Vakgroep Chemische Procestechologie

Verslag behorende bij het fabrieksvoorontwerp van

J. van Hout

P. van Keep

onderwerp:

The Design of an Interconnected Fluidized Bed Combustor

with Regenerative Sulphur Retention

adres: Coendersstraat 37, Delft

devlouw 48, Delft

opdrachtdatum: 10 Februari 1992

verslagdatum: 14 Oktober 1992

TU Delft

Faculteit der Scheikundige Technologie en der Materiaalkunde

Technische Universiteit Delft
THE DESIGN OF A INTERCONNECTED FLUIDIZED BED COMBUSTOR WITH REGENERATIVE SULPHUR RETENTION

J. van Hout & P. van Keep

Delft, October 1992
Discussion issues

a. Design conditions
   (Table III)
   Sensitivity analysis

b. Models
   bed 1: page 14: $t \leftrightarrow T$ residence time
   page 15: simplification w.r.t. shrink core
   remark: degrees of freedom?
   how chosen

   bed 2: page 18: what happens to crushed sorbent particles
   to top or bottom?

   bed 3: -
   bed 4: -

c. Flow sheet:
   particle res. time distr: Gaussian? Why?
   page 24
   distinction between distributed attrition rate
   and residence time

   attr. res. graph

   d. Particle size dependencies
      - Table 7 and 8. page 25-26

   e. What are the innovations/improvements
      in the design?

   f. Steam cycle page 29, 30, 31
      - pinch technology applied to select exch. patterns?
      - heat losses to environment
      - exergy analysis?

   g. Filters
      page 35: bag filters: how much dust do they have
      to handle?

   h. Gun distributor: -

   i. Safety page 40: expl. of Fig.5.

   j. Cost analysis & Sensitivity analysis
      page 42: 336 employees? Interests on loan?
      page 43: profit after taxes

   k. page 44: Conclusions: is the process competitive?
Summary

The usage of coal for power generation will augment. The reserves of coal are far greater than the reserves of other fossil energy sources. Furthermore, coal is cheaper. To convert coal into usable energy, a fluidized bed combustor is an option.

The advantages of a fluidized bed combustor over conventional powder combustor lies in the following facts:

- It has a lower NOx-emission, due to the lower combustion temperature of 850 °C.
- In a fluidized bed, it is possible to adsorb the sulphur with a sorbent.

In this design sulphur caption takes place, using a regenerative sorbent, CaO on a carrier. The combustor is a four-cell interconnected fluidized bed. In the first bed, the largest, the combustion takes place as well as the desulphurization. In the second bed, the segregation of ashes and sorbent take place. Bed three is used for the regeneration of the sorbent. All these three beds are fluidized beds. The fourth bed is a hopper bed and is used to regulate the sorbent flow through the system.

An overview of the characteristics of the design of the interconnected fluidized bed combustor are given in the following table.

<table>
<thead>
<tr>
<th>Power generated</th>
<th>118 MW&lt;sub&gt;e&lt;/sub&gt;</th>
</tr>
</thead>
<tbody>
<tr>
<td>Coal flow for combustion</td>
<td>10.4 kg/s</td>
</tr>
<tr>
<td>Bed heights</td>
<td></td>
</tr>
<tr>
<td>Surface bed 1</td>
<td>4 m</td>
</tr>
<tr>
<td>Surface bed 2</td>
<td>90 m&lt;sup&gt;2&lt;/sup&gt;</td>
</tr>
<tr>
<td>Surface bed 3</td>
<td>5 m&lt;sup&gt;2&lt;/sup&gt;</td>
</tr>
<tr>
<td>Surface bed 4</td>
<td>2 m&lt;sup&gt;2&lt;/sup&gt;</td>
</tr>
<tr>
<td>Surface bed 5</td>
<td>0.5 m&lt;sup&gt;2&lt;/sup&gt;</td>
</tr>
<tr>
<td>Sulphur retention</td>
<td>99± %</td>
</tr>
<tr>
<td>SO&lt;sub&gt;x&lt;/sub&gt;-concentration in stack gas</td>
<td>13.14 mg/m&lt;sup&gt;3&lt;/sup&gt;</td>
</tr>
<tr>
<td>Air flow in combustor</td>
<td>16.32 g/J</td>
</tr>
<tr>
<td>Make up sorbent flow</td>
<td>364 m&lt;sup&gt;3&lt;/sup&gt;/s</td>
</tr>
<tr>
<td>Coal combustion in bed 1</td>
<td>27 g/s</td>
</tr>
<tr>
<td>Coal combustion in bed 1</td>
<td>100 %</td>
</tr>
<tr>
<td>Gas cleaning</td>
<td></td>
</tr>
<tr>
<td>28 cyclones</td>
<td></td>
</tr>
<tr>
<td>165 m&lt;sup&gt;2&lt;/sup&gt; filter cloth</td>
<td></td>
</tr>
<tr>
<td>Total investment (in Dutch Guilders)</td>
<td>196 million</td>
</tr>
<tr>
<td>Profit (in Dutch Guilders)</td>
<td>25.3 million/annum</td>
</tr>
<tr>
<td>Pay out time</td>
<td>7.35 years</td>
</tr>
<tr>
<td>Return on investment</td>
<td>12.9 %</td>
</tr>
</tbody>
</table>

Is this process competitive?
Introduction

The world nowadays strongly depends on the usage of fossil energy sources, mainly oil. Considering the world's oil consumption rate, the oil reserves will run dry within fifty years. A very good alternative fossil energy source is coal, of which the reserves at the present consumption rates will last us more than three hundred years (figure 1). Furthermore, coal is relatively cheap, which makes its usage even more attractive.

![Energy reserves, based on the present-day consumption (Moulijn & Makkee, 1990).](image)

Using coal as an energy source is done mainly by combustion. This has some disadvantages concerning the environment. Let alone the emission of carbon dioxide, the main problems are the relatively high emissions of sulphur oxides, of nitrogen oxides and of dust. Because the advantages of coal combustion are of great weight, much research has been done to solve the environmental problems. This has led to the development of fluidized bed coal combustion. This kind of combustion has many advantages:
- In the fluidized bed coal combustor, it's possible to adsorb the sulphur on a sorbent, which makes a separate desulphurization installation unnecessary.

- Because of the low combustion temperature of 850°C, the production of the nitrogen oxides is low and the ash, which has a melting temperature of 950°C, stays solid and can be easily removed.

- Due to the high flame stability in a fluidized bed coal combustor, it is possible to use coal with a high ashes or water content as well as garbage with a low energy content.

- The used coal may be coarser, which makes the grinding process cheaper.

- The heat transfer coefficient of the tube bundles in the bed is very high, which makes a compact reactor possible.

Presently, limestone is used as sorbent. It reacts with sulphur dioxide in the following way:

\[
\text{CaCO}_3 \rightarrow \text{CaO} + \text{CO}_2 \\
\text{CaO} + \text{SO}_2 + \frac{1}{2}\text{O}_2 \rightarrow \text{CaSO}_4
\]

Limestone is very cheap, but regenerating it is hard. This process produces lots of gypsum, more than is used. So this only shifts the problem from air pollution to an excess solid problem.

To solve this problem, research has been done to find a regenerative sorbent. The outcome is the usage of CaO on a carrier (Van Den Bleek et al., 1989-1990).

The first design of a coal combustion power station, using the regenerative sorbent, was done by Berends and Rikken (1989). This design consists of two atmospheric fluidized beds connected by pipes. In the first bed coal is combusted and the sulphur dioxide adsorbed. In the other bed, the sorbent is regenerated. Sorbent transport is done pneumatically. 85% of the sulphur dioxide is withdrawn from the outlet gas, which is very promising.

The forces acting on the sorbent particles in the pipes in this design are of such a magnitude that there is a considerable attrition. Therefore, a system had to be designed in which the sorbent is transported in another way. The next step was the use of a four-cell interconnected fluid bed. The design of this system has been done by Rozemeijer and Van Der Sluis (1990).
Their system consists of two combustor beds, a regeneration bed and a segregation bed, all fluidized. The first combustor bed is highly fluidized. A good mixing of coal and sorbent takes place which promotes the retention of sulphur. In the second combustor bed, the remaining coal is burned. The fluidization gas velocity is low.

In the third bed the regeneration of the sorbent takes place. This bed is highly fluidized. The fluidization gas used for the regeneration, comes from partial coal gasification. In the fourth bed, the segregation of ashes and sorbent takes place. The fluidization gas velocity is low.

Due to changing insight in hydrodynamical and kinetical behaviour of the system, combined with stricter environmental regulations, a new design is necessary.

a new or a redesign
Interconnected Fluid Bed System

Configuration

The interconnected fluid bed system is a system in which the combustion of coal, the acceptation of sulphur dioxide, the regeneration of sorbent and the separation of ashes and sorbent, are integrated into one system. It consists of four beds, three of them are fluidized and one is a hopper bed (figure 2).

The four beds are connected in such a way that the fluidized particles from bed 1 flow over the separation wall between beds 1 and 2 into bed 2. Bed 2 is connected to bed 3 with an orifice. The flow between bed 3 and 4 proceeds in the same manner as between bed 1 and 2. Between bed 4 and 1 there is again an orifice. This completes the cycle in which the sorbent particles circulate. The moving force for the sorbent - flow is the difference in bulk densities between the beds connected with an orifice. These differences are caused by differing fluidization velocities.

The first bed, bed 1 is a highly fluidized bed in which the constantly-inflowing coal is burnt and the liberated sulphur is adsorbed by sorbent. The bed is fluidized by air. Bed 2 is a lowly fluidized bed in which the ashes and the sorbent are segregated. This bed is also fluidized by air. The off-gas of beds 1 and 2 are led through cyclones and bag filters to the stack.

Bed 3 is again a highly fluidized bed. The regeneration of the sorbent takes place in this bed which is fluidized by a mixture of 90% nitrogen and 10% hydrogen. Bed 4 is a moving or hopper bed. Pure nitrogen is used as fluidization gas. This bed is used to regulate the sorbent flow through the whole system. This flow is regulated by the gas velocity in this bed. The off-gas of beds 3 and 4 are led to a Claus-plant, so as to recover the sulphur.
Figure 2. Interconnected fluid bed combustor with four beds.
Design parameters

Conditions for the plant
The most important design condition is to meet the requirements concerning the environment, set by the Dutch Government. These affect the emissions of carbon monoxide, sulphur dioxide, nitrogen oxides and dust.
In 1982, two sets of requirements for coal combustion existed, one dating from September 1981, the other of September 1982. In 1987, on the tenth of April, new regulations for large coal fired power stations were introduced (Heek, 1988). These will even become more strict as can be seen in Table 1.

Table 1. Regulations for emissions from coal fired power stations.

<table>
<thead>
<tr>
<th></th>
<th>sept '81</th>
<th>sept '82</th>
<th>1987</th>
<th>newest</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO</td>
<td>86 g/GJ</td>
<td></td>
<td>700 mg/m³</td>
<td>700 mg/m³</td>
</tr>
<tr>
<td>SO₂</td>
<td>230 g/GJ</td>
<td>600 g/GJ</td>
<td>700 mg/m³</td>
<td>700 mg/m³</td>
</tr>
<tr>
<td>NOₓ</td>
<td>150 g/GJ</td>
<td>190 g/GJ</td>
<td>500 mg/m³</td>
<td>100 mg/m³</td>
</tr>
<tr>
<td>dust</td>
<td>35 g/GJ</td>
<td>20 g/GJ</td>
<td>50 mg/m³</td>
<td>20 mg/m³</td>
</tr>
</tbody>
</table>

Internal publication of the ministry of VROM, 1992
The strictest regulations have been used as the starting-point of the design. These and other conditions for the design can be seen in Table 2.

Table 2. Conditions for the design.

<p>| | |</p>
<table>
<thead>
<tr>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Electrical power</td>
<td>100 MW</td>
</tr>
<tr>
<td>Steam pressure</td>
<td>100 bar</td>
</tr>
<tr>
<td>Steam temperature</td>
<td>500 °C</td>
</tr>
<tr>
<td>Feed water temperature</td>
<td>50 °C</td>
</tr>
<tr>
<td>Fuel</td>
<td>Coal (Polish-5), gas</td>
</tr>
<tr>
<td>Sorbent</td>
<td>SGC-500</td>
</tr>
<tr>
<td>SO₂-emission</td>
<td>&lt; 700 mg/m³</td>
</tr>
<tr>
<td>NOₓ-emission</td>
<td>&lt; 100 mg/m³</td>
</tr>
<tr>
<td>Dust</td>
<td>&lt; 20 mg/m³</td>
</tr>
<tr>
<td>Bed temperature</td>
<td>850 °C</td>
</tr>
<tr>
<td>Temperature stack-gas</td>
<td>&lt; 120 °C</td>
</tr>
<tr>
<td>Number of beds</td>
<td>4</td>
</tr>
<tr>
<td>Attrition</td>
<td>60% in 1500 hours</td>
</tr>
<tr>
<td>Sorbent deactivation</td>
<td>0.3% per cycle</td>
</tr>
<tr>
<td>Bed heights</td>
<td>≤ 4 m</td>
</tr>
</tbody>
</table>
Raw Materials

Coal:

The coal used in this design is Polish-5. It has a relatively low sulphur content. The characteristics and composition are given in table 3. These are based on an assumptive composition.

Table 3. The characteristics and composition of Polish-5.

<table>
<thead>
<tr>
<th></th>
<th>weight-%</th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>C</td>
<td>75.6</td>
<td>S</td>
<td>0.75</td>
<td>H</td>
<td>4.7</td>
<td>N</td>
<td>1.3</td>
</tr>
<tr>
<td>Fly ash</td>
<td></td>
<td>Bottom ash</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Water</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Energy release</td>
<td>29370</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Volatile part</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Density bottom ash</td>
<td>2448</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Particle size b. ash</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

: based on experiments

Sorbent:

The synthetic regenerative sorbent consists of CaO on a porous gamma-alumina carrier. This combination is made by using the sol-gel method, existing of the following steps. A A100H-sol is trickled into a two-phase system. The floating phase is kerosine in which the A100H forms little spherical particles. The bottom phase is a watery solution of NH₄⁺. This phase turns the particles into a alumina-gel. The particles are subsequently dried and calcinated at a temperature of 850°C.

This process leads to the SGC-500 (Sol Gel Condea) sorbent. It has the following characteristics.

Table 4. Characteristics of SGC-500.

<table>
<thead>
<tr>
<th></th>
<th>weight-%</th>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Ca-content</td>
<td>8.91</td>
<td>Pore-volume</td>
<td>0.40</td>
<td>Porosity</td>
<td>0.56</td>
<td>Density</td>
</tr>
</tbody>
</table>
Bed configuration parameters

Bed 1
Bed 1 is the fluidized bed in which the actual coal combustion takes place. Furthermore, here the sulphur oxides are captured, using the sorbent. The residence time for the bulk gases has to be large enough to ensure a good adsorption.

In this bed the main part of the heat is absorbed by the heat exchangers, so the bed has to be large enough to accommodate a heat exchanging system capable of draining the excess heat in bed 1.

The particles in this bed are assumed to behave like in a Continuously Stirred Tank Reactor (CSTR). The fluidization regime is slugging or bubbling.

The combustion of coal is complete. The residence time in bed 1 is fixed by the condition that 95% of the carbon reacts in bed 1. The remaining five percent is blown out into the freeboard as fly ash. Some of it will combust here. By recycling the fly ash, a complete combustion of the coal is realised.

During the combustion, nitrogen oxides will be formed. Due to the low operating temperature, the concentrations of these oxides will remain low. If they nevertheless exceed the environmental standards, the oxides can be reduced to ammonia using ureum. This requires equipment for treating the stack gas in this way.

Bed 2
In bed 2 segregation between ashes and sorbent takes place by means of the application of a low fluidization velocity. Then, the ash, which will remain at the bottom of the bed, can be withdrawn, using a special valve. By configuring the segregation in this way, the loss of sorbent can be held as minimal as possible.

Bed 3
Bed 3 is the regeneration bed. The sorbent is regenerated by using a mixture of nitrogen and hydrogen as fluidization gas. Because the regeneration is weakly endothermic, a way to get heat into bed 3 has to be found. The off-gas from this bed can be led to a Claus-plant, where the sulphur will be recovered or to a sulphurous acid plant, where the sulphur dioxide is transferred into sulphurous acid. The choice between these two options, depends on the situation concerning the use of sulphur in the world.
Bed 4
Bed 4 is a moving- or hopper bed. It is used to regulate the sorbent flow through the whole system. The flow between two beds depends on the difference in bed densities. The sorbent will flow from a bed with a higher bed density to one with the lower bed density. Because bed 4 is used to regulate the flow of sorbent through the system, the difference in densities between beds 4 and 1 has to be lower than the difference between beds 2 and 3. This restricts the choice of the fluidization velocities in beds 2 and 3.
Calculation Models

Model for the bed voidage

The bed voidage at the minimal fluidization can be calculated using a correlation based on the number of Lyaschenko:
(equation 1)

\[ \epsilon_{mf} = 0.513 \times Ly^{-0.017} \]

where:
\[ \epsilon_{mf} = \text{bed voidage at minimal fluidization conditions} \]

The Lyaschenko number is calculated in the following way:
(equation 2)

\[ Ly = \frac{U_t^3}{g \mu (\rho_s - \rho_{gas})} \]

where:
\[ U_t = \text{terminal fluidization velocity (m/s)} \]
\[ g = \text{acceleration of gravity (m/s}^2) \]
\[ \mu = \text{viscosity (Ns)} \]
\[ \rho_s = \text{density solid particles (kg/m}^3) \]
\[ \rho_{gas} = \text{density gas (kg/m}^3) \]

In this equation \( U_t \) is the terminal fluidization velocity, the gas velocity at which all the particles are blown out. This velocity is calculated with the following equation:
(equation 3)

\[ U_t = \frac{4 \times g \times d_p \times (\rho_s - \rho_{gas})}{\sqrt{3 \times \rho_{gas} \times 0.431}} \]

where:
\[ d_p = \text{diameter solid particle (m)} \]

The voidage at velocities higher than the minimal fluidization velocity are calculated with the following model (Van den Bleek, 1989)
(equation 4)

\[ d_{eq} = \frac{(u-U_{mf})^{0.5} \times (0.4 \times H)^{0.75}}{g^{0.25}} \]

where:
\[ d_{eq} = \text{help variable (-)} \]
\[ u = \text{gas velocity (m/s)} \]
\[ u_{mf} = \text{minimal fluidization gas velocity (m/s)} \]
\[ H = \text{bed height (m)} \]
(equation 5)
\[ U_a = U - U_{mf} + 0.71 \sqrt{g \cdot d_{eq}} \]

where:
- \( u_a \) = help variable (m/s)

(equation 6)
\[ \varepsilon = \frac{U - U_{mf}}{U_a} \]

where:
- \( \varepsilon \) = bed voidage (-)

Model for the sulphur dioxide capture

For the sulphur dioxide capture in bed 1, the Sulphur Retention, SURE2-model (Wolff, 1991) is used. This model is based on the shrinking core model. The model consists of two differential equations, one for the changing sulphur dioxide concentration in the bulk gas phase, the other for the changing core radius. Both the equations are valid for a system that behaves as a continuously stirred tank reactor (CSTR). The fluidized bed can be considered as a CSTR, as told before.

(equation 7)
\[ \frac{dC_{SO_2}}{dt} = \frac{\phi_f}{V_g} \left( C_{SO_2, feed} - C_{SO_2} \right) - \ldots \]

\[ \ldots - \left( \frac{1}{V_g \left( 1 + K_{eq} \cdot C_{eq}^{\frac{1}{2}} \right)} \right) \left( \frac{1}{K_a \cdot K_{eq} \cdot C_{eq}^{\frac{1}{2}}} + r_c^2 \cdot \left( a_{SO_3, film} + a_{SO_3, shell} \right) \right) \]

where:
- \( C_{SO_2} \) = concentration sulphur dioxide (mg/m³)
- \( t \) = residence time (s)
- \( \phi_f \) = mass flow fluidum (kg/s)
- \( C_{SO_2, feed} \) = conc. sulphur dioxide in feed stream (mg/m³)
- \( V_g \) = gas volume in tank (m³)
- \( K_{eq} \) = equilibrium constant of \( SO_2/SO_3 \) equilibrium (m³/mole)½
- \( C_{eq} \) = concentration oxygen (mg/m³)
- \( r_c \) = unreacted core radius (m)
- \( N_{part} \) = number of particles (-)
- \( V_r \) = reactor volume (m³)
- \( k_s \) = reaction rate constant of \( SO_3 \) on core (m³ gas/m² core s)
- \( a_{SO_4} \) = help variable (-)
The model considers three reaction steps, which together define the overall rate of the process:

**Step 1.** Diffusion through the film surrounding the particle, to its surface. It's corresponding transport parameters are the SO\textsubscript{x} film mass transport coefficient, \( k_{SOx,film} \) (m/s) and the diffusion coefficient of SO\textsubscript{x} in the film, \( D_{SOx,film} \) (m\(^2\)/s). Experiments have shown that this step is not rate-controlling.

**Step 2.** Diffusion of the reactant through the shell to the surface of the unreacted core. This process is described with the diffusion coefficient of SO\textsubscript{2} in the shell, \( D_{SOx,shell} \) (m\(^2\)/s).

**Step 3.** The adsorption of SO\textsubscript{3} on the external surface area of the core. The adsorption is first order in SO\textsubscript{3} and proportional to the surface area of the core. The parameter describing the adsorption is \( k_s \). The equilibrium constant for the SO\textsubscript{x} equilibrium, \( SO_2 + \frac{1}{2} O_2 \rightarrow SO_3 \), is \( K_{eq} \).

The influence of the diffusion is found in the alpha's. The alpha's are calculated with the following equations:

(equation 9)

\[
\alpha_{SO_x,film} = \frac{1}{D_{SO_2,film} + D_{SO_2, film} * K_{eq} * C_{O_2}^{\frac{1}{2}}} \left( \frac{1}{R} - \frac{1}{R + \delta} \right)
\]

(equation 10)

\[
\alpha_{SO_x,shell} = \frac{1}{D_{SO_2,shell} + D_{SO_2, shell} * K_{eq} * C_{O_2}^{\frac{1}{2}}} \left( \frac{1}{R_c} - \frac{1}{R} \right)
\]

\[
\frac{dr_c}{dt} = \frac{\alpha_v}{a_v + 6\alpha_v r_c^3} \quad \text{where:}
\]

\[
a_1, r_c + \frac{1}{3} r_c^3 = a_0 r_c + C
\]

\[
a_1, (r_c^3 - r_c) + \frac{1}{3} (r_c^3 - r_c^3) = a_0 r_c
\]

\[
r_c(t) \rightarrow \{ t \}
\]
The delta is calculated from:
(equation 11)
\[ \delta = \frac{1}{k_{SO_2, film} \cdot \frac{1}{Ds_{SO_2, film} R}} \]

The mass transport coefficient \( k_{SO_2, film} \) is calculated with the following set of equations:
(equation 12)
\[ St^{\frac{1}{2}} Sc^{\frac{1}{2}} = (0.60 \pm 0.1) \cdot Re_p^{\frac{1}{2}} \]

where:
- \( St \) = Stokes number (-)
- \( Sc \) = Schmidt number (-)
- \( Re_p \) = Reynolds number (-)

(equation 13)
\[ St = \frac{k_{SO_2, film} \cdot \varepsilon_b}{U_f} \]

where:
- \( \varepsilon_b \) = bed voidage (-)
- \( U_f \) = superficial gas velocity (m/s)

Because the combustor is in steady state operation, the model becomes simpler and linear. The time dependence disappears:
(equation 14)
\[ \frac{d(C_{SO_2})}{dt} = 0 \]

Furthermore, the core radius is taken as a constant, which leads to:
(equation 15)
\[ \frac{dr_c}{dt} = 0 \]

The core radius used henceforth is the average core radius.
\[ \langle r_c \rangle = \int_0^{\infty} r_c(t') \cdot \zeta(t') \, dt' \]
\[ \zeta(t') = \frac{1}{\tau^\theta} e^{-\tau'/\theta} \]
\[ r_c(t') = \int_0^\infty \frac{\zeta(t') \tau_c(t')}{\partial \tau} \, dt \]
This makes equation 8 obsolete. Equation 7 combined with equation 14, becomes a linear equation,\
\[ C_{\text{SO}_2} = C_{\text{SO}_2,\text{ing}} - \frac{1}{\phi_r*(1+K_{eq}*C_{\text{C}_2}^\frac{1}{2})} \times \frac{4*\pi*R_c^2*C_{\text{SO}_2}*N_{\text{part}}*V_r}{K_{eq}*K_{c}*C_{\text{C}_2}^\frac{1}{2}} \]

This equation is used in the model to calculate the sulphur retention. The calculated value is \( \alpha \), the surface occupied by \( \text{CaSO}_4 \) divided by the total surface occupied by \( \text{Ca} \). \( \alpha \) is connected with \( r \), through the dimensionless radius:
\[ Y = \frac{r}{R} \]
where:
\[ R = \text{sorbent particle radius (m)} \]
The equation which relates these two is the following:
\[ \alpha = 1 - Y^3 \]
The remaining surface fraction is the fraction \( \text{CaO} \).

