \rho_g}{\rho_g}}$$

where	$v_{max}$ :	Maximum velocity	(m·s <sup>-1</sup> )
	$\rho_l$ :	Density of the water	(kg·m <sup>-3</sup> )
	$\rho_g$ :	Density of the raw gas	(kg·m <sup>-3</sup> )

Table 12: Specifications and dimensions of the quench and the scrubbing tower.

	Quench (23 bar)	Scrubbing tower (22 bar)
$G_{\text{water}}$ ( $\text{kg}\cdot\text{s}^{-1}$ )	113.04	113.49
$C_{p, \text{water}}$ ( $\text{kJ}\cdot\text{kg}^{-1}\cdot\text{K}^{-1}$ ) [21]	4.190	4.185
$C_{p, \text{gas, mean}}$ ( $\text{kJ}\cdot\text{kg}^{-1}\cdot\text{K}^{-1}$ )	1.7001	1.6913
$T_{\text{water, in}}$ ( $^{\circ}\text{C}$ )	90.41	90
$T_{\text{water, out}}$ ( $^{\circ}\text{C}$ )	115	90.41
$T_{\text{gas, in}}$ ( $^{\circ}\text{C}$ )	250	131
$T_{\text{gas, out}}$ ( $^{\circ}\text{C}$ )	131	129
$\rho_l$ ( $\text{kg}\cdot\text{m}^{-3}$ )	963	971
$\rho_g$ ( $\text{kg}\cdot\text{m}^{-3}$ )	12.013	12.998
$F_{\text{water}}$ ( $\text{m}^3\cdot\text{s}^{-1}$ )	0.117	0.117
$F_{\text{gas}}$ ( $\text{m}^3\cdot\text{s}^{-1}$ )	4.792	4.429
$F_{\text{total}}$ ( $\text{m}^3\cdot\text{s}^{-1}$ )	4.909	4.546
Residence time (s)	5.62	11.65
$v_{\text{max}}$ ( $\text{m}\cdot\text{s}^{-1}$ )	0.890	0.859
D (m)	2.65	2.60
H (m)	5	10

### Venturi scrubber

The mass flows of  $113.49 \text{ kg}\cdot\text{s}^{-1}$  through the scrubbing tower and  $113.04 \text{ kg}\cdot\text{s}^{-1}$  through the quench result in a mass flow of  $0.45 \text{ kg}\cdot\text{s}^{-1}$  through the Venturi scrubber.  
The assumed pressure drop is 1 bar (23 bar inlet and 22 bar outlet pressure).

## Pump

Absolute pressure head: 16.5 m  
of which  
- Desired pressure increase: 10.2 m (1 bar);  
- Static height difference: 5.0 m;  
- Pressure drop over pipe: 1.3 m (0.13 bar).

Calculation pressure drop over pipe with internal diameter  $d$ :

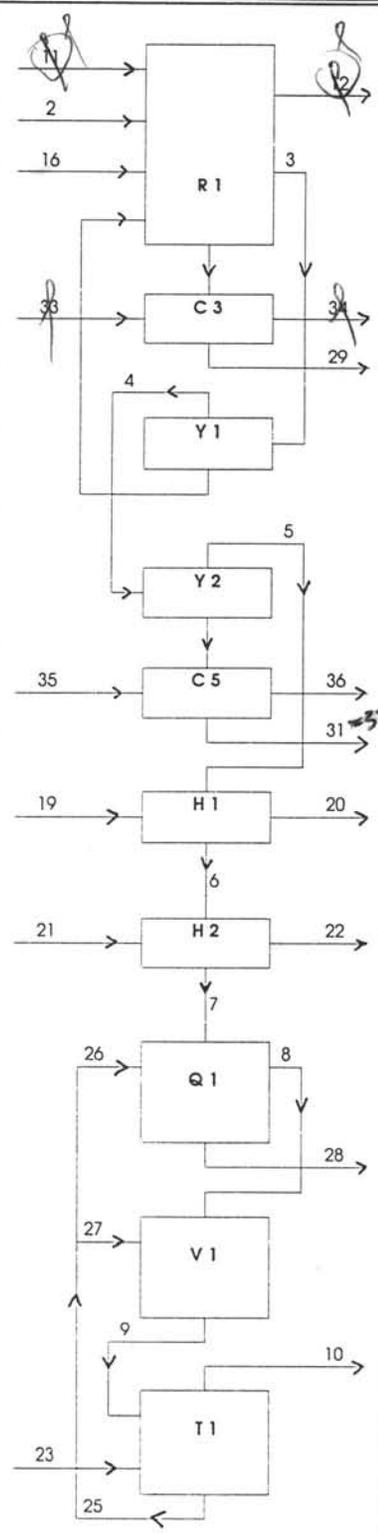
Mass flow water:	113.5	$\text{kg}\cdot\text{s}^{-1}$
Density water:	971.5	$\text{kg}\cdot\text{m}^{-3}$ (90°C, 22.5 bar)
Volume flow water:	0.1168	$\text{m}^3\cdot\text{s}^{-1}$
Dynamic viscosity water:	0.318	$\text{mPa}\cdot\text{s}$
Reynolds number:	$4.54\cdot 10^5 / d$	
Pipe length:	20	m
Friction factor from [21]		
Pressure drop formula from [22]		

Choose  $d=0.1\text{m}$  -->  $\text{Re} = 4.54\cdot 10^6$  -->  $f = 0.0020$  -->  $\Delta p = 1.7$  bar;  
Choose  $d=0.2\text{m}$  -->  $\text{Re} = 2.27\cdot 10^6$  -->  $f = 0.0024$  -->  $\Delta p = 0.13$  bar.

Using an even wider pipe is not advisable because of the very small percentual decrease of the needed pump boosting height and the quadratically rising material costs.

### Appendix III: Molar and heat balances

IN			OUT		
M [kmole/s]	M [kg/s]	Q [kW]	M [kmole/s]	M [kg/s]	Q [kW]
0.964	17.37 cooling water (11)	1,496.0	0.964	17.37 steam out of gasifier (12)	54,374.2
2.784	35.14 coal feed (2)	102,852.4	0.063	1.13 steam (34)	3,643.3
1.126	28.60 O <sup>2</sup> /steam feed (16)	118,766.0	0.066	4.93 ash (29)	69,270.0
0.063	1.13 water (35)	97.3			
0.023	0.41 water (35)	35.3	0.023	0.41 steam (36)	1,321.9
1.040	18.73 water (19)	1,613.0	6.300	1.96 dust (31)	5,237.8
0.401	7.22 water (21)	621.8	1.040	18.73 steam (20)	60,389.3
			0.401	7.22 steam (22)	20,613.8
			6.275	113.04 waste water (28)	54,541.8
6.300	113.49 water (23)	42,744.4	2.906	57.57 product gas (10)	158,805.0
		808,220.0			600,340.3
					15,010.0
	222.19	1,076,476.2		222.19	1,076,476.2



Molar and heat balances

Table 13: Molar and heat balances, part 1.