The formation of nitrogen oxide
Because of the relatively low combustion temperature of 850 °C, none of the incoming nitrogen molecules will react to form nitrogen oxide. But the nitrogen atoms in the coal may do so. The nitrogen contained in the coal is released in two forms. The volatile part of 31.8% is released as ammonia, the other part is released as nitrogen oxide. The ammonia as well as the nitrogen oxide may undergo further reaction. Three reactions are possible (Lin et al., 1990):

\[ \text{NO} + \text{C} \rightarrow \frac{1}{2} \text{N}_2 + \text{CO} \quad k_a = 1.3 \times 10^5 e^{(34000/RT)} \quad [\text{m/s}] \]
[Horio, 1977]

\[ \text{NH}_3 + \frac{1}{2} \text{O}_2 \rightarrow \text{NO} + \frac{1}{2} \text{H}_2\text{O} \quad k_b = 0.5-2 \quad [\text{m}^3/\text{mol} \cdot \text{s}] \]
catalysed by CaO

\[ 2 \text{NH}_3 + \frac{1}{2} \text{O}_2 \rightarrow \text{N}_2 + 3 \text{H}_2\text{O} \quad k_c = 5.2-8.6 \quad [\text{m}^3/\text{mol} \cdot \text{s}] \]
catalysed by char
With these three reactions and their kinetic constants, a model for the calculation of the nitrogen oxide concentration can be established. This model is based on a steady-state situation:

(equation 22)

\[
C_{NH_3} = \frac{\phi_f \cdot C_{NH_3, feed}}{\phi_f + k_b \cdot S_{CaO} \cdot C_{O_2} + k_c \cdot S_c \cdot C_{O_2}}
\]

where:
- \( C_{NH_3} \) = concentration NH\(_3\) (mg/m\(^3\))
- \( C_{NH_3, feed} \) = concentration NH\(_3\) in feed stream (mg/m\(^3\))
- \( S_{CaO} \) = surface CaO (m\(^2\))
- \( S_c \) = surface char (m\(^2\))

(equation 23)

\[
C_{NO} = \frac{\phi_f \cdot C_{NO, feed}}{\phi_f + k_a \cdot S_c} + \frac{k_b \cdot S_{CaO} \cdot C_{O_2} \cdot \phi_f \cdot C_{NH_3, feed}}{(\phi_f + k_a \cdot S_c) \cdot (\phi_f + k_b \cdot S_{CaO} \cdot C_{O_2} + k_c \cdot S_c \cdot C_{O_2})}
\]

where:
- \( C_{NO} \) = concentration NO (mg/m\(^3\))
- \( C_{NO, feed} \) = concentration NO in feed (mg/m\(^3\))

The surfaces \( S_c \) and \( S_{CaO} \) are calculated in the following way:

(equation 24)

\[
S_c = \frac{6 \cdot \phi_{coal} \cdot \tau}{p_{s} \cdot d_{coa1}}
\]

where:
- \( \phi_{coal} \) = mass flow coal (kg/s)
- \( p_{s} \) = density coal (kg/m\(^3\))
- \( d_{coa1} \) = diameter coal particle (m)

(equation 25)

\[
S_{CaO} = 4 \cdot \pi \cdot N_{part} \cdot r_c^2
\]
Ashes and sorbent segregation

The segregation of the ashes arising from the coal combustion and the sorbent is done in bed 2. This segregation is necessary because 20% of the arising ashes stays in the system and if it is not removed it will accumulate in the system. In bed 2, because of the segregation based on different size, densities and minimal fluidization velocities, the ash particles stay at the bottom of the bed and the sorbent particles "floats" on top of it, as has been shown in experiments. By changing the diameter of the particles used, a totally different situation can occur and the ash might "float" on top of the sorbent. As the design is in such a way that the sorbent has to float on top of the ash, such a situation has to be avoided. The modelling of segregation is difficult, especially determining which particles will float and which will sink. To be sure which is which, an experiment with the particles should be performed. If such an experiment points out that the sorbent will sink, then the ash has to be removed using an overflow.

In this design the ash is assumed to sink, so it will be removed through a valve. A model for calculating the effectiveness of the segregation is given by M. Pell (1990):

Firstly, the fraction in the whole bed is calculated:

\[ x = \frac{\text{fraction bottom ash} \cdot \phi_{\text{coal}}}{\text{fraction bottom ash} \cdot \phi_{\text{coal}} + \phi_s} \]

where:
\[ x = \text{ash fraction (-)} \]
\[ \phi_{\text{coal}} = \text{mass flow coal (kg/s)} \]
\[ \phi_s = \text{mass flow sorbent (kg/s)} \]

When the bed is completely mixed, the ash fraction in the toplayer, the top ten percent of the bed, is the same as in the overall bed. The mixing index \( M \), which is defined in equation 27, has in this case the value 1. If there is no ash in the toplayer at all and the segregation is complete, then the mixing index is 0. The mixing index is a function of the velocity and the fluidization characteristics of the dissimilar solids. It is correlated using the following equations:

\[ M = \frac{1}{1 + e^{-z}} \]
Where $z$ is the velocity ratio defined by,
(equation 28)

$$z = \frac{U_F*UTO*e^{-\frac{U_F}{UTO}}}{U_F*U_{mf, sorbent}}$$

where

$U_{mf, sorbent} =$ minimal fluidization velocity for the sorbent particles (m/s)

UTO is defined as the takeover velocity, the velocity at which the mixing index is 0.5. It is given by:

(equation 29)

$$\frac{UTO}{U_{mf, sorbent}} = (\frac{U_{mf, ashes}}{U_{mf, sorbent}})^{1.2} + 0.9*(\frac{\rho_{ashes}}{\rho_{sorbent}})^{1.1} + 2.2*\sqrt{\bar{x}}*(1-e^{-\frac{H}{dt}})^{1.4}$$

where

$U_{mf, ashes} =$ minimal fluidization velocity for the ash particles (m/s)

$\rho_{ashes} =$ density ash particles (kg/m$^3$)

$\rho_{sorbent} =$ density sorbent particles (kg/m$^3$)

$H =$ bed height (m)

dt =$ tank diameter (m)

Using this mixing index we can calculate the fraction ashes in the top layer, which is a measure for the degree of segregation:
(equation 30)

$$x = M\bar{x}$$

**Regeneration**

The sorbent which has been sulfated in bed 1 has to be regenerated. This regeneration takes place in bed 3. This bed is fluidized by a mixture of 90% nitrogen and 10% hydrogen. 10% hydrogen is shown in experiments to be the optimal value between cost and effectiveness (Van Hout & Kok, 1992).

The hydrogen takes care of the regeneration. The following reactions occur(Korbee, to be published):
(equation 31) 
\[ \text{CaSO}_4 + \text{H}_2 \rightarrow \text{CaO} + \text{SO}_2 + \text{H}_2\text{O} \quad k_1 = 0.95 \text{ m/s} \]

(equation 32) 
\[ \text{SO}_2 + 2 \text{H}_2 \rightarrow \frac{1}{2} \text{S}_2 + 2 \text{H}_2\text{O} \quad -\Delta G_0 = 93.98 \text{ kJ/mol} \]

(equation 33) 
\[ \frac{1}{2} \text{S}_2 + \text{H}_2 \rightarrow \text{H}_2\text{S} \quad -\Delta G_0 = 53.32 \text{ kJ/mol} \]

(equation 34) 
\[ \text{CaO} + \text{H}_2\text{S} \rightarrow \text{CaS} + \text{H}_2\text{O} \quad k_4 = 0.005 \text{ m/s} \]

The regeneration model which is based on the SURE-model (Scouten and Van den Bleek), assumes that:
- The fluid bed can be considered a CSTR.
- The rates of the gas-solid reactions are first order with respect to the gas concentration and the outer-surface of the particles.

Looking at the reactions, it is clear that three solid phases are to be found on a particle: CaSO₄, CaO and CaS. These three parts of the sorbent have all been expressed as a fractional surface:

(equation 35) 
\[ \alpha_1 = \frac{\text{surface CaSO}_4}{\text{total surface Ca}} \]

(equation 36) 
\[ \alpha_2 = \frac{\text{surface CaO}}{\text{total surface Ca}} \]

(equation 37) 
\[ \alpha_3 = \frac{\text{surface CaS}}{\text{total surface Ca}} \]

Using these variables and reactions, a model has been made for calculating the average alpha's, the average taken over the whole bed (Van Hout & Kok, 1992):

The average alpha's are given by,

(equation 38) 
\[ \bar{\alpha}_1 = \frac{\alpha \cdot q}{q + k_1 \cdot C_{H_2} \cdot \tau} \]

where
- \( q \) = number of moles Ca per square meter particle surface (mole/m²)
- \( C_{H_2} \) = concentration \( \text{H}_2 \) (mg/m³)
- \( \tau \) = residence time in bed (s)

(equation 39) 
\[ \bar{\alpha}_2 = \frac{k_1 \cdot C_{H_2} \cdot \bar{\alpha}_1}{k_4 \cdot C_{H_2} \cdot \bar{\alpha}_1 + \frac{1 - \alpha}{k_4 \cdot C_{H_2} + k_1 \cdot C_{H_2}}} + \frac{\alpha \cdot k_1 \cdot C_{H_2}}{1 + k_4 \cdot C_{H_2} \cdot \frac{\tau}{q}} \]
where
\[ C_{\text{H}_2\text{S}} = \text{concentration \ H}_2\text{S (mg/m}^3\text{)} \]
(equation 40)
\[ \bar{\alpha} = 1 - \bar{\alpha}_1 - \bar{\alpha}_2 \]

The two equilibria that influence the gas concentrations are introduced in the model in the following way:
(equation 41)
\[ 5.51874 \times 10^8 \times C_{\text{SO}_2}^2 \times C_{\text{H}_2\text{O}}^4 = 10.7105 \times C_{\text{S}_2} \times C_{\text{H}_2\text{O}}^4 \]

where
\[ C_{\text{SO}_2} = \text{concentration \ SO}_2 (\text{mg/m}^3) \]
\[ C_{\text{S}_2} = \text{concentration \ S}_2 (\text{mg/m}^3) \]
\[ C_{\text{H}_2\text{O}} = \text{concentration \ H}_2\text{O (mg/m}^3\text{)} \]

(equation 42)
\[ 1929.2299 \times C_{\text{H}_2\text{S}}^2 \times C_{\text{S}_2} = 10.7105 \times C_{\text{H}_2\text{S}}^2 \]

where
\[ C_{\text{H}_2\text{S}} = \text{concentration \ H}_2\text{S (mg/m}^3\text{)} \]

**Sorbent flow and gas leak**

Bed 4 is used to control the flow of the sorbent through the system. The retention in bed 1 requires a certain flow of sorbent, which will be set by regulating the fluidizing velocity in bed 4 at a value below the minimum fluidization velocity.

Because the sorbent flows from bed 4 to bed 1 through an orifice, a gas leak may be expected. This gas leak is supposed to have the same direction as the sorbent flow, because it occurs through entrainment by the particle flow.

The sorbent flow as a function of the fluidizing velocity and the gas leak can be calculated with the following model.
The required sorbent flow is set by the operation of bed 1. Using this, the absolute particle velocity through the orifice can be calculated with:

\[ U_{\text{abs}} = \frac{\phi_s}{\rho_s(1-\epsilon_{mf})A_{\text{orifice}}} \]

where 
\[ A_{\text{orifice}} = \text{surface area orifice (m}^2\text{)} \]

The overall gas-relation for bed 4 is:

\[ \frac{U_{mf}}{\epsilon_{mf}} = U_s + U_{\text{gas}} \]

The solids flow in bed 4, which is positive when going down, equals:

\[ U_s = \frac{\phi_s}{\rho_s(1-\epsilon_{mf})A_b} \]

where 
\[ A_b = \text{surface area bed (m}^2\text{)} \]

The gas velocity in the bed, which is positive when going up, equals:

\[ U_{\text{gas}} = \frac{U_f - \phi_{gl}}{\epsilon_{mf}} \]

where 
\[ \phi_{gl} = \text{mass flow gas leak (kg/s)} \]

Next, the pressure difference at orifice level between, in this case, beds four and one, is calculated:

\[ \Delta P_0 = (H-h_{\text{orifice}}) \left( \frac{dP_4}{dh} - \frac{dP_1}{dh} \right) \]

where 
\[ h_{\text{orifice}} = \text{height of orifice in bed (m)} \]
Using this pressure difference, the horizontal slip velocity of the gas in the orifice can be calculated with:
(equation 48)

\[
\frac{\Delta P_0}{L_{bed}} = 150 \frac{(\frac{1}{\varepsilon_{mf}} - 1)^2 \mu_{gas}}{(\phi_{sphere} \cdot c_p)^2} \Delta U_{hor} + 1.75 \frac{(\frac{1}{\varepsilon_{mf}} - 1) \rho_{gas}}{\phi_{sphere} \cdot c_p} \Delta U_{hor}^2
\]

The horizontal slip velocity consists of two parts:
(equation 49)

\[
\Delta U_{hor} = -U_{abs} + U_{g, hor}
\]

From this, the gas leak can be calculated:
(equation 50)

\[
\phi_{gl} = U_{g, hor} \cdot A_{orifice} \cdot \varepsilon_{mf}
\]
Bed Configurations Calculated

The design of the interconnected fluid bed combustor is shown in appendices III.1 and III.2. Considering all conditions, environmental as well as for the design, leads to the dimensions of the interconnected fluid beds as shown in table 5.

Table 5. General dimensions of the interconnected fluid bed combustor.

<table>
<thead>
<tr>
<th>Feed Flow Coal</th>
<th>10.4 kg/s</th>
</tr>
</thead>
<tbody>
<tr>
<td>Sorbent Flow Through System</td>
<td>7 kg/s</td>
</tr>
<tr>
<td>Make Up Sorbent Flow</td>
<td>27 g/s</td>
</tr>
<tr>
<td>Height of Beds</td>
<td>4 m</td>
</tr>
<tr>
<td>Average Number of Cycles of Sorbent</td>
<td>136</td>
</tr>
<tr>
<td>Average Activity of Sorbent$^*$</td>
<td>0.67</td>
</tr>
<tr>
<td>Orifice Height Bed 2 --&gt; 3</td>
<td>0.3 m</td>
</tr>
<tr>
<td>Orifice Height Bed 4 --&gt; 1</td>
<td>0.1 m</td>
</tr>
</tbody>
</table>

$^*$ as a fraction of the activity of fresh sorbent

The make up sorbent flow is calculated from the fact that the attrition is 60% in 1500 hours. This leads to the following equation:

(equation 51)

$$\phi_{make\ up} = \frac{0.6 \times \phi_{sorbent} \times (\tau_1 + \tau_2 + \tau_3 + \tau_4)}{1500 \times 3600}$$

An important factor in the retention of sulphur is the activity of the sorbent used. Due to chemical deactivation, the sorbent loses each cycle 0.3% of its activity. The mean number of cycles the sorbent particles remain in the system is fixed by the attrition. Supposing the residence time of a sorbent particle in the system is a Gauss-function, the mean residence time in the system can by calculated by choosing a standard deviation. A standard deviation of 700 seconds is chosen. This leads to a mean residence time of 1325 hours. So the mean number of cycles can be calculated by:

(equation 52)

$$n = \frac{1325 \times 3600}{(\tau_1 + \tau_2 + \tau_3 + \tau_4)}$$

So, the mean activity as a percentage of the activity of fresh sorbent is 0.997% times 100%.
The minimum height of the orifice between beds 2 and 3, is calculated with the fact that 20% of the ash is bottom ash and assuming that the segregation is complete. 10.3% of the coal is ash. This leads to the following equation:

\[
h_{\text{orifice}} = \frac{0.2 \times 0.103 \times \phi_{\text{coal}}}{0.2 \times 0.103 \times \phi_{\text{coal}} + \phi_{\text{sorbent}}} \times H_{\text{bed}}
\]

The residence times of ash as well as sorbent are supposed to be equal in this equation.

Table 6 shows the parameters that are specific for each of the beds.

The minimal product of area and gas velocity in bed 1 are fixed by the amount of oxygen needed for the combustion and sulphur retention. Decreasing the area of the bed makes an increase in gas velocity necessary and the other way around. A balance between the two has to be searched. Increasing the gas velocity increases the compressor costs. Increasing the area of the bed makes the investment higher. A optimisation concerning the costs should be made.

Table 6. The dimensions of the interconnected fluid beds.

<table>
<thead>
<tr>
<th>Bed number</th>
<th>Area (m²)</th>
<th>Gas velocity (m/s)</th>
<th>Voidage (−)</th>
<th>Residence Time (h)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>90</td>
<td>4</td>
<td>0.55</td>
<td>8.79</td>
</tr>
<tr>
<td>2</td>
<td>5</td>
<td>0.7</td>
<td>0.40</td>
<td>0.66</td>
</tr>
<tr>
<td>3</td>
<td>2</td>
<td>1.5</td>
<td>0.42</td>
<td>0.26</td>
</tr>
<tr>
<td>4</td>
<td>0.5</td>
<td>−</td>
<td>−</td>
<td>−</td>
</tr>
</tbody>
</table>

*: These values depend on the particle size.

To keep the cost of the design low, small beds as small as possible are used. The area of bed 2 is set by the assumption that particles need time to settle, so the residence time in bed 2 should not be too short. The surface area in bed 3 is determined by the residence time needed in bed 3 to get a good regeneration of the sorbent. The surface area of bed 4 is not set by a particular demand, only making it too big might cause the bed not to operate as a hopper bed, but tunneling might occur. By keeping the surface area small, this problem will not occur. The gas velocity in bed two is chosen so close to the minimal fluidization velocity of bottom ash, to make the segregation optimal.
The gas velocity in bed three influences the regeneration of the sorbent. The gas velocities in bed two and three determine the solids flow through the orifice between the two beds. The gas velocities have to be chosen in such a way that a solids flow of 7 kilogram per second is no problem.

The gas velocity in bed four determines the solids flow through the system. This velocity depends on the particle size of the sorbent used.

All the dimensions aforementioned, are independent of the particle size. The particle size does have an influence on the retention of the sulphur and the regeneration of the sorbent. This influence is shown in the concentrations of the gases in beds one and three. The particle size also determines the gas velocity in bed 4. The most important gas concentrations of bed 1, as well as the retention are shown in table 7 at a stack gas temperature of 120 °C. These concentrations are far below the regulations. (See table 1)

Table 7. The particle dependence of the retention in bed 1.

<table>
<thead>
<tr>
<th>Particle size (mm)</th>
<th>2.0</th>
<th>1.5</th>
<th>1.0</th>
</tr>
</thead>
<tbody>
<tr>
<td>Concentration Sulphur dioxide (mg/m³)</td>
<td>13.14</td>
<td>3.09</td>
<td>3.30</td>
</tr>
<tr>
<td>Concentration Sulphur dioxide (g/GJ)</td>
<td>16.32</td>
<td>3.88</td>
<td>4.15</td>
</tr>
<tr>
<td>Sulphur Retention (%)</td>
<td>99.22</td>
<td>99.80</td>
<td>99.79</td>
</tr>
<tr>
<td>Concentration Nitrogen oxide (mg/m³)</td>
<td>27.49</td>
<td>22.78</td>
<td>22.30</td>
</tr>
<tr>
<td>Concentration Nitrogen oxide (g/GJ)</td>
<td>34.13</td>
<td>28.65</td>
<td>28.04</td>
</tr>
</tbody>
</table>

The regeneration of the sorbent is as mentioned influenced by the particle size. The off gases of this regeneration are led to a Claus plant. Again the temperature of this gas is 120 °C. The concentrations of sulphur-containing components in this gas are given in table 8.
Table 8. The gas concentrations in the stack gas of bed 3.

<table>
<thead>
<tr>
<th>Particle size (mm)</th>
<th>2.0</th>
<th>1.5</th>
<th>1.0</th>
</tr>
</thead>
<tbody>
<tr>
<td>Concentration Sulphurdiox. (mg/m³)</td>
<td>0.15</td>
<td>0.15</td>
<td>0.11</td>
</tr>
<tr>
<td>Concentration Hydrogen sul. (g/m³)</td>
<td>0.26</td>
<td>0.21</td>
<td>0.33</td>
</tr>
<tr>
<td>Concentration S₂ (g/m³)</td>
<td>0.65</td>
<td>0.46</td>
<td>0.07</td>
</tr>
</tbody>
</table>

Comparing the stack gas concentrations from bed 1, suggests the use of a sorbent particle that is as small as possible. Apart from the fact that a smaller particle may cost more, the size reduction has a negative effect on the segregation. Particles bigger than 2 mm, will segregate from the dust, but the dust will "float" on the sorbent particles, as was tested in an experiment with dust and sorbent of 2.5 mm. A size smaller than 2 mm leads to a mixing index that is too high according to the theory. In practice smaller particles seem to give a better segregation. So, the choice for the particle size will be based on economics.

The loss of sorbent with the bottom dust-removal is smaller than 1 % of the make-up sorbent flow. This means that it is not necessary to take this into account when developing a model.

Although the gas leaks between the different beds aren't that significant, they can be lowered by using special orifices instead of holes.
Other dependences on the particle size are shown in table 9.

Table 9. The particle size dependence of some parameters.