*feed in* *feed in*

Stream number	2	4	5	6	7	8	9
Components	Molar flow (kmole·s <sup>-1</sup> )						
C (s)	1.581	0.063	0	0	0	0	0
H <sub>2</sub>	0.695	1.014	0.986	0.986	0.986	0.986	0.986
O <sub>2</sub>	0.154	0.0060	0	0	0	0	0
H <sub>2</sub> O	0.234	0.357	0.357	0.357	0.357	0.357	0.357
N <sub>2</sub>	0.010	0.041	0.041	0.041	0.041	0.041	0.041
S	0.0044	0	0	0	0	0	0
SiO <sub>2</sub> (s)	0.0495	0.002	0	0	0	0	0
Al <sub>2</sub> O <sub>3</sub> (s)	0.018	0.0006	0	0	0	0	0
Fe <sub>2</sub> O <sub>3</sub> (s)	0.0022	0	0	0	0	0	0
SO <sub>3</sub>	0.0005	0	0	0	0	0	0
CH <sub>4</sub>	0	0.0370	0.037	0.037	0.037	0.037	0.037
H <sub>2</sub> S	0	0.005	0.005	0.005	0.005	0.005	0.005
CO	0	1.12	1.12	1.12	1.12	1.12	1.12
CO <sub>2</sub>	0	0.363	0.363	0.363	0.363	0.363	0.363
COS	0	0.0003	0.0002	0.0002	0.0002	0.0002	0.0002
NH <sub>3</sub>	0	0	0	0	0	0	0
Total	2.748	3.006	2.906	2.906	2.906	2.906	2.906
	Heat flow (kW)						
Sensible	102,852.4	258,084.9	249,464.8	190,685.8	170,681.8	159,035.3	158,999.7
Combustion	808,220.0	633,269.1	600,940.3	600,940.3	600,940.3	600,940.3	600,940.3
Total	911,072.4	891,354.0	850,405.1	791,626.1	771,622.1	759,975.6	759,940.0

Molar and heat balances

Table 14: Molar and heat balances, part 2.

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Stream number	10	11	12	16	19	20	21	22
Components	Molar flow (kmole s <sup>-1</sup> )							
C (s)	0	0	0	0	0	0	0	0
H <sub>2</sub>	0.986	0	0	0	0	0	0	0
O <sub>2</sub>	0	0	0	0.604	0	0	0.401	0.401
H <sub>2</sub> O	0.357	0.964	0.964	0.522	1.040	1.040	0	0
N <sub>2</sub>	0.041	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0	0
SiO <sub>2</sub> (s)	0	0	0	0	0	0	0	0
Al <sub>2</sub> O <sub>3</sub> (s)	0	0	0	0	0	0	0	0
Fe <sub>2</sub> O <sub>3</sub> (s)	0	0	0	0	0	0	0	0
SO <sub>3</sub>	0	0	0	0	0	0	0	0
CH <sub>4</sub>	0.037	0	0	0	0	0	0	0
H <sub>2</sub> S	0.0046	0	0	0	0	0	0	0
CO	1.12	0	0	0	0	0	0	0
CO <sub>2</sub>	0.36	0	0	0	0	0	0	0
COS	0.0002	0	0	0	0	0	0	0
NH <sub>3</sub>	0	0	0	0	0	0	0	0
Total	2.906	0.964	0.964	1.126	1.040	1.040	0.401	0.401
	Heat flow (kW)							
Sensible	158,805.0	1,495.9	54,374.2	118,765.6	1,571.2	60,350.2	605.5	20,609.2
Combustion	600,940.3	0	0	0	0	0	0	0
Total	759,745.3	1,495.9	54,374.2	118,765.6	1,571.2	60,305.2	605.5	20,609.2

Molar and heat balances

Table 15: Molar and heat balances, part 3.

*Handwritten scribbles*

*ash.*

Stream number	23	24	25	26	27	28	29
Components	Molar flow (kmole.s <sup>-1</sup> )						
C (s)	0	0	0	0	0	0	0
H <sub>2</sub>	0	0	0	0	0	0	0
O <sub>2</sub>	0	0	0	0	0	0	0
H <sub>2</sub> O	6.300	6.300	6.300	6.275	0.025	6.275	0
N <sub>2</sub>	0	0	0	0	0	0	0
S	0	0	0	0	0	0	0
SiO <sub>2</sub> (s)	0	0	0	0	0	0	0.0470
Al <sub>2</sub> O <sub>3</sub> (s)	0	0	0	0	0	0	0.0170
Fe <sub>2</sub> O <sub>3</sub> (s)	0	0	0	0	0	0	0.0020
SO <sub>3</sub>	0	0	0	0	0	0	0
CH <sub>4</sub>	0	0	0	0	0	0	0
H <sub>2</sub> S	0	0	0	0	0	0	0
CO	0	0	0	0	0	0	0
CO <sub>2</sub>	0	0	0	0	0	0	0
COS	0	0	0	0	0	0	0
NH <sub>3</sub>	0	0	0	0	0	0	0
Total	6.300	6.300	6.300	6.275	0.025	6.275	0.0660
	Heat flow (kW)						
Sensible	42,774.4	42,968.4	42,968.4	42,798.0	170.4	54,444.5	69,270.0
Combustion	0	0	0	0	0	0	0
Total	42,774.4	42,968.4	42,968.4	42,798.0	170.4	54,444.5	69,270.0