<table>
<thead>
<tr>
<th>Particle size (mm)</th>
<th>2.0</th>
<th>1.5</th>
<th>1.0</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>General</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Minimal Fluid. Gas velocity (m/s)</td>
<td>0.48</td>
<td>0.29</td>
<td>0.13</td>
</tr>
<tr>
<td><strong>Bed 1</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>$\alpha$ (-)</td>
<td>0.309</td>
<td>0.299</td>
<td>0.302</td>
</tr>
<tr>
<td><strong>Bed 2</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Mixing Index M (-)</td>
<td>0.006</td>
<td>0.103</td>
<td>0.204</td>
</tr>
<tr>
<td>Dust Fraction Top Layer (-)</td>
<td>0.0002</td>
<td>0.0031</td>
<td>0.0061</td>
</tr>
<tr>
<td><strong>Bed 3</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>$\alpha_1$ (-)</td>
<td>0.057</td>
<td>0.047</td>
<td>0.027</td>
</tr>
<tr>
<td>$\alpha_2$ (-)</td>
<td>0.926</td>
<td>0.935</td>
<td>0.931</td>
</tr>
<tr>
<td>$\alpha_3$ (-)</td>
<td>0.017</td>
<td>0.018</td>
<td>0.042</td>
</tr>
<tr>
<td><strong>Bed 4</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Gas velocity (m/s)</td>
<td>0.486</td>
<td>0.293</td>
<td>0.133</td>
</tr>
<tr>
<td>Voidage (-)</td>
<td>0.39</td>
<td>0.40</td>
<td>0.41</td>
</tr>
<tr>
<td>Residence time (s)</td>
<td>244</td>
<td>241</td>
<td>238</td>
</tr>
<tr>
<td><strong>Gas leaks</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Leak bed 2→3 (vol.%)</td>
<td>0.1</td>
<td>4.9</td>
<td>0.1</td>
</tr>
<tr>
<td>Leak bed 4→1 (vol.%)</td>
<td>2.6</td>
<td>3.2</td>
<td>5.4</td>
</tr>
</tbody>
</table>
Steam cycle

The medium used to convert heat into electricity is steam. The steam used in this design has a temperature of 500 °C and a pressure of 100 bar. These design conditions are analogous to Verhoeff (1988).

Generating electricity with steam is done by using turbines. The highest yields are begotten when using more than one expansion. In this case, two expansions are used, one in a high pressure turbine, the other in a low pressure turbine.

The total power that has to generated must exceed 100 MW. This capacity corresponds to the electricity usage of a community of 45000 heads. The actual power generated in this design is 118.485 MW. The turbines operate with a efficiency of 86%.

The produced heat in the combustor has to be transferred to water (see appendix III.3). This water comes from a feed water drum V 42. After leaving the drum, the water which is at a pressure of 0.12 bar, is compressed to 100 bar. This stream is split into two equal streams, the first is led through the economiser H 38, the other absorbs the heat in the off gases of beds three and four in the heat exchanger H 44. The two streams are joined and led through bed one in the heat exchanging tubes set H 37. Henceforth, it is fed to drum V 33.

Drum V 33 is used to separate the water from the steam. From it water flows through H 32 and the walls and grid of bed one (H 36), due to free circulation. Here the water is converted partly into steam.

The steam in drum V 33 is superheated to a temperature of 500 °C in the heat exchanger in the off gas of beds one and two, H 35. The superheated steam is fed to the first expander or turbine C 39. The pressure drops to 25 bar. The steam is then again superheated again to 500 °C with the off gases of beds one and two in the heat exchanger H 34. The stream's temperature is held on 500 °C by using an attemperator, which injects cold water into the steam stream. The attemperator is situated in between the two turbines. The resulting steam is fed to the second and last expander C 40. The pressure drops further to 0.12 bar.

The steam is then cooled to form water in condensor H 41. The water is led to drum V 42 which completes the cycle.

The thermic yield of a steam cycle lies between 30 % and 40 %. Rozemeijer and van der Sluis (1990) calculated a yield of 39 %. With a 100 % C-yield of the coal and combustion heat of 29.37 MJ per kg coal, the amount of coal needed can be calculated with:

\[ \text{equation 54} \]

\[ 117 \text{ MWu} / (29.37 \text{ MJ/kg} \times 0.39) \approx 10 \text{ kg coal/s} \]
As to be on the safe side, a coal stream of 10.4 kg/s has been chosen. This leads to a combustion heat of 10.4 * 29.37 MW or 305.448 MW. About 50 % of this heat produced in bed one is transferred to the pipes in bed one. Another 15 % goes to the walls and grid, the remaining heat stays in the off gases or goes into the regeneration reactions of bed three (Verhoeff, 1988).

The off gases of beds one and two are led through two heat exchangers H34 and H35 and finally through an economiser H38. The temperature of the gases are now approximately 190 °C. Henceforth, the gases is roughly cleaned in some cyclones and before entering the bagfilters, they preheat the combustion air used in bed one and two. The temperature now drops to 120 °C, as the gases are led to the stack.

The off gases of beds three and four are led through one heatexchanger, H44. Hereupon following the gases preheat the nitrogen and nitrogen-hydrogen mixture used in beds three and four, before being cleaned and led to a Claus plant.

Properties of the in and out going streams for each apparatus used in the steam cycle are given in table 10. These tsreams concern the water or steam stream. The amount of steam entering the turbines and circulating the cycle is 91.39 kg/s.

Table 10. The properties of in and out going steam or water streams for each apparatus in the steam cycle.

<table>
<thead>
<tr>
<th>Apparatus</th>
<th>T_{in} (°C)</th>
<th>T_{out} (°C)</th>
<th>P_{in} (bar)</th>
<th>P_{out} (bar)</th>
<th>M (kg/s)</th>
<th>Q (kW)</th>
</tr>
</thead>
<tbody>
<tr>
<td>H 32, bed 1</td>
<td>311</td>
<td>311</td>
<td>100</td>
<td>100</td>
<td>fc</td>
<td>136667</td>
</tr>
<tr>
<td>H 34, off gas</td>
<td>312</td>
<td>500</td>
<td>25</td>
<td>25</td>
<td>91.39</td>
<td>38611</td>
</tr>
<tr>
<td>bed 1+2</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>H 35, off gas</td>
<td>311</td>
<td>500</td>
<td>100</td>
<td>100</td>
<td>91.39</td>
<td>33333</td>
</tr>
<tr>
<td>bed 1+2</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>H 36, walls</td>
<td>311</td>
<td>311</td>
<td>100</td>
<td>100</td>
<td>fc</td>
<td>45833</td>
</tr>
<tr>
<td>bed 1+2+3+4</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>H 37, bed 1</td>
<td>96</td>
<td>233</td>
<td>100</td>
<td>100</td>
<td>91.39</td>
<td>55556</td>
</tr>
<tr>
<td>H 38, off gas</td>
<td>52</td>
<td>135</td>
<td>100</td>
<td>100</td>
<td>45.69</td>
<td>15972</td>
</tr>
<tr>
<td>bed 1+2</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>C 39, expander</td>
<td>500</td>
<td>312</td>
<td>100</td>
<td>25</td>
<td>91.39</td>
<td>-30780</td>
</tr>
<tr>
<td>1</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>C 40, expander</td>
<td>500</td>
<td>49</td>
<td>25</td>
<td>0.12</td>
<td>91.39</td>
<td>-87705</td>
</tr>
<tr>
<td>2</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>H 41, condenser</td>
<td>49</td>
<td>49</td>
<td>0.12</td>
<td>0.12</td>
<td>91.39</td>
<td>-209722</td>
</tr>
<tr>
<td>P 43, pump</td>
<td>49</td>
<td>52</td>
<td>0.12</td>
<td>100</td>
<td>91.39</td>
<td>1260</td>
</tr>
<tr>
<td>H 44, off gas</td>
<td>52</td>
<td>57</td>
<td>100</td>
<td>100</td>
<td>45.69</td>
<td>853</td>
</tr>
<tr>
<td>bed 3+4</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

fc is free circulation
Looking at that table and using the figures from it, the thermic yield of the cycle can be calculated. The total amount of energy entering and exiting the cycle, is equal to the amount of heat used in the turbines plus the amount of heat removed in condensor H 41. This gives a total of 328.207 MWₘₑ. The heat used in the turbines equals 118.485 MWₘₑ. This leads to the overall thermic yield:

\[ \frac{118.485}{328.207} \times 100\% = 36.1\% \]

The difference of the total heat removed, which is 328.207 MWₘₑ and the heat produced by combustion, which is 305.448 MWₘₑ is caused by the preheating of the combustion air, the nitrogen-hydrogen mixture and nitrogen pure, before it enters the combustor or other beds. This preheating is partly done with the heat still left in the stack gases and partly by firing with natural gas.

The dimension and characteristics for each heat exchanger is given in table 11.

Table 11. The dimensions and characteristics for each heat exchanger used in the steam cycle.

<table>
<thead>
<tr>
<th>Apparatus</th>
<th>Q (kW)</th>
<th>DeltaTₘₑ (°C)</th>
<th>U (kW/m²°C)</th>
<th>A (m²)</th>
<th>N (-)</th>
<th>Dₑ (m)</th>
</tr>
</thead>
<tbody>
<tr>
<td>H 32</td>
<td>136667</td>
<td>540</td>
<td>0.18</td>
<td>1406</td>
<td>4640</td>
<td>1.72</td>
</tr>
<tr>
<td>H 34</td>
<td>38611</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>H 35</td>
<td>33333</td>
<td>301</td>
<td>0.06</td>
<td>1846</td>
<td>6092</td>
<td>1.95</td>
</tr>
<tr>
<td>H 36</td>
<td>45833</td>
<td>540</td>
<td>0.05</td>
<td>1698</td>
<td>5604</td>
<td>1.88</td>
</tr>
<tr>
<td>H 37</td>
<td>55556</td>
<td>683</td>
<td>0.18</td>
<td>452</td>
<td>1492</td>
<td>1.00</td>
</tr>
<tr>
<td>H 38</td>
<td>15972</td>
<td>151</td>
<td>0.06</td>
<td>1763</td>
<td>5818</td>
<td>1.90</td>
</tr>
<tr>
<td>H 41</td>
<td>-209722</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>H 44</td>
<td>853</td>
<td>375</td>
<td>0.10</td>
<td>22.7</td>
<td>75</td>
<td>0.265</td>
</tr>
</tbody>
</table>

With the thermal efficiency of the cycle known, it is possible to calculate the amount of coal needed to produce exactly 100 MWₑ. The amount of heat added through preheating can easily be calculated by subtracting the amount of heat produced through combustion from the total amount of heat removed from the cycle. This leads to the following equation:

\[ 328.207 \text{ MW}_\text{ₑ} - 305.448 \text{ MW}_\text{ₑ} = 22.759 \text{ MW}_\text{ₑ} \]
The heat needed from the combustion can be calculated using:
(equation 57)

\[
\frac{100.000\, MW_{th}}{HEAT_{combustion\, MW_{th}} + 22.759 \, MW_{th}} = 0.361
\]

This amounts to a heat needed of 254.249 MW. The combustion heat of coal is 29.370 MJ per kg coal, so this leads to the quantity of coal needed per second of 8.7 kg.
Stack gas cleaning

Before the gases are led to the stack, the dust particles have to be removed. To do so, inertial separators, cyclones and bag filters are used, operating in such a way that the environmental requirement for dust of 20 mg/m³ is met.

Inertial separators

A sudden change of direction of the flow causes particles to separate from the main gas stream. Due to their inertia, they continue to move in the original direction. The power requirements of such a separator are low, but the efficiency isn't outstanding. Therefore it is used as a first step in a chain of gas cleaning apparatuses.

Cyclones

Cyclones find an extensive application in the field of gas cleaning, where they are applied for the separating of relatively coarse dusts. The advantages of the use of cyclones include low running costs, relatively low capital cost, reliability in use and suitability for higher temperatures.

There are two main designs for a cyclone: the reverse-flow cyclone and the "uniflow" cyclone. The former will be used.

The reverse-flow cyclone consists of a cylindrical section joined to a conical section and the clean gas outlet is through a pipe which extends some distance axially into the cyclone body through the top (see figure 3.). Under the cyclone a discharge hopper is installed to remove the dust particles.

In this type of cyclone the inlet gas is brought tangentially into a cylindrical section and a strong vortex is thus created. Particles in the flow are subjected to centrifugal forces which move them radially outwards, towards the inside cyclone surface on which the solids separate.

Many different standard geometries for the design of a cyclone are given in the literature as can be seen in table 12. The letters correspond with the letters seen in figure 3. All the values are based on the diameter D of 1 m.
Table 12. An overview of the different standard geometries for cyclones given in the literature.

<table>
<thead>
<tr>
<th>Source</th>
<th>A</th>
<th>B</th>
<th>C</th>
<th>E</th>
<th>F</th>
<th>L</th>
<th>K</th>
<th>M</th>
<th>ksi</th>
</tr>
</thead>
<tbody>
<tr>
<td>Stairmand HE</td>
<td>4.0</td>
<td>2.5</td>
<td>1.5</td>
<td>0.38</td>
<td>0.5</td>
<td>0.2</td>
<td>0.5</td>
<td>0.5</td>
<td>320</td>
</tr>
<tr>
<td>Stairmand HR</td>
<td>4.0</td>
<td>2.5</td>
<td>1.5</td>
<td>0.58</td>
<td>0.88</td>
<td>0.38</td>
<td>0.75</td>
<td>0.75</td>
<td>46</td>
</tr>
<tr>
<td>Breuer</td>
<td>3.53</td>
<td>1.94</td>
<td>1.59</td>
<td>0.32</td>
<td>0.80</td>
<td>0.32</td>
<td>0.74</td>
<td>0.59</td>
<td>86</td>
</tr>
<tr>
<td>Swift HE</td>
<td>3.9</td>
<td>2.5</td>
<td>1.4</td>
<td>0.4</td>
<td>0.5</td>
<td>0.21</td>
<td>0.44</td>
<td>0.4</td>
<td>665</td>
</tr>
<tr>
<td>Swift HR</td>
<td>3.7</td>
<td>2.0</td>
<td>1.7</td>
<td>0.4</td>
<td>0.85</td>
<td>0.35</td>
<td>0.8</td>
<td>0.75</td>
<td>55</td>
</tr>
<tr>
<td>Swift GP</td>
<td>3.75</td>
<td>2.0</td>
<td>1.75</td>
<td>0.4</td>
<td>0.6</td>
<td>0.25</td>
<td>0.5</td>
<td>0.5</td>
<td>300</td>
</tr>
<tr>
<td>Lapple GP</td>
<td>4.0</td>
<td>2.0</td>
<td>2.0</td>
<td>0.25</td>
<td>0.63</td>
<td>0.25</td>
<td>0.5</td>
<td>0.5</td>
<td></td>
</tr>
</tbody>
</table>

HE = High Efficiency  
HR = High Flow rate  
GP = General purpose

Figure 3. Schematic diagram of a reverse-flow cyclone with tangential inlet: 1, gas inlet; 2, cylindrical part; 3, conical part; 4, gas outlet; 5, top cover.
The main gas that has to be cleaned is the off-gas of beds one and two. The flow rate of this off-gas is 531633 m³ gas per hour. As the efficiency of a cyclone drops when the flow rate exceeds 20000 m³ gas per hour (Fayed, 1984), a battery of 27 cyclones will have to be used. This means that each cyclone has a throughput of 19690 m³ per hour. The design of the cyclones is done according to the rules of Stairmand's high efficiency cyclone. The corresponding diameter D is calculated according to Fayed (1984) with:

\[ \frac{Q_t}{D^2} = 5500 \]

This gives diameter of 1.89 m per cyclone.

The cyclone that cleans the off-gas of beds three and four is dimensioned in the same manner. The flow rate through it is 5119 m³ gas per hour, so only one cyclone is needed. The diameter is 0.97 m. The temperature of these gases which is 190 °C is no problem for the cyclones.

**Bag filters**

As a final gas treatment, bag filters as well electrostatic precipitators can be used. The advantages of the filters over the precipitators are:
- The filters play a role in the retention of the sulphur dioxide, as the sorbent particles on the filter cloth adsorb it.
- The filters consume less electricity.

The disadvantages are:
- The filters can clog, if little particles nest in the cloth
- The filters can clog, because of the CaSO₄ in the gas which forms a cement like substance with moisture.

High dust loadings and high gas volumes can be treated in bag filters. Because cloth is used, only gases with a temperate temperature can be treated. In this case the gas has a temperature of 120 °C. Some fibres of which the filter cloth can be made of are listed in table 13. as well as the highest temperature they can withstand.
Table 13. The characteristics of different fibres used as filter cloth in a bag filter.

<table>
<thead>
<tr>
<th>Fibres</th>
<th>Max. Temperature</th>
</tr>
</thead>
<tbody>
<tr>
<td>Cotton</td>
<td>80 °C</td>
</tr>
<tr>
<td>Wool</td>
<td>95 °C</td>
</tr>
<tr>
<td>Nylon</td>
<td>95 °C</td>
</tr>
<tr>
<td>Polypropylene</td>
<td>95 °C</td>
</tr>
<tr>
<td>Polyester</td>
<td>135 °C</td>
</tr>
<tr>
<td>Acrylic fibre</td>
<td>135 °C</td>
</tr>
<tr>
<td>Aramid</td>
<td>220 °C</td>
</tr>
<tr>
<td>Teflon</td>
<td>230 °C</td>
</tr>
<tr>
<td>Glass fibre</td>
<td>300 °C</td>
</tr>
<tr>
<td>Mineral-fibre needle felts</td>
<td>320 °C</td>
</tr>
<tr>
<td>Stainless-steel needle felts</td>
<td>450 °C</td>
</tr>
</tbody>
</table>

Based on the table and on the temperature, a cloth made of polyester or acrylic fire should be used. But, because of the corrosive characteristics of the off-gas, a glass fibre cloth will be used.

The design of a bag filter is a specialistic task, based on experience. Therefore, only a shallow treatment of the design of a filter will be given here.

One of the key parameters of the design of a bag filter is the air-to-cloth-ratio. This is the amount of air per unit cloth area that can be treated. It depends on the type of filter. These types differ in the way the dust cake is removed from the cloth. A pulse-jet unit (see figure 4) is chosen, because it has the highest air to cloth ratio.

With this type, the dust cake is removed periodically by an air pulse of high pressure going through the cloth in the opposite direction of the normal operating way. The air to cloth ratio of this type is about 2.5 m³/m²*min.

Two different filters have to be used, one which treats the off-gas of beds one and two, the other of bed three and four. The former has to have a cloth area of 131.3 m² (531633 m³/hr with a temperature of 120 °C), the latter an area of 34.1 m² (5119 m³/hr with a temperature of 120 °C).
Figure 4. Schematic diagram of a pulse-unit bag filter
Gas distributor design

The gas must be distributed uniformly into the fluidized bed without causing attrition of the bed material. This can be done by using grid plates, containing enough holes to ensure a good uniform aeration of the bed. The design of these grids is done following the rules given by Pell (1990).

Grid pressure drop

Due to the bubbling of the gas, there are constantly small changes in the local pressure drops at the grid. Because gas tries to get through the grid at places where the pressure drops are the lowest, the fluidization will lose its uniformity. To prevent this, the pressure drop over the grid has to be large enough to overcome these disturbances.

Furthermore, when the bed is at rest, most of the holes will be blocked by particles. During start-up, the initial gas flow will blow out some of the holes. Others remain blocked. The gas prefers to go through the unblocked holes and again the fluidization is not uniform. So the pressure drop has to be large enough to clear all the holes.

In practice the pressure drop is set by operating experience. When the pressure drop is a cost factor, the minimal pressure drop that has to be used is 0.3 times the pressure drop over the bed. For the design of the interconnected fluid bed combustor, this leads to the values, as shown in table 14. The pressure drops over the beds are calculated, using the next equation:

\( \Delta P_{\text{bed}} = (1-\varepsilon) \times \rho_s \times H \times g \)

Table 14. The pressure drops over the bed and over the grid, given for each bed separately.

<table>
<thead>
<tr>
<th>Bed number</th>
<th>Delta P bed</th>
<th>Delta P grid</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>24563 Pa</td>
<td>7369 Pa</td>
</tr>
<tr>
<td>2</td>
<td>32962 Pa</td>
<td>9900 Pa</td>
</tr>
<tr>
<td>3</td>
<td>31863 Pa</td>
<td>9600 Pa</td>
</tr>
<tr>
<td>4</td>
<td>32962 Pa</td>
<td>9900 Pa</td>
</tr>
</tbody>
</table>

Using these pressure drops, the orifice gas velocity can be calculated with the following equation:

\( u_o = 0.8 \times \left( \frac{2 \times g \times \Delta P_{\text{bed}}}{\rho_f} \right)^{0.5} \)
Using this hole velocity and the pressure drops, the total hole area can be calculated by:

\[ A_o * N_o = \frac{\Phi_{gas}}{u_o} \]

When the holes are triangularly pitched as is the fact in this design, a relation between the hole diameter and pitch, \( p \), is given as shown in the next equation:

\[ d_o^2 * \frac{2}{\sqrt{3} * p^2} = \frac{\Phi_{gas}}{u_o} * \frac{16}{\pi^2 * d_t^2} \]

Using holes of a diameter, \( d_o \), of 5 mm, leads to the number and pitches as shown in Table 15.

**Table 15. The grid characteristics as calculated for each bed.**

<table>
<thead>
<tr>
<th>Bed</th>
<th>Hole velocity (m/s)</th>
<th>( A_o*N_o ) (m²)</th>
<th>number of holes (-)</th>
<th>Tank diameter (m)</th>
<th>Pitch (mm)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>404</td>
<td>0.89</td>
<td>45383</td>
<td>10.7</td>
<td>48</td>
</tr>
<tr>
<td>2</td>
<td>468</td>
<td>7.6e-3</td>
<td>382</td>
<td>2.52</td>
<td>12.3</td>
</tr>
<tr>
<td>3</td>
<td>461</td>
<td>6.5e-3</td>
<td>332</td>
<td>1.60</td>
<td>84</td>
</tr>
<tr>
<td>4</td>
<td>468</td>
<td>5.2e-3</td>
<td>265</td>
<td>0.80</td>
<td>47</td>
</tr>
</tbody>
</table>
Safety

An old Dutch saying goes: "To avoid is better than to cure". So, during the design of a plant, it is wise to take safety along. Three methods of preventing unsafe situations during design exist:

1. Organizing measures, such as laws, rules, checklists and a procedure for a safety analysis.

2. System analytic research, such as a process safety analysis, a disturbance analysis, risk analysis. These lead to organizing measures as well as technological measures, like regulators, safety valves, sprinkles and so on.

3. Physical-chemical research of the behaviour and characteristics of the substances used.

Organizing measures

A law which affects the design and operation of the power station is the Steam Law. All apparatuses that make or use steam have to be passed by an institute called "Dienst voor het Stoomwezen". This institute checks newly-build steam-using plants as well as existing plants. Also, apparatuses that are under pressure are inspected by the institute, although it's not obligatory. The influence of this institute is felt already during the design-stage, for the designs have to be approved by it.

Safety analysis

A safety analysis has to be done as part of work safety report required by the Government as is written in a law concerning labour, the so-called ARBO-law. This safety analysis contains the steps shown in figure 5.

![Figure 5. A safety analysis diagram.](image-url)
System analytic research
Doing a process safety analysis, is doing a systematic re­search of the behaviour of the process, as a function of its parameters, like pressure, temperature, substances and so on. The target is to find out what the safety limits of the process are. During the analysis, the search of potential hazards that may lead to greater hazards, like fire and explosions, stands central. Potential hazards in the design of the combustor are the gas leaks through the orifices between the beds, steam blanketing and dust explosions.

Gas leaks
Through the orifice between beds two and three gas leaks, caused by the sorbent flow. The gas leak is minor, so no explosion danger will exist. Only if one of the two fluidization gas flows stops, a explosive mixture may form. For this reason the other flow should also be stopped, thus reducing the explosion danger. The gas leak between beds one and four doesn’t lead to a dangerous situation, because bed four is fluidized by the inert gas Nitrogen.