Table 16: Molar and heat balances, part 4.

Stream number	32	33	34	35	36
Components	Molar flow (kmole·s <sup>-1</sup> )				
C (s)	0.063	0	0	0	0
H <sub>2</sub>	0.028	0	0	0	0
O <sub>2</sub>	0.006	0	0	0	0
H <sub>2</sub> O	0	0.0627	0.0627	0.0228	0.0228
N <sub>2</sub>	0	0	0	0	0
S	0.0003	0	0	0	0
SiO <sub>2</sub> (s)	0.002	0	0	0	0
Al <sub>2</sub> O <sub>3</sub> (s)	0.0006	0	0	0	0
Fe <sub>2</sub> O <sub>3</sub> (s)	0	0	0	0	0
SO <sub>3</sub>	0	0	0	0	0
CH <sub>4</sub>	0	0	0	0	0
H <sub>2</sub> S	0	0	0	0	0
CO	0	0	0	0	0
CO <sub>2</sub>	0	0	0	0	0
COS	0.0001	0	0	0	0
NH <sub>3</sub>	0	0	0	0	0
Total	0.1000	0.0627	0.0627	0.0228	0.0228
	Heat flow (kW)				
Sensible	5,237.8	97.32	3,643.3	35.31	1,321.9
Combustion	32,328.8	0	0	0	0
Total	37,566.6	97.32	3,643.3	35.31	1,321.9

## Appendix IV: Safety, health and environment

### IV.1: Properties and toxicology of the components present in the process

Table 17: Properties and toxicology of the components present in the process.

Properties	CO	H <sub>2</sub>	O <sub>2</sub>	H <sub>2</sub> S	COS	NH <sub>3</sub>
molar weight	28.01	2.02	32.00	34.08	60.07	17.04
colour	colourless	colourless	colourless	colourless	colourless	colourless
flammability	flammable	flammable	not flammable	flammable	flammable	not flammable
odour	odourless	odourless	odourless	offensive odour	offensive odour	extremely pungent odour
melting point (°C)	-207	-259.18	-218.40	-85.50	-138.00	-77.70
boiling point (°C)	-191.3	-252.80	-182.96	-60.40	49.9	-33.35
density (kg·m <sup>-3</sup> )	1.250	0.0899	1.429	1.539		0.817
MAC-value	30 ppm				1200 ppm	50 ppm

**IV.2: Hazard and operability Study of the HTW-gasifier**

Constituent: HTW-fluidised bed gasifier  
 Intention: Gasification of North-Bohemian lignite

Pipe nr.: 1  
 Intention: Connecting the lignite output flow of the screw feeder with the gasifier.

Guide word	Deviation	Cause	Consequences and action
NO or NONE	Flow	Malfunctioning of the screw feeder.	The gasifier is running dry, there is no material to gasify so temperature decreases, chemistry changes, steam and oxygen are flowing to the gas treating-unit.
MORE	Flow	Failure in the screw feeder control system or failure in the pressure vessels.	Too much lignite is fed into the gasifier, fluidisation is not optimal, temperature drops and so chemistry changes. Liquid tars could be formed.
LESS	Flow	Failure in the screw feeder control system or failure in the pressure vessels.	Too less lignite is fed into the gasifier, fluidisation is not optimal, temperature increases and chemistry changes.
REVERSE	Flow	Malfunctioning of the screw feeder.	Lignite flows back into the pressure vessel-unit and pressure will increase drastically.

Pipe nr.: 2  
 Intention: Feeding the gasifier with a oxygen/steam mixture.

Guide word	Deviation	Cause	Consequences and action
NO or NONE	Flow	Malfunctioning of the compressor.	No gasifying agent, accumulation of lignite in the gasifier, temperature decrease and changing chemistry.
MORE	Flow	Malfunctioning of the compressor.	Too much gasifying agent, fluidisation velocity changes and the bed becomes instable.
LESS	Flow	Malfunctioning of the compressor.	Shortage of gasifying agent, fluidisation velocity drops, accumulation of lignite and the bed becomes instable.
REVERSE	Flow	Malfunctioning of the compressor.	Accumulation of oxygen and steam in compressor/pipe system including lignite accumulation in gasifier.

Pipe nr.: 3  
 Intention: Output of the syngas leaving the gasifier.

Guide word	Deviation	Cause	Consequences and action
NO or NONE	Flow	Blockage somewhere in cyclone-pipe system	Accumulation of syngas in the gasifier, pressure increase and destabilisation of the fluidised bed
MORE	Flow	Malfunctioning somewhere in cyclone-pipe system	Volume flow of syngas becomes too big for cyclone recycle system. Instability of the whole system.
LESS	Flow	Blockage somewhere in cyclone-pipe system	Fluidisation regime changes, accumulation of syngas in the gasifier and gasification equilibria change.
INVERSE	Flow	Blockage somewhere in cyclone-pipe system	Drastic destabilisation of the fluidised bed, pressure increase, potentially explosive environment.

Pipe nr.: 4  
 Intention: Waterflow into the gasifier shell for cooling the gasifier.

Guide word	Deviation	Cause	Consequences and action
NO or NONE	Flow	Malfunctioning of the pump	Drastic pressure and temperature increase in the gasifier. A dangerous situation develops.
MORE	Flow	Malfunctioning of the pump	Decrease of temperature in the gasifier, chemistry changes and lignite is accumulated.
LESS	Flow	Malfunctioning of the pump	Increase in temperature in the gasifier, chemistry changes and because of high temperature failure of the construction.
REVERSE	Flow	Malfunctioning of the pump	Water flow is not connected with the heat exchange system so the whole system destabilises.

Pipe nr.: 5  
 Intention: Guiding the ash flow from the gasifier to the ash screw discharger.

Guide word	Deviation	Cause	Consequences and action
NO or NONE	Flow	Malfunctioning of the ash screw discharger.	Instable fluidisation, the bed weight increases and chemistry changes.
MORE	Flow	Malfunctioning of the ash screw discharger.	
LESS	Flow	Malfunctioning of the ash screw discharger.	Instable fluidisation, the bed weight increases and chemistry changes.
REVERSE	Flow	Malfunctioning of the ash screw discharger.	Instable fluidisation, the bed weight increases and chemistry changes.

## Appendix V: Cost estimation

### V.1 Fixed capital costs

#### V.1.1 Taylor's method

Taylor's method is a "step" method. [15]  
It is based on the following formula:

$$I_B = 182 \cdot f \cdot P^{0.39} \cdot \frac{C_I}{300}$$

where	p:	Capacity of the system = 1817.7	(ktons·yr <sup>-1</sup> )
	C <sub>i</sub> :	EPE index = 427 (1993 value)	
	I <sub>B</sub> :	Fixed capital costs	(DM)
	f:	Total costliness index	

The total costliness index can be approximated with:

$$f = \sum_{i=1}^N (1.3)^{S_i}$$

where	f:	Total costliness index
	S <sub>i</sub> :	Score for the complexity of a significant process step

Alternatively, the total costliness index can be read from a table converting scores to costliness indexes.