Steam Blanketing
Steam blanketing is the separation of the water and the steam produced, in the pipes of the heat exchanger. The water will flow on the bottom of the pipe, while the steam flows on top of it. This causes salts to crystallize on the upper part of the pipe wall, which leads to corrosion. Steam blanketing occurs when the flow through the pipes is too low. Because of the special design of a fluidized bed combustor, rules for avoiding steam blanketing differ from rules concerning steam blanketing in conventional combustors. It might occur in the heat exchanger in bed 1, because of the great heat flow combined with the great length and diameter of the pipes. Care should be taken that the flow through these pipes is high enough (Verhoeuff, 1988).

Dust Explosions
There are some apparatuses in the design where the potential hazard of a dust explosion exits. These are the hoppers in which the coal is stored and the cyclones used for the cleaning of the stack gases.

In the coal hoppers, a hazardous mixture of dust and air may form during the filling of the hoppers. An explosion may be ignited by static electricity. By grounding all the contact surfaces, ignition may be prevented. If an explosion occurs after all, it can be suppressed with a build-in extinguisher such as shown in figure 4. The cyclones and bag filters are used for cleaning the air coming from the combustor. In the air, some hot particles may occur. These could serve as an ignition source for an eventual dust explosion. So they have to be extinguished. This can be done by using an infrared detector combined with a spraying system.
Cost analysis

Investment
Many methods that serve the purpose of assessing the investment involved in building a plant, have been designed for the chemical industry. Better is it, to compare a plant with an existing one and adjusting the investment costs for the new plant. In this case, an atmospheric fluidized bed boiler with sulphur retention with limestone, was used. This plant was delivered in 1988 at AKZO in Hengelo. This installation has a thermic power of 90 MW, with a kettle capacity of 115 tons steam per hour. Van Heek (1988) gives a summary of the investments done since 1983. In the AKZO plant the fluidized bed has an area of 96 m². In the design of the interconnected fluid bed combustor, beds two, three and four will be built into bed one. This leads to a total area of 97 m². So the investments are comparable. Here follows a list of the investment needed for the interconnected fluid bed system. They have been adjusted for inflation and alike, so they differ from those for the AKZO plant that are given between brackets:
(The amounts of money are given in Dutch Guilders)

- Combustor 47 mln (41 mln)
- Fuel supply 11 mln (10 mln)
- Electronics and instrumentation 5 mln (4 mln)
- Pipelines and water supply 5 mln (4 mln)
- Engineering 9 mln (8 mln)
- Pumps and compressors 81 mln (- mln)
  (800 DG/kW)
- Sorbent 2 mln (- mln)
  (280 tons at 6096 DG/ton (van den Bleek et al., 1991))
- Remaining 3 mln (3 mln)

Investment reality 163 mln (70 mln)

Plus:
- Work capital (6%) 10 mln
- Know-how, start up (14%) 23 mln

Total investment 196 mln

Production costs
Operating the combustor brings along extra costs. A list of these costs is given below. All the costs given are per year.
- The maintenance takes 4% of the reality investment, which gives 0.04 * 163 mln = 6.5 mln.
- The depreciation of the investment will be done in ten years. So this leads to 0.1 * (total investment-work capital) = 18.6 mln.
- The Stein relation calculates that 336 employees are needed to operate the plant. This equals 61 places * 350000/place = 21.4 mln.
- For the cooling water and the steam, 144 million tons of river water is used at 6 cents a ton. This is 8.7 mln.
- In a year operation, 299520 tons of coal is used at 139 DG a ton, which gives 41.6 mln.
- Sorbent disappears into the cyclones and filters, because of attrition. A make-up sorbent flow goes into the combustor. This is 786 tons sorbent a year at 6096 DG a ton, so 4.8 mln.
- The ash produced by the combustion has to be removed. This costs 35.36 DG a ton. 31636 tons a year are produced so the cost is 1.1 mln.
- The electricity used to operate the system costs 16 cents a kWh. Per year 88 million kWh are needed for the pumps, compressors and other equipment. This leads to 14 mln.
- To warm the fluidization gases, natural gas is used. This costs 6.64 DG a gJ. As 690000 gJ is needed, this costs 4.6 mln.
- The nitrogen used to fluidize beds three and four costs 0.3 DG a m³, at a use of 13581600 m³ a year gives 4.1 mln.
- Hydrogen is used for the regeneration of the sorbent. The price of hydrogen is 1800 DG a ton. As 175.9 tons a year are needed, this costs 0.3 mln.

The total operation costs during the first ten years is 125.7 million Dutch Guilders a year.

**Cost evaluation**

The yield is obtained from the electricity produced which is sold at a price of 16 cents a kWh. The power of the plant is 118 MW. A plant operates about 8000 hours a year, so the production of this plant is $9.6 \times 10^8$ kWh. So the yield is 151 million Dutch Guilders a year.

As the operating costs amount to 125.7 million, the profit per year is 25.3 million Dutch Guilders. With this amount the pay out time can be calculated:

Pay out time = \frac{Total investment - Work capital}{Profit surplus cash}

Using this equation, a pay out time of 7.4 years is calculated. The return on investment is calculated with:

\[ \text{ROI} = \frac{\text{Net profit}}{\text{Total investment}} \]

This amounts to a return on investment of 12.9 %

\[ \frac{25.3}{125.7} = \text{bruto profit before taxes} \]

\[ \frac{196.0}{(\star)} \]

\[ (\star) \text{ wrong definition.} \]
The interconnected fluid bed combustor can be designed according to the plan in appendix III. Some characteristics of the design are listed below.

- The bed height is 4 m.
- The beds have a comparatively small total area of 97 m².
- The sulfur retention in bed 1 is between 99.22 and 99.80 %.
- The off gas of beds three and four, that is led to a Claus plant, contains 146.28 g/m³.
- The SO₂-concentration in the stack gas, is with 13.14 mg/m³ well below the regulation of 700 mg/m³ and with 16.32 g/GJ well below 600 g/GJ.

The only requirement that is not met, is that the residence time in bed one is fixed by the condition that 95% of the coal reacts in bed one. Following this condition, leads to a particle residence time in bed one, that is too small, to effectively adsorb the sulphur in the gases. For this reason, this condition has been dropped, as the regulation for the sulphur oxides in the stack gas has been considered of higher importance.

The power generated with the calculated configuration of the ICFBC is 118 MW. The coal used to generate such a power is 10.4 kilograms per second. If a power of 100 MW, is wanted, only 8.7 kilograms coal per second is needed. In that case the combustor bed area can be reduced.

All the off gases are led through a series of cyclones and bag filters. In total 28 cyclones are needed and a total cloth area for the bag filters of 165.4 m². This way of cleaning the off gases is supposed to be sufficient to meet the regulation for dust in stack gas of 20 mg/m³.

The calculated investment involved in building such an ICFBC, is 196 million Dutch Guilders. A cost analysis leads to an annual production cost estimation of 125.7 million Dutch Guilders. The yield is gotten from the sale of electricity which goes presently at 16 cents a kWh. With an operation of 8000 hours a year, this leads to an annual income of 151 million Dutch Guilders. This gives a profit of 25.3 million Dutch Guilders per annum. Henceforth, the pay out time can be calculated and it equals 7.35 years. This means a return on investment of 12.9 %.

All things considered, the ICFBC looks promising. It is possible to meet all the environmental regulations and making money at the same time. A return on investment of 12.9 % isn’t bad. And considering that the price of electricity will rise, due to the free market mechanisms, the ICFBC has a future.
### Symbols used

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Meaning</th>
<th>Unit</th>
</tr>
</thead>
<tbody>
<tr>
<td>A</td>
<td>surface area</td>
<td>m²</td>
</tr>
<tr>
<td>Cₓ</td>
<td>concentration of compound x</td>
<td>mg/m³</td>
</tr>
<tr>
<td>D</td>
<td>diameter cyclone</td>
<td>m</td>
</tr>
<tr>
<td>d</td>
<td>diameter</td>
<td>m</td>
</tr>
<tr>
<td>dₑq</td>
<td>help variable</td>
<td>m</td>
</tr>
<tr>
<td>dt</td>
<td>diameter tank</td>
<td>m</td>
</tr>
<tr>
<td>Dₓ₀ₓ,film</td>
<td>diffusion coefficient of SOₓ in film</td>
<td>m²/s</td>
</tr>
<tr>
<td>Dₓ₀ₓ,shell</td>
<td>diffusion coefficient of SOₓ in shell</td>
<td>m²/s</td>
</tr>
<tr>
<td>g</td>
<td>acceleration of gravity, 9.8 m/s²</td>
<td>m/s²</td>
</tr>
<tr>
<td>H</td>
<td>bed height, 4 m</td>
<td>m</td>
</tr>
<tr>
<td>h</td>
<td>height</td>
<td>m</td>
</tr>
<tr>
<td>Kₑq</td>
<td>equilibrium constant of SO₂ + 1/2 O₂, &lt;--→ SO₃</td>
<td>(m²/mole)⁴</td>
</tr>
<tr>
<td>kₛ</td>
<td>reaction rate constant of SO₃ on core</td>
<td>m³ gas/(m² core s)</td>
</tr>
<tr>
<td>kₓ₀ₓ,film</td>
<td>SOₓ film mass transport coefficient</td>
<td>m/s</td>
</tr>
<tr>
<td>Lₚₑd</td>
<td>distance over which pressure drop P₀ is taken</td>
<td>m</td>
</tr>
<tr>
<td>Ly</td>
<td>Lyaschenko number</td>
<td>--</td>
</tr>
<tr>
<td>M</td>
<td>mixing index</td>
<td>--</td>
</tr>
<tr>
<td>N</td>
<td>number of ...</td>
<td>--</td>
</tr>
<tr>
<td>n</td>
<td>mean number of cycli of sorbent part.</td>
<td>--</td>
</tr>
<tr>
<td>P</td>
<td>pressure</td>
<td>bar</td>
</tr>
<tr>
<td>q</td>
<td>number of moles Ca per square meter particle surface</td>
<td>mole/m²</td>
</tr>
<tr>
<td>Qₛ</td>
<td>gas throughput through cyclone</td>
<td>m³/s</td>
</tr>
<tr>
<td>R</td>
<td>sorbent particle radius</td>
<td>m</td>
</tr>
<tr>
<td>rₛ</td>
<td>unreacted core radius</td>
<td>m</td>
</tr>
<tr>
<td>Reₚ</td>
<td>Reynolds number</td>
<td>--</td>
</tr>
<tr>
<td>Sₓ</td>
<td>surface of compound x, c =char</td>
<td>m²</td>
</tr>
<tr>
<td>Sc</td>
<td>Schmidt number</td>
<td>--</td>
</tr>
<tr>
<td>St</td>
<td>Stokes number</td>
<td>--</td>
</tr>
<tr>
<td>t</td>
<td>time</td>
<td>s</td>
</tr>
<tr>
<td>u</td>
<td>gas velocity</td>
<td>m/s</td>
</tr>
<tr>
<td>uₛ</td>
<td>help variable</td>
<td>m/s</td>
</tr>
<tr>
<td>Uₛₜₜ</td>
<td>absolute particle velocity through orifice</td>
<td>m/s</td>
</tr>
<tr>
<td>Uᵣ</td>
<td>superficial gas velocity</td>
<td>m/s</td>
</tr>
<tr>
<td>U_gas</td>
<td>gas velocity in bed going up</td>
<td>m/s</td>
</tr>
<tr>
<td>Uₜ</td>
<td>terminal fluidization velocity</td>
<td>m/s</td>
</tr>
<tr>
<td>U_TOₛₜₜ</td>
<td>help variable</td>
<td>m/s</td>
</tr>
<tr>
<td>U_TO</td>
<td>take over velocity, the velocity at which the mixing becomes predominant</td>
<td>m/s</td>
</tr>
<tr>
<td>Vₛ</td>
<td>gas volume in a tank</td>
<td>m³</td>
</tr>
<tr>
<td>Vₛₜₜ</td>
<td>reactor volume</td>
<td>m³</td>
</tr>
<tr>
<td>x</td>
<td>ash fraction</td>
<td>--</td>
</tr>
<tr>
<td>Y</td>
<td>dimensionless core radius fraction</td>
<td>--</td>
</tr>
<tr>
<td>z</td>
<td>velocity ratio for mixing</td>
<td>--</td>
</tr>
</tbody>
</table>
### Greek symbols

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Meaning</th>
<th>Unit</th>
</tr>
</thead>
<tbody>
<tr>
<td>(\alpha)</td>
<td>surface (\text{CaSO}_4) / surface (\text{Ca}) (bed 1)</td>
<td>--</td>
</tr>
<tr>
<td>(\alpha_1)</td>
<td>surface (\text{CaSO}_4) / surface (\text{Ca}) (bed 3)</td>
<td>--</td>
</tr>
<tr>
<td>(\alpha_2)</td>
<td>surface (\text{CaO}) / surface (\text{Ca}) (bed 3)</td>
<td>--</td>
</tr>
<tr>
<td>(\alpha_3)</td>
<td>surface (\text{CaS}) / surface (\text{Ca}) (bed 4)</td>
<td>--</td>
</tr>
<tr>
<td>(\alpha_{\text{SO}_x})</td>
<td>help variable</td>
<td>--</td>
</tr>
<tr>
<td>(\Gamma)</td>
<td>stoichiometric coefficient = 1</td>
<td>mole (\text{CaO})/mole (\text{SO}_3)</td>
</tr>
<tr>
<td>(\delta)</td>
<td>help variable</td>
<td>--</td>
</tr>
<tr>
<td>(\epsilon)</td>
<td>bed voidage</td>
<td>--</td>
</tr>
<tr>
<td>(\mu)</td>
<td>viscosity</td>
<td>Ns</td>
</tr>
<tr>
<td>(\rho)</td>
<td>density</td>
<td>(\text{kg/m}^3)</td>
</tr>
<tr>
<td>(\phi)</td>
<td>mass flow</td>
<td>(\text{kg/s})</td>
</tr>
<tr>
<td>(\phi_{\text{GL}})</td>
<td>gas leak</td>
<td>(\text{m}^3/\text{s})</td>
</tr>
<tr>
<td>(\phi_{\text{sphere}})</td>
<td>sphericity factor = 0.95</td>
<td>--</td>
</tr>
<tr>
<td>(\tau_x)</td>
<td>residence time in bed (x)</td>
<td>s</td>
</tr>
</tbody>
</table>

### Subscripts

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Meaning</th>
</tr>
</thead>
<tbody>
<tr>
<td>ashes</td>
<td>ashes or dust</td>
</tr>
<tr>
<td>(b, b)</td>
<td>bed</td>
</tr>
<tr>
<td>coal</td>
<td>coal</td>
</tr>
<tr>
<td>(f)</td>
<td>fluid, usually gas</td>
</tr>
<tr>
<td>gas</td>
<td>gas</td>
</tr>
<tr>
<td>(\text{,hor})</td>
<td>flowing horizontally</td>
</tr>
<tr>
<td>(\text{,ing})</td>
<td>ingoing</td>
</tr>
<tr>
<td>make up</td>
<td>make up sorbent flow</td>
</tr>
<tr>
<td>mf</td>
<td>at minimal fluidization conditions</td>
</tr>
<tr>
<td>(\text{,min})</td>
<td>minimal</td>
</tr>
<tr>
<td>o</td>
<td>concerning the orifices in the grid</td>
</tr>
<tr>
<td>orifice</td>
<td>concerning the orifice between beds</td>
</tr>
<tr>
<td>(p, part)</td>
<td>particle</td>
</tr>
<tr>
<td>s</td>
<td>sorbent or solid</td>
</tr>
<tr>
<td>sorbent</td>
<td>sorbent</td>
</tr>
</tbody>
</table>

When a symbol is overlined, the mean value is meant.
Literature


Bleek, C.M. van den & Gerritsen, A.W., dictaat Reactorkunde II (st75), Vakgroep Chemische Technologie TU Delft (1988)


DGM, Overzicht van emissie-eisen na herziening Bees WLV Kopie eisen VROM, (1992)


Heek, P.H.G. van, Realisatie wervelbedketel K10 Hengelo, PT Proces techniek, 43, nr. 10, p. 31-35 (1988)

Horio, M., Shigekatsu, M., Muchi, I., A model study for the development of low NOx fluidized-bed coal combustors, 5th Proc. on 7th Int. FBC conference, (1977)

Hout, J. van, Kok, L., Een model voor de regeneratie van het sorbent gebruikt voor zw zwavelvangst bij de verbranding van steenkool, Eindopdracht reactorkunde II TU Delft (1992)


Lin, W., Valkenburg, P.J.M., Bleek, C.M. van den, Prediction of NOx and SOx emissions in FBC of coal using easy to determine coal and sorbent characteristics, Proc. 2nd Int. Rolduc Symp. on Coal Science, p. 399-405, Amsterdam, 1990


Nauze, R.D. La, Fundamentals of coal combustion in fluidised beds, CSIRO Division of Fossil Fuels, North Ryde, Australia, (1988)


Rozemeijer, F.A. & Sluis, F.J.P. van der, Een kolengestookte centrale op basis van een vier-cel interconnected fluïde bed systeem met regeneratieve zwavelvangst, Fabrieksvoortontwerp nr.: 2854, Vakgroep Chemische Procestechnologie, TU Delft (1990)

Schouten, J.C. and Bleek, C.M. van den, The D.U.T.-SURE-model: a simple approach in FBC sulfur retention modeling, Department of Chemical Engineering and Chemistry, TU Delft


Versteeg, R., Modellering van een regeneratieve zwavelvangst bij de wervelbed-verbranding van steenkool, afstudeerscriptie TU Delft, Vakgroep Chemische Procestechnologie (1990)
WEBCI/WUBO prijzenboekje, redactie Dutch Association of Cost Engineers, 14, (1989)

Extension

Introduction

With the models, worked out in previous chapters, some calculations have been made. These comprise calculations of trends, in order to publicize them. To make things easier, the models of beds one and three have been simplified. These are also the only models in addition of the model for the formation of nitrogen oxides in bed one, that have been used for the calculations. The models for beds two and four stay unchanged.

Calculation path for bed one

The model for bed one has very much been simplified. First of all, the actual coal feed needed to produce 100 MW, has been calculated. In the first design, a coal feed stream of 10.4 kg/s is used. This leads to a power generation of 118.485 MW.

Using these figures, the coal feed needed can be calculated in the following way:

\[\text{coal feed needed} = \frac{10.4 \times 100}{118.485}\]

This leads to a coal feed flow of 8.8 kg/s. Based on this the amount of air needed, can be calculated. For the combustion of 1 kilogram coal, 73.8 mole oxygen is needed or 351 mole air. One mole air takes 22.4 liters at room temperature. This has to be corrected for the temperature in the reactor. The factor for this is 1/0.23. Knowing all these figures, the minimal gas flow into the reactor can be derived. This amount is multiplied by a factor of 1.2, to ensure the 6 % oxygen in the stack gas as requested in the emission specifications set by the government. The total gas flow needed to operate the reactor with a coal flow of 8.8 kg/s is 361.158 m³/s air.

The next step is implementing the emission specification for sulphur oxides. Knowing this concentration and the gas flow, the maximum mole flow sulphur oxides into the stack can be calculated. Furthermore, the incoming mole flow sulphur oxides, arising from the coal combustion, is known. Combining these two, the sulphur retention in mole SO₂ per second is known. In the case of the model, a stack concentration for the sulphur oxides of 600 mg/m³ (at 20 °C) is taken, to be on the safe side.

As the SO₂ flow released by combustion is 2 mole/s and the maximum flow into the stack 0.778 mole/s, a sulphur retention of 62 % has to be reached. This means that 1.28 mole SO₂/s has to be captured by the sorbent. By choosing the sulphurization degree alpha, the flow of active calcium oxide is set. If the average activity of the sorbent is known, the actual sorbent flow can be calculated with the fact that 8.91 mass percent of the sorbent is taken by calcium.

This sorbent flow is the used along with the mole flow captured sulphur dioxide, to determine the volume of the bed, using the SURE2-model. A graph of this calculation path is shown in figure 6.
Figure 6. The calculation path for the model for bed one.
A problem that rises, using the forementioned path, is that this would only work if the regeneration of the sorbent is to be complete. Alas, this is not the case, but basically the calculation order in the model is the same.

Simplifications of the model for bed one

As was mentioned before, the model as described in the chapter Calculation models has been simplified. First of all the area and height of the bed have been replaced by using only the volume of the bed, which is actually calculated in the model, rather than being an input variable as before.

Because the feed gas flow has become an input variable, the gas velocity has been made obsolete. This is possible, when all the influences it has on the model, can be ruled out. The velocity first of all, influences the bed voidage. Models for the bed voidage however apply only for small beds and won't function properly in the case of the model. As the voidage lies between 0.4 and 0.6, a constant voidage of 0.55 has been chosen. This value is based on the value calculated in the calculations with the "old" model using a sorbent with a diameter of 2 mm.

The velocity also influences the help variable delta in the calculation of alfafilm. As this delta is a weak function of the velocity, attributing it a constant value will not introduce a significant error. The average value of 2.1 e-9 m has been assigned to it, based on calculations with the "old" model.

Correcting an error

Sad enough, the old model is not without error. In the calculation of the number of particles Npart, a misinterpretation of its definition has occurred. The model of Wolff states that it is the number of sorbent particles per m$^3$ reactor volume. In the SURE2 model, this value is multiplied by the volume itself, leading to the total number of particles in the reactor.

In the "old" model in this script however, Npart was interpreted as the total number of sorbent particles in the reactor. This caused the model to calculate the conditions that would arise if the number of m$^3$ reactor volume times the number of particles was being used. This leads to a small bed with surprising possibilities. This error has been corrected, but alas, the time fails to correct the mass balances as given in the appendices. No influence on the steam cycle is expected, when correcting this error, nor will the gas cleaning change.
Simplifications in the model for bed three

As in bed one, the bed voidage and gas feed flow have been made input variables. The values attributed to these variables are respectively 0.42 and 3 m³/s. The area and gas velocity are no longer variables, but the residence time in the bed has become an input variable. This value determines the regeneration of the sorbent, along with the volume of the reactor which is an output variable.

Models for the calculation of the average activity of the sorbent

In the "old" model, the average activity was calculated as follows:
(equation 64)

\[ \text{activity} = 0.997 \left( \frac{\text{tau}_{\text{system}}}{\text{t}_{\text{cycle}}} \right) \]

where \( \text{tau}_{\text{system}} = \) average residence time of sorbent in the system as a whole
\( \text{t}_{\text{cycle}} = \) the average time it takes for a sorbent particle to complete one cycle

Two new models have been added. In the first model the deactivation of the sorbent is attributed to just being in the system at such a high temperature. The values of the parameters in this model are based on an experiment on the deactivation of the sorbent. The cycle time of the sorbent was 4.3 hours and the deactivation was 1.03 promille over 10³ cycles. This leads to the following model:
(equation 65)

\[ \text{activity} = 1 - \frac{0.103}{10³ \times 4.3} \times \tau_{\text{system}} \]

The other model attributes the loss of activity to the changing of the surface occupation. So, it couples the deactivation to the number of cycles. This leads to this equation:
(equation 66)

\[ \text{activity} = 1 - \frac{0.103}{10³} \times \frac{\tau_{\text{system}}}{\text{t}_{\text{cycle}}} \]

The first model leads with the average residence time in the system of 1325 hours, to an average activity fraction of 0.69. This value is independent of the volumes of the beds or of anything else.
The second model however is influenced by the different volumes, through the residence times. Both models are used in the calculation of bed three, but the calculated value is iterated with bed one.