The score for the complexity of a process step can be read from a table [15], thereby taking into account throughput, temperature and pressure extremes, materials of construction etc. It was assumed that the gasifier section will be constructed from stainless steel, which is believed to be realistic because of the corrosive nature of the lignite fuel and the produced gases.

The scores for each process step can be found in table 12. For the total costliness index a value of 31.15 was found. This resulted in an estimated fixed capital cost of 151 million DM for the gasifying section, closely approaching the Lurgi estimate.

Table 13: Calculation of costliness index (Taylor).

Storage / handling	Throughput	Construction material	Reaction / storage time	Pressure / temperature	Other	Total score	Costliness index (f)
fuel coal	-1	1	0	0	0	0	1
oxygen	-1	1	0	1.5	0	1.5	1.5
steam	-2	1	0	1.5	0	0.5	1.15
water (recycle)	1	1	0	0	0	2	1.7
water (make-up)	1	1	0	0	0	2	1.7
Process							
bunker	-1	1	0	0	0	0	1
pressure chambers	-1	1	0	1.5	0	1.5	1.5
reactor (gasifier)	0	1	0	2 + 1.5 = 3.5	1 (fluid bed)	5.5	4.25
ash lock hoppers	-3	1	0	2 + 1.5 = 3.5	0	1.5	1.5
cyclones	0	1	0	2 + 1.5 = 3.5	0	4.5	3.25
dust lock hoppers	-3	1	0	1.5	0	-0.5	0.9
heat exchangers	0	1	0	2 + 1.5 = 3.5	0	4.5	3.25
quench	2	1	0	1.5	0	4.5	3.25
Venturi scrubber	0	1	0	1.5	0	2.5	1.95
scrubbing tower	2	1	0	1.5	0	4.5	3.25
Total score		(Note: all stainless steel recommended by Lurgi due to corrosion problems)	(Note: residence time ~ seconds or minutes (not hours or weeks))	(Note: 2 + 1.5 = temperature + pressure extremes)			31.15

### V.1.2 Wilson's method

Wilson's method [15] is also a "step" method, which is based on the following formula:

$$I_B = f \cdot N \cdot (AUC) \cdot F_p \cdot F_t \cdot F_m$$

where	AUC:	Average unit cost (1971 base year) = £ 352693 / unit
	$F_p$ :	Pressure factor = 1.06
	$F_t$ :	Temperature factor = 1.15
	$F_m$ :	Material factor = 1.8
	f:	Investment factor = 1.5
	N:	Number of items of process equipment of roughly equal size = 11

It was assumed that each lock hopper train with conveyor belt constitutes one item of equipment. Furthermore the six cyclones (1 large and 5 small) were approximated as being two items of equipment.

The above resulted in an estimated investment of £ 12.8 million in base year 1971. This was corrected using the CPE price index [9] to being approximately £ 79 million in 1993. At the current exchange rate (November 1993) of 1 DM equals £ 0.41, this resulted in fixed capital costs of 192 million DM.

### V.1.3 Lang's method

The method proposed by Lang is a factorial method which takes the specific details of the equipment which has been designed into account. According to Coulson & Richardson [9] it can be used to make a quick estimate of capital cost in the early stages of project design, when the preliminary flow sheets have been drawn up and the main items of equipment are roughly sized. This made Lang's method ideal for our purposes.

This method is based on the following formulae:

$$Cf = f_L \cdot C_e$$

where	$C_f$ :	Fixed capital costs
	$f_L$ :	Lang factor = 4.7 for predominantly fluids processing plant
	$C_e$ :	Total delivered cost of all the major equipment items

$$C_e = C \cdot S^n$$

where	C:	Cost constant
	S:	Characteristic size parameter for an item of equipment
	n:	Index for specific type of equipment

The calculations (Table 13) resulted in an estimated fixed capital cost for the HTW installation of DM 121 million.

Table 13: Calculation of the fixed capital cost using Lang's factorial method.