**How are the calculations made?**

There exist five models, for each bed one and for bed one even two. It is not necessary to iterate all models consequently to obtain a result. The calculation scheme used for the models is explained below and graphically given in figure 7.

**General input:**

particle sizes, densities and consequential minimal fluidization velocities for sorbent and ash.

**Bed one:**

input: $\phi_{\text{coal}}$, $\phi_{\text{gas}}$, $\alpha$, SO$_2$ concentration, $\varepsilon$

from other beds: $\tau_2$, $\tau_3$, $\tau_4$

output to NO$_x$ model: $\varphi_2$, $V$, $N_{\text{part}}$, O$_2$ concentration, $\tau_1$

$\phi_{\text{coal}}$, $\phi_{\text{gas}}$, $\alpha$

output to model bed2: $\phi_{\text{coal}}$, $\varphi_1$

output to model bed4: $\varphi_1$, $u_1$, $\varepsilon_1$

iteration with bed three:

output: $\varphi_1$, $\tau_1$

input: $\alpha_1$, $\alpha_2$, $\alpha_3$, average activity

Overall output: $\tau_1$, $V_1$, $\varphi_1$, $\phi_{\text{make-up}}$, O$_2$ concentration

**Bed two:**

input: $u_2$

from bed1: $\phi_{\text{coal}}$, $\varphi_1$

output: $\tau_2$ (also to bed1 and bed3), M, x, $\phi_{\text{gas}}$

**Bed three:**

input: $\tau_3$, $\phi_{\text{gas}}$, H$_2$ feed concentration, $\varepsilon_3$

iteration with bed 1:

input: $\varphi_1$, $\tau_1$

output: $\alpha_1$, $\alpha_2$, $\alpha_3$, $\tau_3$, average activity

**General output:** $\alpha_1$, $\alpha_2$, $\alpha_3$, average activity, $\tau_3$, SO$_2$, H$_2$O, H$_2$S, S$_2$, H$_2$ concentrations, V
Bed four:

input: $A_{\text{orf}}$, $h_{\text{orf}}$, $V_{\text{bed}}$

from bed1: $\phi_4$, $u_1$, $\varepsilon_1$

output: $r_4$ (also to bed1 and bed3), $u_4$, $\phi_{\text{gas}}$, $\phi_{\text{so}}$
Figure 7. The calculation scheme for all the five models and their in- and outputs
Calculations and evaluation

Using the newly constructed models, the influence of varying the sorbent charge, on beds one and three has been calculated. These calculations have been done, following the new models for the loss of activity.

Model 2:

The first calculations are done with the use of equation 65. The use of this model implicates that the average activity of the sorbent remains constant. As the total residence time of the sorbent equals 1325 hours the average activity becomes 0.69 or 69% of the Ca is active in the sulphur retention.

The influence of the varying sorbent charge \( \alpha \) is first calculated for the effect it has on the sulphur retention in bed one. The trends are shown in table 16.

Table 16. The influence of varying the sorbent charge on bed one.

<table>
<thead>
<tr>
<th>( \alpha ) (-)</th>
<th>( \phi_{\text{make-up}} ) (kg/s)</th>
<th>( V_1 ) (m³)</th>
<th>( \phi_{\text{sorbent}} ) (kg/s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.3</td>
<td>0.0766</td>
<td>1095</td>
<td>4.0047</td>
</tr>
<tr>
<td>0.4</td>
<td>0.1070</td>
<td>1562</td>
<td>2.9342</td>
</tr>
<tr>
<td>0.5</td>
<td>0.1446</td>
<td>2143</td>
<td>2.3014</td>
</tr>
<tr>
<td>0.6</td>
<td>0.1920</td>
<td>2881</td>
<td>1.8699</td>
</tr>
<tr>
<td>0.7</td>
<td>0.2559</td>
<td>3888</td>
<td>1.5552</td>
</tr>
<tr>
<td>0.8</td>
<td>0.3520</td>
<td>5419</td>
<td>1.3077</td>
</tr>
<tr>
<td>0.9</td>
<td>0.5312</td>
<td>8309</td>
<td>1.0814</td>
</tr>
</tbody>
</table>

Calculating the models with an \( \alpha \) smaller than 0.3 did not work, so these values have been left out. Looking at table 16, the conclusion may be drawn that it is best to operate with a charge that is as small as possible. By decreasing this charge, the sorbent stream increases, which is logical, as the amount of sulphur retained is constant. As the amount of sulphur retained is calculated by multiplying the charge with the sorbent stream and a constant, this trend is to be expected.

Furthermore, with decreasing charge, the volume decreases. This is caused by the fact that a smaller charge on the sorbent needs a smaller residence time. The residence time decreases with increasing sorbent flow and decreasing bed volume. The charging of the sorbent has a bigger effect than can be compensated by a increase in sorbent flow.
The next step is to look at the effect the variation of the charge has on the regeneration in bed three. The trends coming from these calculations are shown in table 17.

Table 17. The effect of a varying charge on the regeneration in bed three.

<table>
<thead>
<tr>
<th>α</th>
<th>α₁</th>
<th>α₂</th>
<th>α₃</th>
<th>V₃ (m³)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.3</td>
<td>0.0222</td>
<td>0.9061</td>
<td>0.0717</td>
<td>7.89</td>
</tr>
<tr>
<td>0.4</td>
<td>0.0267</td>
<td>0.8788</td>
<td>0.0945</td>
<td>5.78</td>
</tr>
<tr>
<td>0.5</td>
<td>0.0307</td>
<td>0.8534</td>
<td>0.1159</td>
<td>4.54</td>
</tr>
<tr>
<td>0.6</td>
<td>0.0342</td>
<td>0.8283</td>
<td>0.1375</td>
<td>3.69</td>
</tr>
<tr>
<td>0.7</td>
<td>0.0371</td>
<td>0.8034</td>
<td>0.1596</td>
<td>3.06</td>
</tr>
<tr>
<td>0.8</td>
<td>0.0354</td>
<td>0.7778</td>
<td>0.1891</td>
<td>2.58</td>
</tr>
<tr>
<td>0.9</td>
<td>0.0391</td>
<td>0.7430</td>
<td>0.2179</td>
<td>2.13</td>
</tr>
</tbody>
</table>

Once again, the results show that it is best to operate with a charge that is as small as possible. These calculations have been done with a constant residence time for the particles in bed three. The residence time remained at a value of 1600 s. This explains why the bed volume increases with decreasing charge or increasing sorbent flow.

The table also shows that the regeneration is better with decreasing charge, which is to be expected. Here, the influence of the charge on the CaO fraction α₂ is meant.

The third effect registered is the effect on the stack gases in bed three. These are shown in table 18.
Table 18. The effect of the varying charge on the stack gases.

<table>
<thead>
<tr>
<th>$\alpha$ (-)</th>
<th>$SO_2$ (mol/m$^3$)</th>
<th>$H_2S$ (mol/m$^3$)</th>
<th>$S_2$ (mol/m$^3$)</th>
<th>$H_2O$ (mol/m$^3$)</th>
<th>$H_2$ (mol/m$^3$)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.3</td>
<td>0.4001</td>
<td>0.0071</td>
<td>0.0080</td>
<td>1.0581</td>
<td>0.0059</td>
</tr>
<tr>
<td>0.4</td>
<td>0.3861</td>
<td>0.0097</td>
<td>0.0118</td>
<td>1.0548</td>
<td>0.0066</td>
</tr>
<tr>
<td>0.5</td>
<td>0.3731</td>
<td>0.0122</td>
<td>0.0158</td>
<td>1.0517</td>
<td>0.0072</td>
</tr>
<tr>
<td>0.6</td>
<td>0.3554</td>
<td>0.0149</td>
<td>0.0202</td>
<td>1.0483</td>
<td>0.0078</td>
</tr>
<tr>
<td>0.7</td>
<td>0.3342</td>
<td>0.0179</td>
<td>0.0247</td>
<td>1.0448</td>
<td>0.0085</td>
</tr>
<tr>
<td>0.8</td>
<td>0.3148</td>
<td>0.0216</td>
<td>0.0248</td>
<td>1.0393</td>
<td>0.0102</td>
</tr>
<tr>
<td>0.9</td>
<td>0.2592</td>
<td>0.0263</td>
<td>0.0355</td>
<td>1.0343</td>
<td>0.0104</td>
</tr>
</tbody>
</table>

These effects follow the effects of the regeneration. As the regeneration is less effective, the $SO_2$ concentration in the stack gases decreases. As a concentration that is as high as possible is wanted, these trends also point to using a charge that is as small as possible.

Model 3:

Using the third model, the same calculations have been made, but now the residence times of the separate beds influence the average activity and via that the operation of the system. The residence times of beds two, three and four are taken constant. Their respective values are 2359 s, 1600 s and 244 s.

The effect on the operation of bed one and consequently on the whole system are shown in table 19.

Table 19. The effect of varying the charge on the operation of bed one.

<table>
<thead>
<tr>
<th>$\alpha$ (-)</th>
<th>$V_i$ (m$^3$)</th>
<th>$\Phi_{\text{sorbent}}$ (kg/s)</th>
<th>$\Phi_{\text{make-up}}$ (kg/s)</th>
<th>$\tau_i$ (s)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.08</td>
<td>301</td>
<td>15.7849</td>
<td>0.0283</td>
<td>11946</td>
</tr>
<tr>
<td>0.1</td>
<td>368</td>
<td>10.9861</td>
<td>0.0306</td>
<td>20887</td>
</tr>
<tr>
<td>0.2</td>
<td>714</td>
<td>4.5200</td>
<td>0.0510</td>
<td>97239</td>
</tr>
<tr>
<td>0.3</td>
<td>1109</td>
<td>2.8370</td>
<td>0.0762</td>
<td>237600</td>
</tr>
<tr>
<td>0.4</td>
<td>1581</td>
<td>2.0471</td>
<td>0.1063</td>
<td>463200</td>
</tr>
<tr>
<td>0.5</td>
<td>2161</td>
<td>1.5844</td>
<td>0.1429</td>
<td>806367</td>
</tr>
<tr>
<td>0.6</td>
<td>2895</td>
<td>1.2734</td>
<td>0.1881</td>
<td>1324228</td>
</tr>
<tr>
<td>0.7</td>
<td>3891</td>
<td>1.0478</td>
<td>0.2483</td>
<td>2128881</td>
</tr>
<tr>
<td>0.8</td>
<td>5374</td>
<td>0.8666</td>
<td>0.3364</td>
<td>3489476</td>
</tr>
<tr>
<td>0.9</td>
<td>8110</td>
<td>0.6976</td>
<td>0.4945</td>
<td>6375895</td>
</tr>
</tbody>
</table>
The trends found are the same as found with model two and they have the same explanations. The residence time in bed one increases notably with increasing charge. Increasing the residence time increases the cycle time which in turn increases the average activity of the sorbent.

Contrary to the calculations with model two, the calculation with model three incorporated the calculation of the stack gas concentrations of NO and NH$_3$ in bed one. The trends for these are shown in table 20.

Table 20. Calculations on the formation of nitrogen products with varying charge.

<table>
<thead>
<tr>
<th>$\alpha$ (-)</th>
<th>$O_2$ (mole/m$^3$)</th>
<th>NO (mole/m$^3$)</th>
<th>NH$_3$ (mole/m$^3$)</th>
<th>N$_{\text{part}}$ (#/m$^3$)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.08</td>
<td>0.42771</td>
<td>8.9e-4</td>
<td>3.1e-6</td>
<td>106728356</td>
</tr>
<tr>
<td>0.1</td>
<td>0.42772</td>
<td>5.0e-4</td>
<td>2.1e-6</td>
<td>106425150</td>
</tr>
<tr>
<td>0.2</td>
<td>0.42786</td>
<td>1.0e-4</td>
<td>5.4e-7</td>
<td>105025768</td>
</tr>
<tr>
<td>0.3</td>
<td>0.42800</td>
<td>4.1e-5</td>
<td>2.4e-7</td>
<td>103658199</td>
</tr>
<tr>
<td>0.4</td>
<td>0.42818</td>
<td>2.1e-5</td>
<td>1.3e-7</td>
<td>102265689</td>
</tr>
<tr>
<td>0.5</td>
<td>0.42838</td>
<td>1.2e-5</td>
<td>7.4e-8</td>
<td>100831949</td>
</tr>
<tr>
<td>0.6</td>
<td>0.42860</td>
<td>7.2e-6</td>
<td>4.5e-8</td>
<td>99343399</td>
</tr>
<tr>
<td>0.7</td>
<td>0.42885</td>
<td>4.5e-6</td>
<td>2.8e-8</td>
<td>97749605</td>
</tr>
<tr>
<td>0.8</td>
<td>0.42919</td>
<td>2.7e-6</td>
<td>1.8e-8</td>
<td>95954564</td>
</tr>
<tr>
<td>0.9</td>
<td>0.42979</td>
<td>1.5e-6</td>
<td>9.6e-9</td>
<td>93529991</td>
</tr>
</tbody>
</table>

The concentrations are given in mole per m$^3$ at a temperature of 850 C. To obtain the concentration at 20 C in mg/m$^3$, these values have to be multiplied with a factor 130e3. Doing this for the highest value of 8.9e-4 mole/m$^3$, a concentration at room temperature of 116 mg/m$^3$ is obtained.

Looking at the emission specifications as given in table 1, learns that the maximum allowed concentration is 100 mg/m$^3$. But, this is the concentration at stack gas temperature which equals about 120 C. At this temperature the density of the gas is lower and so also the concentration.

Furthermore, the model for the formation of nitrogen oxide turned out to be on closer look a very pessimistic one. It states that the non-volatile part of the nitrogen in the coal is released as nitrogen oxide, but it is believed that it is released as nitrogen also.

The effect of the charge on the regeneration in bed three is to be found in table 21.
Table 21. The effect of the variation of the charge on the regeneration in bed three.

<table>
<thead>
<tr>
<th>( \alpha ) (-)</th>
<th>average activity</th>
<th>( V_3 ) (m(^3))</th>
<th>( \alpha_1 ) (-)</th>
<th>( \alpha_2 ) (-)</th>
<th>( \alpha_3 ) (-)</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.08</td>
<td>0.7046</td>
<td>31.1033</td>
<td>0.0092</td>
<td>0.9716</td>
<td>0.0191</td>
</tr>
<tr>
<td>0.1</td>
<td>0.8099</td>
<td>21.6475</td>
<td>0.0117</td>
<td>0.9646</td>
<td>0.0237</td>
</tr>
<tr>
<td>0.2</td>
<td>0.9530</td>
<td>8.9064</td>
<td>0.0207</td>
<td>0.9326</td>
<td>0.0467</td>
</tr>
<tr>
<td>0.3</td>
<td>0.9803</td>
<td>5.5902</td>
<td>0.0275</td>
<td>0.9035</td>
<td>0.0690</td>
</tr>
<tr>
<td>0.4</td>
<td>0.9898</td>
<td>4.0337</td>
<td>0.0334</td>
<td>0.8769</td>
<td>0.0897</td>
</tr>
<tr>
<td>0.5</td>
<td>0.9941</td>
<td>3.1253</td>
<td>0.0385</td>
<td>0.8522</td>
<td>0.1093</td>
</tr>
<tr>
<td>0.6</td>
<td>0.9964</td>
<td>2.5092</td>
<td>0.0428</td>
<td>0.8283</td>
<td>0.1289</td>
</tr>
<tr>
<td>0.7</td>
<td>0.9978</td>
<td>2.0646</td>
<td>0.0462</td>
<td>0.8049</td>
<td>0.1489</td>
</tr>
<tr>
<td>0.8</td>
<td>0.9986</td>
<td>1.7076</td>
<td>0.0482</td>
<td>0.7795</td>
<td>0.1724</td>
</tr>
<tr>
<td>0.9</td>
<td>0.9993</td>
<td>1.3746</td>
<td>0.0468</td>
<td>0.7454</td>
<td>0.2079</td>
</tr>
</tbody>
</table>

As mentioned before with model two, an increase in charge means a decrease in regeneration efficiency. The average activity has but a small effect on the regeneration. As shown in the table the activity decreases with decreasing charge. This is due to the decreasing of the residence time. The only notable effect the average activity has, is on the sorbent flow. The average activity calculated with model three is higher than with model two. This has as a consequence that with model two, a greater sorbent flow has to be used to retain the same amount of sulphur at the same charge.

The effect on the stack gases of bed three is to be found in table 22. The effects that occur are similar to the trends with model two as shown in table 18.
Table 22. The effect of the variation of the charge of the sorbent on the concentrations of various compounds in the stack gas of bed three.

<table>
<thead>
<tr>
<th>α</th>
<th>SO(_2) (mol/m(^3))</th>
<th>H(_2) (mol/m(^3))</th>
<th>H(_2)O (mol/m(^3))</th>
<th>H(_2)S (mol/m(^3))</th>
<th>S(_2) (mol/m(^3))</th>
</tr>
</thead>
<tbody>
<tr>
<td>0.08</td>
<td>0.4212</td>
<td>0.0037</td>
<td>1.0656</td>
<td>0.0018</td>
<td>0.0013</td>
</tr>
<tr>
<td>0.1</td>
<td>0.4191</td>
<td>0.0042</td>
<td>1.0643</td>
<td>0.0026</td>
<td>0.0021</td>
</tr>
<tr>
<td>0.2</td>
<td>0.4035</td>
<td>0.0056</td>
<td>1.0592</td>
<td>0.0062</td>
<td>0.0067</td>
</tr>
<tr>
<td>0.3</td>
<td>0.3859</td>
<td>0.0066</td>
<td>1.0547</td>
<td>0.0097</td>
<td>0.0119</td>
</tr>
<tr>
<td>0.4</td>
<td>0.3679</td>
<td>0.0074</td>
<td>1.0505</td>
<td>0.0131</td>
<td>0.0174</td>
</tr>
<tr>
<td>0.5</td>
<td>0.3485</td>
<td>0.0081</td>
<td>1.0464</td>
<td>0.0168</td>
<td>0.0229</td>
</tr>
<tr>
<td>0.6</td>
<td>0.3255</td>
<td>0.0089</td>
<td>1.0421</td>
<td>0.0201</td>
<td>0.0286</td>
</tr>
<tr>
<td>0.7</td>
<td>0.2987</td>
<td>0.0096</td>
<td>1.0375</td>
<td>0.0239</td>
<td>0.0343</td>
</tr>
<tr>
<td>0.8</td>
<td>0.2626</td>
<td>0.0106</td>
<td>1.0318</td>
<td>0.0286</td>
<td>0.0402</td>
</tr>
<tr>
<td>0.9</td>
<td>0.2036</td>
<td>0.0124</td>
<td>1.0225</td>
<td>0.0361</td>
<td>0.0468</td>
</tr>
</tbody>
</table>

Conclusion

Based on the results gotten from the calculations with either activity model, the conclusion may be drawn that it is best to operate the system with a charge that is as small as possible. The only point against it, is that with decreasing charge, the nitrogen oxide concentration in the stack gas of bed one increases and might exceed the maximum allowed level. However, the calculation model is a pessimistic one, so it is reasonable to expect that that will not happen.

It has to be noted, that the conclusion as drawn before, is only based on the calculations as shown in the tables in this chapter. To be sure that the conclusion is correct, an economic optimization should be made, regarding the bed size, make-up stream and compression costs. Due to lack of time this has not been done.

In the following pages the simplified models for the calculation with MERCURY\textsuperscript{TM} are included. These concern the models for bed one, desulphurization and formation of nitrogen oxides, and bed three, regeneration.
Problem: C:\MERCURY\1.EKA

\[ V_r := 400 \]
\[ \phi := 0 \]
\[ d_p := 2e-3 \]
\[ \rho_s := 1400 \]
\[ K_s := 0.149 \]
\[ K_{eq} := 0.164 \]
\[ \phi_{gas} := 361.158 \]
\[ \tau_2 := 2359 \]
\[ \tau_3 := 1600 \]
\[ \tau_4 := 244 \]
\[ \phi_{vers} := 0.6 \times \phi \times (\tau_1 + \tau_2 + \tau_3 + \tau_4) / (3600 \times 1500) \]
\[ \text{fractie} := 0.7048 \]
\[ \alpha := 0.08 \]
\[ d_{kool} := 2e-3 \]
\[ \phi_{kool} := 8.8 \]
\[ \alpha_1 := 0.0093 \]
\[ \alpha_2 := 0.9717 \]
\[ \alpha_3 := 0.0190 \]
\[ C_{SO_2} := 0.23 \times 600e-3 / 64 \]
\[ \epsilon := 0.55 \]

\[ \text{SURE2-MODEL} \]
\[ D_{SO_2\text{shell}} := 7.53e-7 \]
\[ D_{SO_2\text{film}} := 1.34108e-9 \]
\[ D_{SO_3\text{shell}} := 0.9 \times D_{SO_2\text{shell}} \]
\[ D_{SO_3\text{film}} := 0.9 \times D_{SO_2\text{film}} \]
\[ C_{\text{sin}} := (0.0075 / 32.06e-3) \times \phi_{kool} + (\alpha_1 + \alpha_3) \times 0.0891 \times \phi_{gas} \]
\[ \alpha := \left( C_{\text{sin}} - C_{SO_2} \right) \times (\phi_{gas} \times 40.08e-3) / (\text{fractie} \times 0.0891 \times \phi) \]
\[ Y := (1 - \alpha)^{1/3} \]
\[ C_{O_2} := 0.2095 \times 1000 \times \text{rhofluidum} / (22.4 \times 1.29) + \phi_{kool} \times 1.6875 / (\phi_{gas}) + (1.5 \times \alpha_1 - 0.5 \times \alpha_3) \]
\[ \times \text{fractie} \times 0.0891 \times (\phi_{gas} + 40.08e-3) \]
\[ C_{O_2} := C_{O_2}^{\text{in}} - (75.91 \times \phi_{kool} + (1.5 \times \alpha) \times \phi_{gas} \times 40.08e-3) / (\phi_{gas}) \]
\[ \alpha_{film} := (2 / dp - 1 / (0.5 \times dp + \delta)) / (D_{SO_2\text{film}} + D_{SO_3\text{film}} \times K_{eq} \times \sqrt{C_{O_2}}) \]
\[ \alpha_{shell} := (1 / (Y \times 0.5 \times dp) - 2 / dp) / (D_{SO_2\text{shell}} + D_{SO_3\text{shell}} \times K_{eq} \times \sqrt{C_{O_2}}) \]
\[ \eta_{\text{fluidum}} := 464e-7 \]
\[ \text{rhofluidum} := 0.30953 \]
\[ \delta := 2.1e-9 \]
\[ N_{\text{part}} := \phi / (1400 \times (\phi / 1400 + 0.2 \times 0.103 \times \phi_{kool} / 2448)) \]
\[ \times (1 - \epsilon) \times 3 / (4 \times 3.1416 \times (0.5 \times dp) ^ 3) \]
\[ X_{\text{feed}} := 1 \]
\[ X := C_{SO_2} / C_{\text{sin}} \]
\[ \phi_{gas} \times (X_{\text{feed}} - X) \times (1 + K_{eq} \times \sqrt{C_{O_2}}) = (K_{eq} \times \sqrt{C_{O_2}}) \]
\[ \times 4 \times 3.1416 \times (Y \times 0.5 \times dp) ^ 2 \times \text{N_{part} \times V_r} / (1 / K_s + (Y \times 0.5 \times dp) ^ 2 \]
\[ \times (\alpha_{film} + \alpha_{shell}) = 0 \]
\[ \tau_1 := V_r / (1 - \epsilon) / (\phi / 1400 + 0.2 \times 0.103 \times \phi_{kool} / 2448) \]
tijd bed1,s