Apparatus (model)	Number N	Cost index C	Characteristic size parameter S	Equipment type index n	Purchased equipment cost $C_e = NCS^n$	
Gasifier (reactor)	1	£6,000	71.3 m <sup>3</sup>	0.40	£33,061	
Bunker (storage tank)	1	£600	294.5 m <sup>3</sup>	0.65	£24,159	
Lock hoppers in (vertical process tanks)	3	£500	42.4 m <sup>3</sup>	0.59	£13,687	
Belt + 5 screws (conveyors; 1 m wide)	6	£1,500	6 m	0.65	£28,843	
Cyclones (vertical process tanks)	1	£500	26.6 m <sup>3</sup>	0.59	£3,465	
	1 (5)	£500	1.32 m <sup>3</sup>	0.59	£589	
Lock hoppers out (vertical process tanks)	6	£500	6.28 m <sup>3</sup>	0.59	£8,873	
Quench (evaporator)	1	£4,000	5.52 m <sup>2</sup>	0.53	£9,888	
Venturi scrubber (evaporator)	1	£4,000	1.23 m <sup>2</sup>	0.53	£4,458	
Scrubbing tower (vertical process tank)	1	£500	52.95 m <sup>3</sup>	0.59	£5,201	
				TOTAL EQUIPMENT COST $C_e$ :	£132,224	
Heat exchanger 0 (gasifier)	104 m <sup>2</sup> £34,000			MATERIAL CORRECTION	STAINLESS STEEL: (x 3.7)	£489,229
Heat exchanger 1 (high press. steam)	541 m <sup>2</sup> £94,000			PRESSURE CORRECTION	PRESSURE < 30 BAR: (x 1.4)	£684,920
Heat exchanger 2 (low press. steam)	375 m <sup>2</sup> £74,000			EXTRA COST	HEAT EXCHANGERS:	£252,500
Total cost (£)	202,000			TOTAL EQUIPMENT REAL COST $C_e$ :		£937,420
Pressure corr. (£) (x 1.25)	252,500			LANG FACTOR CORRECTION	FIXED CAPITAL COST $C_f$ : ( $C_e \times f_L = 4.7$ )	£4,405,876
				INFLATION CORRECTION	INFLATION 8%/YR MID-'79 -> END-'93: (x 3.0524)	£13,448,588
				EXCHANGE RATE CONVERSION	EXCHANGE RATE £/DM MID-'79: (x 9)	DM 121,037,288

## V.2 Costs dependent on production volume

These costs were calculated based on the price data supplied by Brennstoff Kombinat Vresova A.G. The estimate can be made with:

$$K_p = k_p \cdot P$$

$$k_p = \sum_{i=1}^N v_i \cdot q_i$$

where

P:	plant capacity = 1.82 million tons/year
$v_i$ :	costs per ton of fuel or required flow
$q_i$ :	quantity of $i$ needed per ton of product

The net production of steam by the gasifier plant was not taken into account as a financial benefit.

The costs for lignite, oxygen (95%, 24 bar) and cooling water (20 °C) are 1038.6, 460.0 and 0.35 Czech Crowns per ton respectively.

Using the data from the mass balance sheets one finds that the requirements for lignite, oxygen and cooling water are 0.61, 0.33 and 2.7 tons per ton of product respectively.

This would result in a cost of 624 Czech Crowns per ton of product gas, which equals DM 37.10 per ton of product gas.

If the plant were to run continuously (365 days per year, 24 hours per day) it would produce a maximum of 1.8 million tons of synthesis gas. However, it is envisaged to operate the HTW gasifier only during peak hours (approximately 8 hours per day).

Then, on an annual basis, the costs dependent on the production volume will be a maximum of 378 million Czech Crowns, which equals DM 22.5 million.

During operation of the plant, a maximum of 354 MW of electricity can be produced when the net produced steam from the gasifier section is also used to drive the steam turbine, and the coal dust is burnt in the existing coal combustion plant.

This electricity has a market value of 3000 Czech crowns per MWh, resulting in a turnover of 3099 million Crowns per (or DM 184 million) annually.