Vr>0

--- Solution ---

Variables:

\[ Vr = +301.28206617327 \]
\[ \text{phis} = +15.78468649848424 \]
\[ dp = +0.0020000000000000 \] \{ = +1 / 500 \}
\[ \text{rhos} = +1400.0000000000000 \]
\[ Ks = +0.1490000000000000 \]
\[ Keq = +0.1640000000000000 \]
\[ \text{phigas} = +361.1580000000000 \]
\[ \tau2 = +2359.0000000000000 \]
\[ \tau3 = +1600.0000000000000 \]
\[ \tau4 = +244.00000000000000 \]
\[ \phiVers = 0.6*\text{phis}*(2359+0.45*Vr/(\text{phis}/1400+7.4052287581699E-05)+1600+244)/5400000 \]
\[ = +0.028323666064652423 \]
\[ \tau1 = 0.45*Vr/(\text{phis}/1400+7.4052287581699E-05) \]
\[ = +1194.6201807082125 \]
\[ \text{fractie} = +0.704800000000000 \] \{ = +881 / 1250 \}
\[ \text{alfa} = +0.080000000000000 \] \{ = +2 / 25 \}
\[ \text{dkool} = +8.80000000000000 \] \{ = +44 / 5 \}
\[ \text{alfal} = +0.009300000000000 \] \{ = +93 / 1000 \}
\[ \text{alfa2} = +0.971700000000000 \] \{ = +9717 / 1000 \}
\[ \text{alfa3} = +0.019000000000000 \] \{ = +19 / 1000 \}
\[ \text{Cso2} = +0.002156250000000 \] \{ = +69 / 3200 \}
\[ \varepsilon1 = +0.550000000000000 \] \{ = +11 / 20 \}
\[ \text{Dso2shell} = +7.53000000000000E-07 \]
\[ \text{Dso2film} = +1.341080000000000 \] \{ = +134108 / 100000 \}
\[ \text{Dso3shell} = +6.77700000000000E-07 \]
\[ \text{Dso3film} = +1.206972000000000 \] \{ = +1206972 / 1000000 \}
\[ \text{Csin} = (2.0586400499064+0.001777174344*\text{phiVers}*(2359+0.45*Vr/(\text{phis}/1400+7.4052287581699E-05)+1600+244)/5400000)/0.04008)/361.158 \]
\[ = +0.0076345971296324054 \]
\[ \text{Y} = +0.972588826218856 \]
\[ \text{Co2in} = 0.0411177379429502+0.000279449676* \]
\[ *(\text{phiVers}*(2359+0.45*Vr/(\text{phis}/1400+7.4052287581699E-05)+1600+244)/5400000)/14.47521264 \]
\[ +2.2441353474529 \]
\[ = +2.2855572716554517 \]
\[ \text{rhofluidum} = +0.309530000000000 \] \{ = +30953 / 100000 \}
\[ \text{Co2} = -(668.008+0.0075357216*\text{phiVers}*(2359+0.45*Vr/(\text{phis}/1400+7.4052287581699E-05)+1600+244)/5400000)/14.47521264+2.2441353474529 \]
\[ = +0.42771161036879707 \]
\[ \text{alfafilm} = 0.0020999959/(1.34108E-09+1.97943408E-10) \]
\[ *\text{SQRT}(-668.008+0.0075357216*\text{phiVers}*(2359+0.45*Vr/(\text{phis}/1400+7.4052287581699E-05)+1600+244)/5400000)/14.47521264+2.2441353474529) \]
\[ = +1428049.3524460010 \]
\[ \delta = +2.10000000000000E-09 \]
\[ \alpha_{\text{shell}} = \frac{28.1837270193}{(7.53E-07+1.11428E-07 \times \sqrt{-668.008+0.0075357216 \times \phi / 0.04008}/361.158} \\
+0.041117739429502 + 0.000279449676 \times (\phi - 0.6 \times \phi \times (2359+0.45Vr/(\phi/1400+7.4052287581699E-05)+1600+244)/5400000)}{14.47521264+2.2441353474529} \]

\[ \eta_{\text{fluidum}} = +4.6400000000000E-05 \]

\[ N_{\text{part}} = \frac{\phi}{(1400 \times (\phi/1400+7.4052287581699E-05) \times 3 \times 0.45/1.25664E-08} \]

\[ X_{\text{feed}} = +106728355.70453128 \]

\[ X = 0.00215625/((2.0586400499064+0.001777174344 \times (\phi - 0.6 \times \phi \times (2359+0.45Vr/(\phi/1400+7.4052287581699E-05)+1600+244)/5400000)} / 0.04008)/361.158) \]

\[ \text{Largest residual is 2.1E-14} \]

\text{Method: Iterative}

\text{Numeric processor: 80387}
Problem: C:\MERCURY\3.EKA

phis=15.7849
taul=11946
fractie:=0.9
tau3=1600
Vbed:=8
alfa=0.08
SO2:=0.01
A:=-1
alfa1:=0.025
alfa2:=0.85
alfa3:=0.025
H2:=0.004
H20:=1
H2S:=0.05
u3=1.5
dp=2e-3
rhos=1400
H2in=1.07105
Mca=40.08e-3
k1=0.95
k4=0.005
umf=0.48
alfa1>0
alfa3>0
fractie=1-(0.103/(103*tcycl))*tautot
tautot=1325*3600
tcycl=tau1+tau2+tau3+tau4
tau2=2359
tau4=244

{MODEL TER BEREKENING VAN DE VERBLIJFTIJD}
eps3=0.42
tau3*phis/(rhos*(1-eps3))=Vbed
q=fractie*0.0891*rhos*dp/(Mca*6)
phigas=3

{MODEL DER BEREKENING VAN DE REGENERATIE}
alfa1*(q+k1*H2*tau3)-alfa*q=0
alfa2*(1+k4*H2S*tau3/q)-
A*alfa*q*(1+k4*H2S*tau3/q)/(q+k1*H2*tau3)-(1-alfa-(alfa*A))=0
A=k1*H2/(k4*H2S-k1*H2)
alfa3=1-alfa1-alfa2
H2in=(H2+H20+H2S)
phis*0.0891*fractie*(alfa-alfa1-alfa3)/Mca=
=phigas*(H2S+SO2+2*S2)
phis*0.0891*fractie*
(3*alfa1-4*alfa1-alfa2)/Mca=phigas*(2*SO2+H2O)

(5.51874e8*SO2^2*H2^4)-(S2*10.7105*H2O^4)=0
(1929.2299*H2^2*S2)-(H2S^2*10.7105)=0
Solution

Variables:

\[ \text{phis} = +15.784900000000 \quad \{ = +157849 / 10000 \} \]

\[ \text{tau1} = +1194.600000000000 \]

\[ \text{fractie} = +0.704625673416311 \quad \{ = +3793 / 5383 \} \]

\[ \text{tau3} = +1600.000000000000 \]

\[ \text{Vbed} = +31.1032512315271 \quad \{ = +157849 / 5075 \} \]

\[ \text{alfa} = +0.080000000000000 \quad \{ = +2 / 25 \} \]

\[ \text{SO2} = 0.659354579291232-8.2419321365165 \]

\[ \text{*alfal-8.2419321365165*alfa3-H2S-2*S2} \]

\[ \text{A} = +0.42118303129007056 \quad \{ = 0.95*H2/(0.005*H2S-0.95*H2) \}
\]

\[ \text{alfa1} = +0.00921732718902584 \]

\[ \text{alfa2} = (0.92-0.08*0.95*(0.058479711578144/lalfa1-0.7309963947268)/(0.005*H2S-0.95*(0.058479711578144/lalfa1-0.7309963947268)/1520+0.95*(670.058479711578144/lalfa1-0.7309963947268)/1520/(0.005*H2S-0.95*(0.058479711578144/lalfa1-0.7309963947268)/1520)/(1+8*H2S/0.7309963947268)\]/1520 \]

\[ \text{alfa3} = +0.00921732718902584 \]

\[ \text{H2} = (0.058479711578144/lalfa1-0.7309963947268)/1520 \]

\[ \text{H2S} = +0.001800000000000 \]

\[ \text{H2in} = +1.500000000000000 \quad \{ = +3 / 2 \} \]

\[ \text{dp} = +0.002000000000000 \quad \{ = +1 / 500 \} \]

\[ \text{rhos} = +1400.000000000000 \]

\[ \text{H2in} = +1.071050000000000 \quad \{ = +21421 / 20000 \} \]

\[ \text{Mca} = +0.040080000000000 \quad \{ = +501 / 12500 \} \]

\[ \text{k1} = +0.950000000000000 \quad \{ = +19 / 20 \} \]

\[ \text{k4} = +0.005000000000000 \quad \{ = +1 / 200 \} \]

\[ \text{umf} = +0.480000000000000 \quad \{ = +12 / 25 \} \]

\[ \text{g} = +9.807000000000000 \quad \{ = +9807 / 1000 \} \]

\[ \text{tcycl} = +16149.000000000000 \]

\[ \text{tautot} = +477000.000000000 \]

\[ \text{tau2} = +2359.000000000000 \]

\[ \text{tau4} = +244.000000000000 \]

\[ \text{eps3} = +0.420000000000000 \quad \{ = +21 / 50 \} \]

\[ \text{q} = +0.7309963947268 \]

\[ \text{phigas} = +3.000000000000000 \]

\[ \text{S2} = H2S^2*10.7105/((0.058479711578144/1520^2/1929.2299/1520)^2/1929.2299 \]

\[ = +0.0013189853359123448 \]

Largest residual is 4.6E-15
Method: Iterative
Numeric processor: 80387
Problem: C:\MERCURY\NOX.EKA

phis=15.7849
V1=301
tau=11946
N=106728356
Co2=.42771
alfa=0.08
phigas=361.158
phikool=8.8
g=9.807
umf=0.67
R=0.5*dp
rhos=1400
Ks=0.149
Keq=0.164
dkoool=2e-3
eps1=0.55

Problem:

Variables:

phis = +15.784900000000000000
V1 = +301.000000000000000000
tau = +11946.000000000000000000
N = +106728356.000000000000000000
Co2 = +0.427710000000000000
alfa = +0.080000000000000000
phigas = +361.158000000000000000
phikool = +8.800000000000000000
R = +0.500000000000000000
rhos = +1400.000000000000000000
Ks = +0.149000000000000000
Keq = +0.164000000000000000
dkoool = +0.002000000000000000
eps1 = +0.550000000000000000

Solution

Variables:

phis = +15.784900000000000000  { = +157849 / 10000 }
V1 = +301.000000000000000000
tau = +11946.000000000000000000
N = +106728356.000000000000000000
Co2 = +0.427710000000000000
alfa = +0.080000000000000000
phigas = +361.158000000000000000
phikool = +8.800000000000000000
R = +0.500000000000000000
rhos = +1400.000000000000000000
Ks = +0.149000000000000000
Keq = +0.164000000000000000
dkoool = +0.002000000000000000
eps1 = +0.550000000000000000
ka = +0.0313740000000000
kb = +2.0000000000000000
kc = +5.2000000000000000 \{ = +26 / 5 \}
Y = +0.972588826218856
Npart = +32125235156.000000
Scao = +381858.02526288
Sc = +225267.42857143
Cnh3in = +0.007194951477509
Cnoin = +0.0154306820995639
Cnh3 = +3.1382051154313E-06
Cno = +0.00088817792206119

Largest residual is 0
Method: Exact
Numeric processor: 80387
Appendix I

MERCURY MODELS FOR ICFBC
Problem: B:\BED1.EKA

\[ u_1 = 4 \quad \text{(superficiële snelheid bed1, m/s)} \]
\[ g = 9.807 \quad \text{(valversnelling, m/s}^2) \]
\[ H = 4 \quad \text{(hoogte fluide bedden, m)} \]
\[ \text{umf} = 0.67 \quad \text{(minimale fluid snelheid sorb.+as, m/s)} \]
\[ \rho_s = 1400 \quad \text{(dichtheid sorbent, kg/m}^3) \]
\[ K = 0.149 \quad \text{(constante,-)} \]
\[ \text{Keq} = 0.164 \quad \text{(evenwichtsconst. SO2-SO3,-)} \]
\[ A = 90 \quad \text{(reactoroppervlak, m}^2 \)]
\[ V_r = H \cdot A \quad \text{(reactorvolume, m}^3 \)]
\[ \text{tau1} = V_r \cdot (1 - \epsilon_s) / (\phi_s / 1400 + 0.2 \cdot 0.103 \cdot \phi_kool / 2448) \quad \text{(verblijftijd bed1, s)} \]
\[ \text{phigas} = u_1 \cdot A \quad \text{(gasstroom, m}^3/\text{s)} \]
\[ \text{tau2} = 2359 \quad \text{(*) (verblijftijd bed2, s)} \]
\[ \text{tau3} = 928 \quad \text{(*) (verblijftijd bed3, s)} \]
\[ \text{tau4} = 244 \quad \text{(*) (verblijftijd bed4, s)} \]
\[ \phi_s \text{vers} = 0.6 \cdot \phi_s \cdot (\text{tau1} + \text{tau2} + \text{tau3} + \text{tau4}) / (3600 \cdot 1500) \quad \text{(toevoer verse sorbent, kg/s)} \]
\[ \text{fractie} = 0.665316 \quad \text{(stroom sorbent, kg/s)} \]
\[ d_{kool} = 2 \cdot 10^{-3} \quad \text{(deeltjesdiameter kool, m)} \]
\[ \phi_kool = 10.4 \quad \text{(stroom kool, kg/s)} \]
\[ \alpha_1 = 0.05628 \quad \text{(opp CaSO4/tot.opp.Ca)} \]
\[ \alpha_2 = 0.92613 \quad \text{(opp CaO/tot.opp.Ca)} \]
\[ \alpha_3 = 0.01759 \quad \text{(opp CaS/tot.opp.Ca)} \]
\[ C_{SO2} = 0.23 \cdot 700e-3 \cdot 64 \quad \text{(maximale conc. SO2 in afgas, mg/m}^3) \]

\{MODEL TER BEREKENING VAN POROSITEIT\}

\[ \text{deq1} = ((u_1 - \text{umf})^0.5) \cdot ((0.4 \cdot H)^0.75) \cdot g^{-0.25} \]
\[ u_a(l) = (u_1 - \text{umf}) + 0.71 \cdot (g \cdot \text{deq1})^0.5 \]
\[ \epsilon_s(l) = (u_1 - \text{umf}) / u_a(l) \]

\{SURE2-MODEL\}

\[ D_{SO2}\text{shell} = 7.53e-7 \]
\[ D_{SO2}\text{film} = 1.34108e-9 \]
\[ D_{SO3}\text{shell} = 0.9 \cdot D_{SO2}\text{shell} \]
\[ D_{SO3}\text{film} = 0.9 \cdot D_{SO2}\text{film} \]
\[ C_{\text{Sin}} = (0.0075 / 32.06e-3 \cdot \phi_kool + (\alpha_1 + \alpha_3) \cdot 0.0891 \cdot \text{fractie} \cdot (\phi_s - \text{phigas} / 40.08e-3) / \text{phigas} \]
\[ \alpha = ((C_{\text{Sin}} - C_{SO2}) \cdot (\text{phigas} \cdot 40.08e-3)) / (\text{fractie} \cdot 0.0891 \cdot \phi_s) \]
\[ C_{SO3} = C_{SO2} \cdot \sqrt{C_{\text{Co2}}} \cdot \text{Keq} \]
\[ Y = (1 - \alpha)^{(1/3)} \]
\begin{align*}
C_{02in} & = 0.2095 \times 1000 \times \frac{\rho_{fluidum}}{(22.4 \times 1.29)} + \phi_{kool} \\
& = 1.6875 \times (\frac{\phi_{gas}}{(1.5 \times \alpha_{l} - 0.5 \times \alpha_{3}) \times \text{fractie} \times 0.0891}) \\
& \quad \times (\phi_{is} \times (40.08e-3)) \\
C_{02} & = C_{02in} - (75.91 \times \phi_{kool} + (1.5 \times \alpha_{l}) \times \text{fractie} \times 0.0891) \\
& \quad \times \frac{\phi_{is} / 40.08e-3}{\phi_{gas}} \\
\alpha_{film} & = (2/dp - 1/(0.5 \times dp + \delta)) / (D_{so2film} + D_{so3film} \times K_{eq} \times \sqrt{C_{02}}) \\
\alpha_{shell} & = (1/(Y \times 0.5 \times dp) - 2/dp) / (D_{so2shell} + D_{so3shell} \times K_{eq} \times \sqrt{C_{02}}) \\
\eta_{fluidum} & = 464e-7 \\
\rho_{fluidum} & = 0.30953 \\
Sc & = \eta_{fluidum} / (\rho_{fluidum} \times 1.3e-4) \\
Re & = \rho_{fluidum} \times u_{1} \times dp / \eta_{fluidum} \\
St & = 0.81 / (Re \times 0.5 \times Sc \times 0.66) \quad \{5 < Re < 50\} \\
k_{so2film} & = u_{1} \times St / \epsilon_{1} \\
\delta & = 1 / (k_{so2film} / \nu_{so2film} - 2/dp) \\
N_{part} & = \phi_{is} / (1400 \times (\phi_{is} / 1400 + 0.2 \times \phi_{kool} / 2448) \times \nu_{r} \times (1 - \epsilon_{1})^{3} / (4 \times 3.1416 \times (0.5 \times dp)^{3}) \\
X_{feed} & = 1 \\
X & = C_{so2} / C_{sin} \\
& = \phi_{gas} \times (X_{feed} - X) \times (1 + K_{eq} \times \sqrt{C_{02}}) - (K_{eq} \times \sqrt{C_{02}}) \times 4 \times 3.1416 \times (Y \times 0.5 \times dp)^{2} \times X_{Npart} \times \nu_{r} / (1 / K_{s} + (Y \times 0.5 \times dp)^{2} \times (\alpha_{film} + \alpha_{shell})) = 0
\end{align*}

--- Solution ---

Variables:

\begin{align*}
\text{u}_{1} & = +4.0000000000000000 \\
\text{g} & = +9.8070000000000000 \quad \{ = +9807 / 1000 \} \\
\text{H} & = +4.0000000000000000 \quad \{ = +67 / 100 \} \\
\text{umf} & = +0.6700000000000000 \quad \{ = +1 / 500 \} \\
\text{dp} & = +0.0020000000000000 \quad \{ = +149 / 1000 \} \\
\text{rhos} & = +90.0000000000000000 \quad \{ = +41 / 250 \} \\
\text{Ks} & = +360.0000000000000000 \quad \{ = +152 / 5 \} \\
\text{K}_{eq} & = +10.4000000000000000 \quad \{ = +52 / 5 \} \\
\text{phigas} & = +360.0000000000000000 \\
\text{tau}_{2} & = +2359.0000000000000000 \\
\text{tau}_{3} & = +928.0000000000000000 \\
\text{tau}_{4} & = +244.0000000000000000 \\
\text{phisvers} & = +0.027354427410891 \\
\text{fractie} & = +0.6653160000000000 \\
\text{dkool} & = +0.0020000000000000000 \quad \{ = +1 / 500 \} \\
\text{alpha}_{1} & = +0.0562800000000000000 \quad \{ = +1407 / 25000 \} \\
\text{alpha}_{2} & = +0.9261300000000000 \quad \{ = +1759 / 100000 \} \\
\text{C}_{so2} & = +6.95253901850522e-05 \quad \{ = +7.5300000000000000e-07 \} \\
\text{deq}_{1} & = +1.46699306208502 \\
\text{u}_{1} & = +6.02302550375398 \\
\text{D}_{so2film} & = +1.3410800000000000e-09
\end{align*}
Dso3shell = +6.77700000000000E-07
Dso3film = +1.20697200000000E-09
Csin = +0.0088742859483585
alfa = 0.30857441922808-34.7717462592199*Cso2
= +0.30615690000199347
Y = (1-(0.30857441922808-34.7717462592199*Cso2))^0.333333333333333
Co2in = +0.88529312430454765
rhofluidum = +0.3095300000000000 \{ = +30953 / 100000 \}
Co2 = 0.08878476192208+1.5*Cso2
= +0.88889050007356321
alfafilm = 0.0018365892/(1.34108E-09+1.97943408E-10*SQRT(0.08878476192208
+1.5*Cso2))
= +1311760.0613526655
delta = +1.8365925689455E-09
alfashell = 
(1/(0.001*(1-(0.30857441922808-34.7717462592199*Cso2))^0.333333333333333)-1000)/(7.53E-07+1.111428E-07*SQRT(0.0887847
6192208+1.5*Cso2))
= +164817927.21718791
etafluidum = +4.6400000000000E-05
Sc = +1.15311303241391
Rey = +53.367241379310 \{ = +30953 / 580 \}
St = +0.10092812394521
kso2film = +0.73020139888051
Npart = +37766159878.386
Xfeed = +1.0000000000000000
X = 112.685122591184*Cso2

Largest residual is 2.8E-14
Method: Iterative
Numeric processor: 80387
Problem: B: \NOXBED1.EKA

\[ u_1 = 4 \quad \text{[supercifiele snelheid bed1, m/s]} \]
\[ g = 9.807 \quad \text{[valversnelling, m/s²]} \]
\[ H = 4 \quad \text{[hoogte fluide bedden, m]} \]
\[ \text{umf} = 0.67 \quad \text{[minimale fluide snelheid sorb.+as, m/s]} \]
\[ \text{dp} = 2e-3 \quad \text{[deeltjesdiameter sorbent, m]} \]
\[ R = 0.5 \times \text{dp} \quad \text{[straal deeltjes sorbent, m]} \]
\[ \text{rhos} = 1400 \quad \text{[dichtheid sorbent, kg/m³]} \]
\[ \text{Ks} = 0.149 \quad \text{[constante, -]} \]
\[ \text{Keq} = 0.164 \quad \text{[evenwichtsconst. SO2-SO3, -]} \]
\[ \text{tau} = 31639 \quad \text{[verblijftijd, s]} \]
\[ A = V_r / H \quad \text{[reactoroppervlak, m²]} \]
\[ \text{A} = 90 \]
\[ \text{phigas} = u_1 \times A \quad \text{[gasstroom, m³/s]} \]
\[ \text{phis} = 7 \quad \text{[stroom sorbent, kg/s]} \]
\[ \text{dkool} = 2e-3 \quad \text{[deeltjesdiameter kool, m]} \]
\[ \text{phikool} = 10.4 \quad \text{[stroom kool, kg/s]} \]

\[ \text{deq} = ( ( u_1 - \text{umf} ) ^{0.5} ) \times ( ( 0.4 \times H ) ^{0.75} ) \times g^{-0.25} \]
\[ \text{ua} = ( u_1 - \text{umf} ) + 0.71 \times ( g \times \text{deq} ) ^{0.5} \]
\[ \text{eps1} = ( u_1 - \text{umf} ) / \text{ua} \]

\{ INVOER \}

\[ \text{alfa} = 0.30616 \]
\[ \text{Co2} = 0.088889 \]

\{ NOx-MODEL \}

\[ \text{ka} = 0.031374 \]
\[ \text{kb} = 2 \]
\[ \text{kc} = 5.2 \]
\[ \text{Npart} = 37766159878 \]
\[ \text{Y} = 0.8529 \]
\[ \text{Scao} = \text{Npart} \times 3.1415 \times ( Y \times R ) ^{2} \]
\[ \text{Sc} = 6 \times \text{phikool} \times \text{tau} / ( \text{rhos} \times \text{dkool} ) \]
\[ \text{Cnh3in} = 0.318 \times 0.013 \times \text{phikool} / ( \text{phigas} \times 14e-3) \]
\[ \text{Cnoin} = ( 1 - 0.318 ) \times 0.013 \times \text{phikool} / ( \text{phigas} \times 14e-3) \]
\[ \text{Cnh3} = \text{phigas} \times \text{Cnh3in} / ( \text{phigas} + \text{kb} \times \text{Scao} + \text{kc} \times \text{Sc} \times \text{Co2} ) \]
\[ \text{Cno} = \text{phigas} \times \text{Cnoin} / ( \text{phigas} + \text{ka} \times \text{Sc} ) + \text{kb} \times \text{Scao} \times \text{Co2} \times \text{phigas} \]
\[ \text{Cnh3in} / ( ( \text{phigas} + \text{ka} \times \text{Sc} ) \times ( \text{phigas} + \text{kb} \times \text{Scao} + \text{kc} \times \text{Sc} \times \text{Co2} ) ) \]

Solution

Variables:

\[ u_1 = +4.0000000000000000 \]
\[ g = +9.8070000000000000 \quad \{ = +9807 / 1000 \} \]
\[ H = +4.0000000000000000 \]
\[ \text{umf} = +0.6700000000000000 \quad \{ = +67 / 100 \} \]
\[ \text{dp} = +0.0020000000000000 \quad \{ = +1 / 500 \} \]
\[ R = +0.0010000000000000 \quad \{ = +1 / 1000 \} \]
\[ \text{rhos} = +1400.0000000000000 \]
\[ \text{Ks} = +0.1490000000000000 \quad \{ = +149 / 1000 \} \]
\[ \text{Keq} = +0.1640000000000000 \quad \{ = +41 / 250 \} \]
\[ \text{tau} = +31639.0000000000000 \]
A = +90.000000000000000
Vr = +360.000000000000000
phigas = +360.000000000000000
phis = +7.000000000000000
dkool = +0.002000000000000000 \{ = +1 / 500 \}
phikool = +10.400000000000000 \{ = +52 / 5 \}
deq = +1.46699306208502
ua = +6.02302550375398
eps1 = +0.55287828316923
alfa = +0.3061600000000000 \{ = +3827 / 12500 \}
Co2 = +0.08888900000000000
ka = +0.03137400000000000
kb = +2.0000000000000000
kc = +5.200000000000000 \{ = +26 / 5 \}
Npart = +37766159878.000000
Y = +0.8852900000000000
Scao = +371938.384037260
Sc = +705097.71428571
Cnh3in = +0.0085304761904762
Cnoin = +0.0182949206349206
Cnh3 = +7.8262306152980E-06
Cno = +0.00031597476119777

Largest residual is 0
Method: Exact
Numeric processor: 80387
Problem: B:\BED2.EKA

{floater=sorbent}
{sinkers=as}
{H=as}

g=9.807
u2=0.7
umfsorbent=0.48
umfas=0.67
rhoas=2448
rhosorbent=1400
das=2e-3
dsorbent=2e-3

g=9.807 {valversnelling,m2/s}
u2=0.7 {superficiele gassnelheid bed2,m/s}

umfsorbent=0.48 {min.fluid.snelheid sorbent,m/s}
umfas=0.67 {min.fluid.snelheid as,m/s}
rhoas=2448 {dichtheid as,kg/m3}
rhosorbent=1400 {dichtheid sorbent,kg/m3}
das=2e-3 {deeltjesdiameter as,m}
dsorbent=2e-3 {deeltjesdiameter sorbent,m}

h=4 {bedhoogte,m}
A=5 {oppervlak bed,m2}
dt=sqrt(4*A/3.1415) {diameter tank,m}

phikool=10.4 {stroom kool in bed1,kg/s}
phisorbent=7 {stroom sorbent,kg/s}

{MODEL TER BEREKENING VAN DE SEGREGATIE IN BED2}

xgem=0.2*0.103*phikool/(0.103*0.2*phikool+phisorbent)
UTO1=(umfas/umfsorbent)^1.2
UTO2=0.9*(rhoas/rhosorbent)^1.1
UTO3=2.2*sqrt(xgem)*(1-exp(-h/dt))^1.4
UTO/umfsorbent=UTO1+UTO2-UTO3
z=(u2-UTO)*exp(u2/UTO)/(u2-umfsorbent)
M=1/(1+exp(-z))
x=xgem*M {asfractie in de toplaag}

{BEREKENING VAN DE VERBLIJFTIJD IN BED2}

Vr=A*h
eps2=0.4
tau2=Vr(1-eps2)/((phisorbent/rhosorbent)+
(0.103*0.2*phikool/rhoas))
{verblijftijd bed2,s}

Solution

Variables:

g = +9.807000000000000 { = +9807 / 1000 }

u2 = +0.700000000000000 { = +7 / 10 }

umfsorbent = +0.4800000000000000 { = +12 / 25 }

umfas = +0.6700000000000000 { = +67 / 100 }

rhoas = +2448.0000000000000

rhosorbent = +1400.0000000000000

das = +0.0020000000000000 { = +1 / 500 }

dsorbent = +0.0020000000000000 { = +1 / 500 }

h = +4.0000000000000000

A = +5.0000000000000000

dt = +2.523169729653777

phikool = +10.4000000000000000 { = +52 / 5 }

phisorbent = +7.0000000000000000

xgem = +0.0296968218412473 { = +1339 / 45089 }

UTO1 = +1.49210806708351

UTO2 = +1.66415672751785

UTO3 = +0.27502980850979

UTO = +1.3829927933240
\[ z = \frac{(0.7 - U_T) \times \exp(0.7/U_T)}{0.22} \]
\[ = -5.15045030597 \]

\[ M = \frac{1}{1 + \exp(-((0.7 - U_T) \times \exp(0.7/U_T))/0.22))} \]
\[ = +0.005765707416903 \]

\[ x = 0.0296968218412473 \times \frac{1}{1 + \exp(-((0.7 - U_T) \times \exp(0.7/U_T))/0.22))} \]
\[ = +0.000172231859485 \]

\[ V_r = +0.4000000000000000 \quad \{ = \frac{2}{5} \} \]

\[ \tau_2 = +2358.71478307789 \]

Largest residual is 0
Method: Exact
Numeric processor: 80387
Problem: B:\BED23_2.EKA

--- Solution ---

Variables:

<table>
<thead>
<tr>
<th>Variable</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>phis23</td>
<td>+7.00000000000000000000</td>
</tr>
<tr>
<td>Abed2</td>
<td>+5.00000000000000000000</td>
</tr>
<tr>
<td>u2</td>
<td>+0.70000000000000000000</td>
</tr>
<tr>
<td>umf2</td>
<td>+0.67000000000000000000</td>
</tr>
<tr>
<td>Aorif</td>
<td>+0.00353420000000000000</td>
</tr>
<tr>
<td>horif</td>
<td>+0.30000000000000000000</td>
</tr>
<tr>
<td>dp</td>
<td>+0.00200000000000000000</td>
</tr>
<tr>
<td>Lbed</td>
<td>+0.25000000000000000000</td>
</tr>
<tr>
<td>phib</td>
<td>+0.95000000000000000000</td>
</tr>
<tr>
<td>H</td>
<td>+4.00000000000000000000</td>
</tr>
<tr>
<td>g</td>
<td>+9.80700000000000000000</td>
</tr>
<tr>
<td>rhos</td>
<td>+1400.0000000000000000</td>
</tr>
<tr>
<td>rhogas</td>
<td>+0.30952560000000000000</td>
</tr>
</tbody>
</table>
mu = +4.640000000000000E-05
eps3 = +0.4200000000000000 { = +21 / 50 }
Ut = +173.584370895532
Ly = +8211938.6571230
epsmf2 = +0.39135812349005
drukgrad2 = +8356.5629676106
drukgrad3 = +7963.28400000000
deltaP0 = +1455.132180159
deltauhor = +1.1273426353953
Uabs = +2.32442427748737
ughor = 2.32442427748737+deltauhor
= +3.45176912883
phigo = 0.00138312971200385*ughor
= +0.004774241376120
Ug = 1.78865364436124-0.00070684*ughor
= +1.7862137974365
Us = 0.000234713722328454*phis
= +0.00243984692470
phis = 7+3.0114984031946*deltauhor
= +10.394990546
phigas = +3.5000000000000000 { = +7 / 2 }
relgasoverspraak = 0.039517991771539*ughor
= +0.13640689646057

Largest residual is 2.3E-10
Method: Exact
Numeric processor: 80387
Problem: B:\BED3.EKA

SO2:=0.01
A:=-1
alfa1:=0.1
alfa2:=0.8
alfa3:=0.005
H2:=0.004
H2O:=1
H2S:=0.05
u3=1.5
H=4
Abed=2.0
dp=2e-3
rhos=1400
phis=7
H2in=1.07105
Mca=40.08e-3
k1=0.95
k4=0.005
alfa=.30615
umf=0.48
q=fractie*0.0891*rhos*dp/(Mca*6)
phigas=u3*Abed

{MODEL TER BEREKENING VAN DE VERBLIJFTIJD}

esp3=0.42
tau3=Abed*H*rhos*(1-eps3)/phis
q=fractie*0.0891*rhos*dp/(Mca*6)
phigas=u3*Abed

{MODEL DER BEREKENING VAN DE REGENERATIE}

alfa1*(q+k1*H2*tau3)-alfa*q=0
alfa2*(1+k4*H2S*tau3/q)-A*alfa*q*(1+k4*H2S*tau3/q)/(q+k1*H2*
tau3)-(1-alfa-(alfa*A»=0
A=k1*H2/(k4*H2S-k1*H2)
alfa3=1-alfa1-alfa2

H2in=(H2+H2O+H2S)
phis*0.0891*fractie*(alfa-alfa1-alfa3)/Mca=u3*Abed*(H2S+SO2+2*S2)
phis*0.0891*fractie*(3*alfa1+1-4*alfa1-alfa2)/Mca=u3*Abed*
(2*SO2+H2O)

(5.51874e8*SO2^2*H2^4)-(S2*10.7105*H2O^4)=0
(1929.2299*H2^2*S2)-(H2S^2*10.7105)=0

n=1325*3600/(tau1+tau2+tau3+tau4) {aantal cycli,-}
fractie=(0.997)^n {activiteit sorbent,-}
tau1=31639
tau2=2359
tau4=244

Solution ====

Variables:
\[ S = 0.056548251341 - 3.451080357147 \times \text{alpha1} - 3.451080357147 \times \text{alpha3} - \text{H2S} - 2 \times \text{S2} \]

\[ \text{alpha1} = 0.0562776179336062 \]

\[ \text{alpha2} = \left( 0.69385 - 0.30615 \times 0.95 \left( \frac{0.21130965026811}{\text{alpha1}} - 0.6902160714294 \right) \right) / 881.6 \]

\[ 0.21130965026811 / (0.005 \times \text{H2S} - 0.95 \times (0.21130965026811 / \text{alpha1} - 0.6902160714294) / 881.6) + 0.95 \times (0.21130965026811 / \text{alpha1} - 0.6902160714294) / 881.6 \]

\[ \text{alfa3} = +0.92613354227827194 + 0.01758883978812192 \]

\[ \text{H2} = \left( \frac{0.21130965026811}{\text{alpha1}} - 0.6902160714294 \right) / 881.6 \]

\[ \text{H2S} / 0.6902160714294 \]

\[ \text{H2in} = +1.07105 - (0.21130965026811 / \text{alpha1} - 0.6902160714294) / 881.6 - \text{H2S} \]

\[ \text{H2in} = +1.0647487903360444 \]

\[ \text{H2S} = +0.00282507945234963 \]

\[ \text{u3} = +1.5000000000000000 \quad \{ = 3 / 2 \} \]

\[ \text{H} = +4.0000000000000000 \]

\[ \text{Abed} = +2.0000000000000000 \]

\[ \text{dp} = +0.0020000000000000 \quad \{ = 1 / 500 \} \]

\[ \text{rhos} = +1400.0000000000000 \]

\[ \text{phis} = +7.0000000000000000 \]

\[ \text{H2in} = +1.0710500000000000 \quad \{ = 21421 / 20000 \} \]

\[ \text{Mca} = +0.0400800000000000 \quad \{ = 501 / 12500 \} \]

\[ \text{k1} = +0.9500000000000000 \quad \{ = 19 / 20 \} \]

\[ \text{k4} = +0.0050000000000000 \quad \{ = 1 / 200 \} \]

\[ \text{alpha} = +0.3061500000000000 \quad \{ = 6123 / 20000 \} \]

\[ \text{umf} = +0.4800000000000000 \quad \{ = 12 / 25 \} \]

\[ \text{g} = +9.8070000000000000 \quad \{ = 9807 / 1000 \} \]

\[ \text{eps3} = +0.4200000000000000 \quad \{ = 21 / 50 \} \]

\[ \text{tau3} = +928.0000000000000 \]

\[ \text{q} = +0.6902160714294 \]

\[ \text{fractie} = +0.6653165017530 \]

\[ \text{phigas} = +3.0000000000000000 \]

\[ \text{H2S}^2 \times 10.7105 / ((0.21130965026811 / \text{alpha1} - 0.6902160714294) / 881.6)^2 / 1929.2299 = +0.0036668658487751136 \]

\[ \text{n} = +135.626954791015 \]

\[ \text{tau1} = +31639.0000000000000 \]

\[ \text{tau2} = +2359.0000000000000 \]

\[ \text{tau4} = +244.0000000000000 \]

Largest residual is 5.7E-17

Method: Iterative

Numeric processor: 80387
### Problem: B:\BED4-2.EKA

\[ \text{phis}:=7 \quad \text{ {sorbentstroom, kg/s}} \]
\[ \text{phis41}=7 \quad \text{ {oppervlakte bed, m²}} \]
\[ \text{Abed}=0.5 \quad \text{ {minimale fluidisatiesnelheid, m/s}} \]
\[ \text{u4}=0.4863514 \quad \text{ {minim. fluid. snelh. bed1, m/s}} \]
\[ \text{umf}=0.48 \quad \text{ {totaal oppervlak orifices, m²}} \]
\[ \text{umfl}=0.67 \quad \text{ {totaal oppervlak orifices, m²}} \]
\[ \text{Aorif}=3.5342e-3 \quad \text{ {totaal oppervlak orifices, m²}} \]
\[ \text{L0}=0.1496 \quad \text{ {bedhoogte bij proef voor umf, m}} \]
\[ \text{ul}=4 \quad \text{ {superficiele snelheid bed1, m/s}} \]
\[ \text{horif}=0.1 \quad \text{ {hoogte orifices, m}} \]
\[ \text{dp}=2e-3 \quad \text{ {deeltjesdiameter sorbent, m}} \]
\[ \text{Lbed}=\sqrt{Abed} \quad \text{ {afstand waarover drukval deltaPO, m}} \]
\[ \text{phib}=0.95 \quad \text{ {sfericiteitsfactor deeltjes, -}} \]
\[ \text{H}=4 \quad \text{ {bedhoogte bij proef voor umf, m}} \]
\[ \text{g}=9.807 \quad \text{ {valversnelling, m²/s}} \]
\[ \text{rhos}=1400 \quad \text{ {dichtheid sorbent, kg/m³}} \]
\[ \text{rhogas}=0.3095256 \quad \text{ {dichtheid gas bij 850 °C, kg/m³}} \]
\[ \text{mu}=464e-7 \quad \text{ {viscositeit bij 850 °C, Ns}} \]

\{MODEL TER BEPALING POROSITEIT BIJ MINIMALE FLUIDISATIESNELHEID\}

\[ \text{Ut}=\frac{(4\ast g\ast dp\ast (\text{rhos}\ast \text{rhogas}))/((3\ast \text{rhogas}\ast 0.431)\ast 0.5)}{\text{ {terminale gassnelheid, m/s}}} \]
\[ \text{Ly}=\left(\frac{\text{Ut}}{\text{Ly}}\right)^3/\left(\frac{(g\ast \text{mu}\ast (\text{rhos}\ast \text{rhogas}))/((\text{Ly})\ast 0.017)}{\text{ {Lyman getal, -}}} \right) \]
\[ \text{epsmf}=0.513\ast \left(\frac{(\text{Ly})^{0.017}}{\text{ {porositeit bij minim.flui,-}}} \right) \]

\{MODEL TER BEPALING POROSITEIT\}

\[ \text{deql}=((\text{ul}\ast \text{umfl})^{0.5}\ast ((0.4\ast \text{H})^{0.75}\ast g\ast 0.25) \quad \text{ {modelparameter}} \]
\[ \text{ual}=\text{(ul}\ast \text{umfl})+0.71\ast (g\ast \text{deql})^{0.5} \quad \text{ {modelparameter}} \]
\[ \text{eps1}=((\text{ul}\ast \text{umfl})/\text{ual} \quad \text{ {porositeit in bed1}} \]

\{MODEL TER BEREKENING GASOVERSPRAAK (KORBEE) \}

\[ \text{drukgrad4}=6570.98 \quad \text{ {drukgradient bed4, Pa.m}} \]
\[ \text{x}=32.7\ast \text{u1}/\text{umf} \]
\[ \text{y}=63.065+0.1012\ast x \]
\[ \text{druk1}=9.80665\ast y \]
\[ \text{drukgrad1}=\text{druk1}\ast \text{epsmf}/(\text{L0}\ast \text{eps1}) \quad \text{ {drukgradient bed1, Pa.m}} \]
\[ \text{deltaP0}=\left((\text{H}\ast \text{horif})\ast (\text{drukgrad4}\ast \text{drukgrad1}) \quad \text{ {drukverschil bed4-bed1, Pa}} \right) \]
\[ \text{deltaP0}/\text{Lbed}=150\ast ((1/\text{epsmf})-1)^{2}\ast \text{mu*deltauhor}/(\text{phib*dp})^{2}+1.75\ast ((1/\text{epsmf})-1)\ast \text{rhogas*deltauhor}^2/(\text{phib*dp}) \]
\[ \text{deltauhor}>0 \]
\[ \text{Uabs}=\text{phis41}/(\text{rhos}\ast (1\ast \text{epsmf})\ast \text{Aorif}) \]
\[ \text{deltauhor}=-\text{Uabs+ughor} \]
\[ \text{phigo}=\text{ughor}\ast \text{Aorif}\ast \text{epsmf} \]
\[ \text{gasoverspraak, m³/s} \]
\[ \text{Ug}=((\text{u4}-\text{(phigo/Abed)}) / \text{epsmf} \quad \text{ {gasstroom naar bed4, m³/s}} \]
\[ \text{Us+Ug}/\text{epsmf} \]
\[ \text{Us}=\text{phis}/(\text{rhos}\ast (1\ast \text{epsmf})\ast \text{Abed}) \quad \text{ {sorb. snelh. in bed, m/s}} \]
\[ \text{phigas}=\text{u4}\ast \text{Abed} \quad \text{ {gasstroom naar bed4, m³/s}} \]
\[ \text{relgasoverspraak}\ast \text{phigo*100/phigas} \quad \text{ {percentage gasoverspraak, %}} \]
\[ \text{tau4}=\text{H}\ast \text{Abed}\ast (1\ast \text{epsmf})\ast \text{rhos}/\text{phis} \quad \text{ {verblijftijd bed4, s}} \]
Solution

Variables:

\[
\begin{align*}
\text{phis} & = -6.9144747568 + 3.0114984031946 \times \text{ughor} \\
& = +7.00005495454 \\
\text{phis41} & = +7.0000000000000000 \\
\text{Abed} & = +0.5000000000000000 \\
& = +1 / 2 \\
\text{u4} & = +0.4863514000000000 \\
\text{umf} & = +0.4800000000000000 \\
& = +12 / 25 \\
\text{umfl} & = +0.6700000000000000 \\
& = +67 / 100 \\
\text{Aorif} & = +0.0035342000000000 \\
\text{L0} & = +0.1496000000000000 \\
& = +187 / 1250 \\
\text{ul} & = +4.0000000000000000 \\
\text{horif} & = +0.1000000000000000 \\
& = +1 / 10 \\
\text{dp} & = +0.0020000000000000 \\
& = +1 / 500 \\
\text{Lbed} & = +0.70710678118655 \\
\text{phib} & = +0.9500000000000000 \\
& = +19 / 20 \\
\text{H} & = +4.0000000000000000 \\
\text{g} & = +9.8070000000000000 \\
& = +9807 / 1000 \\
\text{rhos} & = +1400.0000000000000 \\
\text{rhogas} & = +0.3095256000000000 \\
\text{mu} & = +4.6400000000000000 \times 10^{-5} \\
\text{Ut} & = +173.584370895532 \\
\text{Ly} & = +8211938.6571230 \\
\text{epsmf} & = +0.391355812349005 \\
\text{deql} & = +1.46699306208502 \\
\text{ual} & = +6.02302550375398 \\
\text{eps1} & = +0.55287828316923 \\
\text{drukgrad4} & = +6570.9800000000000 \\
x & = +272.5000000000000 \\
& = +545 / 2 \\
y & = +90.6420000000000 \\
& = +45321 / 500 \\
\text{druk} & = +888.8943693000000 \\
\text{drukgrad1} & = +4205.9181672152 \\
\text{deltap0} & = +9223.7411478607 \\
\text{deltauhor} & = +2.2960429612054 \\
\text{Uabs} & = +2.32442427748737 \\
\text{ughor} & = +2.32442427748737 + \text{deltauhor} \\
& = +4.6204672386928 \\
\text{phigo} & = +0.0013831297120385 \times \text{ughor} \\
& = +0.00639070521176 \\
\text{Ug} & = +1.24273457721456 - 0.0070684 \times \text{ughor} \\
& = +1.21007526658458 \\
\text{Us} & = +0.0023471372232845 \times \text{phis} \\
& = +0.01643008954884 \\
\text{phigas} & = +0.2431757000000000 \\
\text{relogasoverspraak} & = +0.56877792970426 \times \text{ughor} \\
& = +2.628019790290 \\
\text{tau4} & = +1704.203725422799 \times \text{phis} \\
& = +243.455763775 \\
\end{align*}
\]

Largest residual is 1.6E-10
Method: Exact
Numeric processor: 80387
Appendix II

CALCULATION OF CONDENSOR H41
Calculation of condenser H 41

The design calculations of the condenser H 41 are done following Kern's method as given in Coulson and Richardson (1983) on pages 545 to 553.

1. The calculation of the stream of cooling water needed to condense steam stream.

The amount of heat that has to be removed, \( Q \) is 209722 kW. The stream of cooling water needed can be calculated using:

\[
Q = M \cdot c_p \cdot \Delta T
\]

where

- \( Q \) is the heat to be removed, \( 209722 \) kW
- \( M \) is the mass flow of cooling water, \( 5017.3 \) kg/s
- \( c_p \) is the heat capacity, \( 4.18 \) kJ/kg °C
- \( \Delta T \) is the temperature difference, \( 10 \) °C

2. The calculation of the heat exchanging surface of the exchanger H 41.

The surface can be calculated using the following equations:

\[
Q = U \cdot A \cdot \Delta T
\]

where

- \( Q \) is the heat transferred, \( 209722 \) kW
- \( U \) is the overall heat transfer coefficient, \( \text{W/m}^2 \text{°C} \)
- \( A \) is the heat transfer area, \( \text{m}^2 \)
- \( \Delta T \) is the mean temperature difference, °C

The mean temperature difference can be calculated, using a temperature correction factor, \( F \), and the logarithmic mean temperature difference given by:

\[
\Delta T_{lm} = \frac{(T_1 - t_2) - (T_2 - t_1)}{\ln \left( \frac{T_1 - t_2}{T_2 - t_1} \right)}
\]

where

- \( T_1 \) is the inlet shell-side fluid temperature, \( 49 \) °C
- \( T_2 \) is the outlet shell-side fluid temperature, \( 30 \) °C
- \( t_1 \) is the inlet tube-side fluid temperature, \( 20 \) °C
- \( t_2 \) is the outlet tube-side fluid temperature, \( 10 \) °C

Choosing 20 mm outer diameter, 16 mm inner diameter, 4.83-m-long tubes, cupro-nickel, gives an area per tube of:

\[
\text{Area of one tube} = 4.83 \times 20 \times 10^{-3} \times \pi = 0.303 \text{ m}^2
\]

This gives a total number of tubes of:

\[
\text{Number of tubes, } N = \frac{25390}{0.303} = 83795 \text{ pipes}
\]
A 1.25 triangular pitch is used, with four passes, which leads to a bundle diameter $D_b$ of:

$$D_b = 20 \times \left( \frac{83795}{0.175} \right)^{\frac{1}{2.285}} = 6.1 \text{m}$$

A split-ring floating head is chosen. Extrapolating figure 12.10 of Coulson and Richardson (1983), leads to a shell-bundle clearance of 0.22 m. This leads to a shell diameter, $D_s$, of 6.32 m.

3. The calculation of the pressure drop over the tube side.

$$v = \text{velocity in the pipe} = 4 \times \frac{M_c \times N_p}{(\pi \times D_i^2 \times \rho \times N)}$$

$M_c = \text{mass flow cooling water} = 5017.3 \text{kg/s}$

$N_p = \text{number of passes} = 4$

$D_i = \text{inner tube diameter} = 16 \times 10^{-3} \text{m}$

$\rho = \text{density cooling water} = 998 \text{kg/m}^3$

$N = \text{number of pipes} = 83795$

This leads to a velocity $v$ of 1.2 m/s.

With this fact the pressure drop can be calculated using:

$$\Delta P_t = N_p \times (8 \times j_f \times \left( \frac{L}{D_i} \right) + 2.5) \times \frac{\rho \times v^2}{2}$$

$j_f = \text{the dimensionless friction factor which can be be gotten from figure 12.24 in Coulson and Richardson (1983)} = 3.5 \times 10^{-3}$

$L = \text{the length of a tube} = 4.83 \text{m}$

Using the equation a pressure drop over the tube side of 62 kPa is calculated which is acceptable.

4. The calculation of the tube side coefficient.

The mean water temperature is $0.5 \times (20+30) = 25 \text{°C}$.

The tube cross-sectional area is $(\pi/4) \times 16^2 = 201 \text{mm}^2$

The number of tubes per pass is $83795/4 = 20949$.

The total flow area is $20949 \times 201 \times 10^{-6} = 4.21 \text{m}^2$

The water mass velocity is $M_c/\text{(total flow area)} = 5017.3/4.21 = 1191.5 \text{kg/s}\text{m}^2$

The linear water velocity is the water mass velocity divided by the density is 1.21 m/s.

Now the tube side coefficient can be calculated with:

$$h_t = 4200 \times (1.35 + 0.02 \times \tau) \times \left( \frac{u_t^{0.8}}{d_i^{0.2}} \right)$$
where \( h_i \) = tube side coefficient, W/m\(^2\) °C
\( t \) = mean water temperature, °C
\( u_t \) = linear water velocity, m/s
\( d_i \) = tube inner diameter, m

This gives the coefficient a value of 5198 W/m\(^2\)°C.

5. The calculation of the shell side coefficient.

The baffle spacing \( l_b \) is one fifth of the shell diameter, equals 1.22 m. The tube pitch \( p_t \) is 1.25 times the outer tube diameter, equals 25 mm.

The cross-flow area, \( A_s \) can be calculated with:

\[
A_s = \frac{(p_t - d_o) \cdot D_t \cdot l_p}{p_t}
\]

So, \( A_s \) equals 1.49 m\(^2\).

The mass velocity \( G_s \) is calculated by dividing the mass flow steam by the cross-flow area, which gives a value of 61.4 kg/s/m\(^2\).

For an equilateral triangular pitch arrangement, the shell side hydraulic or equivalent diameter \( d_e \), is calculated with:

\[
d_e = \frac{1.10}{d_o} \cdot (p_t^2 - 0.917 \cdot d_o^2)
\]

So, \( d_e \) = 14.4 mm

mean shell temperature = 49 °C
density \( \rho \) = 992 kg/m\(^3\)
viscosity \( \eta \) = 653.3 e-6 Ns/m\(^2\)
heat capacity \( c_p \) = 4.136 kJ/kg°C
thermal conductivity \( \lambda \) = 0.578

Reynolds is given by:

\[
Rey = \frac{G_s \cdot d_e}{\eta} = 1353.4
\]

And Prandtl is given by:

\[
Pr = \frac{c_p \cdot \eta}{\lambda} = 4.73
\]
With a baffle cut of 25 % and figure 12.29 of Coulson and Richardson (1983), the shell side heat transfer factor \( j_h \) can be determined. It equals \( 1.5 \times 10^{-2} \).

The shell side heat transfer coefficient can be calculated using:

\[
h_s = \frac{\lambda}{d_e} j_h \cdot Re \cdot Pr^{\frac{1}{3}} = 1368 \text{ W/m}^2\text{C}
\]

6. The determination of the overall heat transfer coefficient \( U \).

The overall heat transfer coefficient can be determined using:

\[
\frac{1}{U} = \frac{1}{h_o} + \frac{1}{h_{od}} + \frac{d_o \ln(d_o/d_i)}{2k_w} + \frac{d_o}{d_i} \cdot \frac{1}{h_o} + \frac{d_o}{d_i} \cdot \frac{1}{h_{id}}
\]

where
\( h_o \) = outside fouling factor = 5000 W/m\(^2\)°C
\( h_{id} \) = inside fouling factor = 12000 W/m\(^2\)°C
\( k_w \) = thermal conductivity of the tube wall material = 380 W/m\(^2\)°C

This leads to an overall heat transfer coefficient of 116 W/m\(^2\)°C.

Using equation one, but with the new overall heat transfer coefficient of 116, leads to a new value from the exchanging area, namely 76565 m\(^2\). This area is three times as big as the first estimation. By making the tubes three times as long, the exchanger will operate correctly, if the tube side pressure drop poses no problem.

7. The tube side pressure drop recalculated for a new tube length.

The new tube length will be 76565/25390 times 4.83 m, equals 14.57 m. Using the equation of the pressure drop over the tube side again, but with this new tube length, gives a pressure drop of 79.665 kPa, which is still acceptable.
Appendix III

FLOWSHEETS
Appendix IV

MASS/HEAT BALANCES
<table>
<thead>
<tr>
<th>IN</th>
<th>Voorwaarts</th>
<th>Retour</th>
<th>UIT</th>
</tr>
</thead>
<tbody>
<tr>
<td>M</td>
<td>Q</td>
<td>M</td>
<td>Q</td>
</tr>
<tr>
<td>112.51</td>
<td>0</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>4539</td>
<td></td>
<td></td>
</tr>
<tr>
<td>121.80</td>
<td>8694</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>112.51</td>
<td>132.33</td>
<td></td>
</tr>
<tr>
<td></td>
<td>24294</td>
<td></td>
<td></td>
</tr>
<tr>
<td>0.027</td>
<td>37166.7</td>
<td></td>
<td></td>
</tr>
<tr>
<td>10.4</td>
<td>305445</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>609.2</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>7.214</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>1.08</td>
<td></td>
<td></td>
</tr>
<tr>
<td></td>
<td>360.23</td>
<td></td>
<td></td>
</tr>
<tr>
<td>121.86</td>
<td>342614.7</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

Diagram:

- Air
- Compression
- Fire
- Sorbent
- Coal
- Reaction
- Steam cycle
- Ash

Diagram nodes:

- C1
- H4
- H9
- R10
- P12
Massa in kg/s
Warmte in kW

Fabrieksvooronkwerp
No:

| 100.84 |
| 0.533  |
| 0.006  |
| 0      |
| 0.533  |
| 0      |
| 0.006  |
| 0      |
| 182.78 |
| 28.01  |
| 7.0    |
| 7.0    |
| 7.0    |
| 7.0    |

245.28 343795 → Totaal → 343314

Fluegas bed 3 + 4

1.1813 113
Appendix V

STREAM/COMPONENTS RECORDS
### Tabel van Apparaatstroom Componenten

<table>
<thead>
<tr>
<th>Componenten</th>
<th>M</th>
<th>Q</th>
<th>M</th>
<th>Q</th>
<th>M</th>
<th>Q</th>
<th>M</th>
<th>Q</th>
<th>M</th>
<th>Q</th>
</tr>
</thead>
<tbody>
<tr>
<td>O2</td>
<td>0.458</td>
<td></td>
<td>1.024</td>
<td></td>
<td>0.075</td>
<td></td>
<td>1.024</td>
<td></td>
<td>0.075</td>
<td></td>
</tr>
<tr>
<td>N2</td>
<td>1.21</td>
<td></td>
<td>0.0016</td>
<td></td>
<td>28.81</td>
<td></td>
<td>0.0016</td>
<td></td>
<td>28.81</td>
<td></td>
</tr>
<tr>
<td>Ar</td>
<td>4.60</td>
<td></td>
<td>0.0034</td>
<td></td>
<td>0.0</td>
<td></td>
<td>0.0</td>
<td></td>
<td>0.0</td>
<td></td>
</tr>
<tr>
<td>SO2</td>
<td>0.857</td>
<td></td>
<td>0.027</td>
<td></td>
<td>0.027</td>
<td></td>
<td>0.027</td>
<td></td>
<td>0.027</td>
<td></td>
</tr>
<tr>
<td>CO2</td>
<td>0.0027</td>
<td></td>
<td>0.0001</td>
<td></td>
<td>0.0001</td>
<td></td>
<td>0.0001</td>
<td></td>
<td>0.0001</td>
<td></td>
</tr>
<tr>
<td>H2</td>
<td>0.006</td>
<td></td>
<td>0.006</td>
<td></td>
<td>0.006</td>
<td></td>
<td>0.006</td>
<td></td>
<td>0.006</td>
<td></td>
</tr>
<tr>
<td>H2O</td>
<td>0.0008</td>
<td></td>
<td>0.0008</td>
<td></td>
<td>0.0008</td>
<td></td>
<td>0.0008</td>
<td></td>
<td>0.0008</td>
<td></td>
</tr>
<tr>
<td>NO</td>
<td>0.001</td>
<td></td>
<td>0.001</td>
<td></td>
<td>0.001</td>
<td></td>
<td>0.001</td>
<td></td>
<td>0.001</td>
<td></td>
</tr>
<tr>
<td>NH3</td>
<td>0.001</td>
<td></td>
<td>0.001</td>
<td></td>
<td>0.001</td>
<td></td>
<td>0.001</td>
<td></td>
<td>0.001</td>
<td></td>
</tr>
<tr>
<td>S2</td>
<td>0.00002</td>
<td></td>
<td>0.00002</td>
<td></td>
<td>0.00002</td>
<td></td>
<td>0.00002</td>
<td></td>
<td>0.00002</td>
<td></td>
</tr>
<tr>
<td>Ash</td>
<td>0.0001</td>
<td></td>
<td>0.0001</td>
<td></td>
<td>0.0001</td>
<td></td>
<td>0.0001</td>
<td></td>
<td>0.0001</td>
<td></td>
</tr>
<tr>
<td>Sorbent</td>
<td>0.000025</td>
<td></td>
<td>0.000025</td>
<td></td>
<td>0.000025</td>
<td></td>
<td>0.000025</td>
<td></td>
<td>0.000025</td>
<td></td>
</tr>
<tr>
<td><strong>Totaal:</strong></td>
<td>0.464</td>
<td>98.42</td>
<td>121.795</td>
<td>11.53</td>
<td>120.91</td>
<td>11735</td>
<td>1.096</td>
<td>929</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

**M in kg/s**

**Q in kW**

*Stroom/Componenten staat*
<table>
<thead>
<tr>
<th>Apparaatstroom Componenten</th>
<th>1</th>
<th>6</th>
<th>7</th>
<th>8</th>
<th>9</th>
</tr>
</thead>
<tbody>
<tr>
<td>M</td>
<td>Q</td>
<td>M</td>
<td>Q</td>
<td>M</td>
<td>Q</td>
</tr>
<tr>
<td>O2</td>
<td>26.17</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>N2</td>
<td>85.13</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>H2O</td>
<td>1.21</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Ar</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>SO2</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>H2S</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>CO2</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>H2</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>NO</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>NH3</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Totaal:</strong></td>
<td>112.51</td>
<td>4539</td>
<td>112.51</td>
<td>13233</td>
<td>0.006</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Apparaatstroom Componenten</th>
<th>10</th>
<th>11</th>
<th>12</th>
<th>13</th>
<th>14</th>
</tr>
</thead>
<tbody>
<tr>
<td>M</td>
<td>Q</td>
<td>M</td>
<td>Q</td>
<td>M</td>
<td>Q</td>
</tr>
<tr>
<td>O2</td>
<td>1.0243</td>
<td>0.25</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>N2</td>
<td>85.26</td>
<td>0.82</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Ar</td>
<td>1.21</td>
<td>0.01</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>SO2</td>
<td>0.0016</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>CO2</td>
<td>28.81</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>H2</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>H2O</td>
<td>4.602</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>NO</td>
<td>0.0034</td>
<td>0.006</td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>NH3</td>
<td>0.000004</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Ash</td>
<td>0.857</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Sorbent</td>
<td>0.027</td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Totaal:</strong></td>
<td>121.795</td>
<td>15448</td>
<td>1.08</td>
<td>360</td>
<td>0.006</td>
</tr>
</tbody>
</table>

*M in kg/s
Q in kW*

*Stroom/Componenten staat*
<table>
<thead>
<tr>
<th>Apparaatstroom Componenten</th>
<th>2</th>
<th>3</th>
<th>4</th>
<th>5</th>
<th>6</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ash</td>
<td>1.0712</td>
<td></td>
<td></td>
<td>1.0712</td>
<td></td>
</tr>
<tr>
<td>C</td>
<td>7.862</td>
<td></td>
<td></td>
<td>7.862</td>
<td></td>
</tr>
<tr>
<td>H</td>
<td>0.589</td>
<td></td>
<td></td>
<td>0.489</td>
<td></td>
</tr>
<tr>
<td>N</td>
<td>0.135</td>
<td></td>
<td></td>
<td>0.135</td>
<td></td>
</tr>
<tr>
<td>O</td>
<td>0.562</td>
<td></td>
<td></td>
<td>0.562</td>
<td></td>
</tr>
<tr>
<td>S</td>
<td>0.078</td>
<td></td>
<td></td>
<td>0.078</td>
<td></td>
</tr>
<tr>
<td>H2O</td>
<td>0.203</td>
<td></td>
<td></td>
<td>0.203</td>
<td></td>
</tr>
<tr>
<td>Al2O3</td>
<td></td>
<td>0.024</td>
<td></td>
<td>0.024</td>
<td></td>
</tr>
<tr>
<td>CaO</td>
<td></td>
<td>0.003</td>
<td></td>
<td>0.003</td>
<td></td>
</tr>
<tr>
<td>CaSO4</td>
<td></td>
<td>0.003</td>
<td></td>
<td>0.003</td>
<td></td>
</tr>
<tr>
<td>CaS</td>
<td></td>
<td></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Totaal:</td>
<td>10.4</td>
<td>0.027</td>
<td>0.027</td>
<td>10.4</td>
<td></td>
</tr>
</tbody>
</table>

M in kg/s
Q in kW

Stroom/Componenten staat
<table>
<thead>
<tr>
<th>Apparaatstroom Componenten</th>
<th>Bed 1 → Bed 2</th>
<th>Bed 2 → Bed 3</th>
<th>Bed 3 → Bed 4</th>
<th>Bed 4 → Bed 1</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ash</td>
<td>M 0.214</td>
<td>Q</td>
<td>M 6.376</td>
<td>Q 0.132</td>
</tr>
<tr>
<td>A12O3</td>
<td>M 6.376</td>
<td>Q 0.132</td>
<td>M 6.376</td>
<td>Q 0.041</td>
</tr>
<tr>
<td>CaSO4</td>
<td>M 0.132</td>
<td>Q 0.132</td>
<td>M 6.376</td>
<td>Q 0.041</td>
</tr>
<tr>
<td>CaO</td>
<td>M 0.492</td>
<td>Q 0.492</td>
<td>M 0.560</td>
<td>Q 0.022</td>
</tr>
<tr>
<td>CaS</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>Totaal:</td>
<td>M 7.214</td>
<td>Q 7.0</td>
<td>M 7.0</td>
<td>Q 7.0</td>
</tr>
</tbody>
</table>

M in kg/s
Q in kW

Stroom/Componenten staat
Appendix VI

APPARATUSES LISTS
### WARMTEWISSELAARSPECIFIEKATIEBLAD

**Apparaatnummer : H. 41**  
**Aantal : 18, Reeks/parallel**

#### ALGEMENE EIGENSCHAPPEN :

**Functie :** Condenseren, stoom in stoomcyclus

| Type | Wärmtewisselaar*
|------|---------------------
|      | Koeier             
|      | Kondensor          
|      | Verdampen         

**Uitvoering :** met vaste pijpplaten*

* floating head  
* haarspeld  
* dubbele pijp  
* platenwarmtewisselaar

**Positie :** horizontaal/verticaal*

#### BEDRIJFSKONDITIES :

<table>
<thead>
<tr>
<th>Soort fluidum</th>
<th>Mantelzijde</th>
<th>Pijpzijde</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>STEAM</td>
<td>COOLING WATER</td>
</tr>
<tr>
<td>Massastroom</td>
<td>91.39 kg/s</td>
<td>5017.3 kgf/s</td>
</tr>
<tr>
<td>Massastroom te verdampen/kondenseren</td>
<td>21.39 kg/s</td>
<td></td>
</tr>
<tr>
<td>Gemiddelde soortelijke warmte</td>
<td>2383 kJ/kg °C</td>
<td>4.18 kJ/kg °C</td>
</tr>
<tr>
<td>Verdampingswarmte</td>
<td>49 kJ/kg</td>
<td>20</td>
</tr>
<tr>
<td>Temperatuur IN</td>
<td>49 °C</td>
<td>30 °C</td>
</tr>
<tr>
<td>Temperatuur UIT</td>
<td>0.12 °C</td>
<td>1 °C</td>
</tr>
<tr>
<td>Druk</td>
<td>STAINLESS</td>
<td>COPRO-NICKEL</td>
</tr>
<tr>
<td>Materiaal</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>

*Doorstrepen wat niet van toepassing is
## Apparatenlijst voor pompens, blowers, kompressoren

<table>
<thead>
<tr>
<th>Apparaat No:</th>
<th>P 43</th>
<th>P 31</th>
<th>C 1</th>
<th>C 6</th>
<th>C 6</th>
</tr>
</thead>
<tbody>
<tr>
<td>Benaming, type</td>
<td>High pressure pump</td>
<td>Circulation pump</td>
<td>centrifugal compressor</td>
<td>Centrifugal compressor</td>
<td>Centrifugal compressor</td>
</tr>
<tr>
<td>te verpompen medium</td>
<td>Water</td>
<td>Water</td>
<td>Air</td>
<td>Hydrogen</td>
<td>Nitrogen</td>
</tr>
<tr>
<td>Capaciteit in kg/s</td>
<td>91.39</td>
<td>91.39</td>
<td>112.51</td>
<td>0.006</td>
<td>0.533</td>
</tr>
<tr>
<td>Dichtheid in kg/m³</td>
<td>988</td>
<td>688</td>
<td>1.18</td>
<td>1.18</td>
<td>1.18</td>
</tr>
<tr>
<td>Zuig-/persdruk in bar abs.</td>
<td>0.12 / 100</td>
<td>100 / -</td>
<td>1.0 / 1.4</td>
<td>1.0 / 1.4</td>
<td>1.0 / 1.4</td>
</tr>
<tr>
<td>temp. in °C in / uit</td>
<td>49 / 52</td>
<td>311 / -</td>
<td>25 / 60</td>
<td>25 / 60</td>
<td>25 / 60</td>
</tr>
<tr>
<td>Vermogen in kW theor./prakt.</td>
<td>1260</td>
<td>-</td>
<td>3404</td>
<td>2.61</td>
<td>16.67</td>
</tr>
<tr>
<td>Speciaal te gebruiken mat.</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>aantal serie/parallel</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
</tr>
</tbody>
</table>
## Apparaatenlijst voor warmtewisselaars, fornuizen

<table>
<thead>
<tr>
<th>Apparaat No.:</th>
<th>H 8</th>
<th>H 11</th>
<th>H 9</th>
<th>H 13</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Benaming, type</strong></td>
<td>fluegas / airheater</td>
<td>fluegas / airheater</td>
<td>gas airheater</td>
<td>gas airheater</td>
</tr>
<tr>
<td><strong>Medium</strong></td>
<td>fluegas / air</td>
<td>fluegas / nitrogen</td>
<td>air</td>
<td>nitrogen</td>
</tr>
<tr>
<td><strong>Capaciteit, uitgewisselde warmte in kW.</strong></td>
<td>8694</td>
<td>84</td>
<td>2429</td>
<td>101</td>
</tr>
<tr>
<td><strong>Warmtewisselend oppervl. in m²</strong></td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td><strong>Aantal serie/paralleel</strong></td>
<td>1</td>
<td>2</td>
<td>1</td>
<td>1</td>
</tr>
<tr>
<td><strong>Abs. druk in bar</strong></td>
<td>1.4 / 1.0</td>
<td>1.4 / 1.0</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td><strong>temp. in / uit in °C</strong></td>
<td>65 / 144</td>
<td>N2 60 / 172</td>
<td>144 / 350</td>
<td>172 / 350</td>
</tr>
<tr>
<td><strong>pijpzijde</strong></td>
<td>190 / 120</td>
<td>170 / 120</td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>mantelzijde</strong></td>
<td>H2 60 / 350</td>
<td>190 / 170</td>
<td></td>
<td></td>
</tr>
<tr>
<td><strong>Speciaal te gebruiken mat.</strong></td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
</tr>
<tr>
<td>Apparaat No:</td>
<td>V 33</td>
<td>V 42</td>
<td></td>
<td></td>
</tr>
<tr>
<td>-------------</td>
<td>------</td>
<td>------</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Benaming, type</td>
<td>Steamdrum</td>
<td>Drum</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Capaciteit</td>
<td>-</td>
<td>-</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Abs. druk in bar</td>
<td>100</td>
<td>0.12</td>
<td></td>
<td></td>
</tr>
<tr>
<td>temp. in °C</td>
<td>311</td>
<td>49</td>
<td></td>
<td></td>
</tr>
<tr>
<td>Inhoud in m³ of afmetingen</td>
<td>-</td>
<td>-</td>
<td></td>
<td></td>
</tr>
<tr>
<td>in m</td>
<td>-</td>
<td>-</td>
<td></td>
<td></td>
</tr>
<tr>
<td>aantal serie/parallel</td>
<td>-</td>
<td>-</td>
<td></td>
<td></td>
</tr>
</tbody>
</table>