

CORROSION PREDICTION AND MATERIALS SELECTION FOR OIL AND GAS PRODUCING ENVIRONMENTS.

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ABSTRACT

A computer model is described which predicts corrosion rates of carbon steel flowlines and tubing in the presence of carbon dioxide. The model can be applied also for processing facility piping and equipment. The influences of dissolved corrosion products, hydrogen sulfide, crude oil type, acetic acid and flow regime on the corrosion rate are predicted. Inhibition scenarios can be compared and, for highly corrosive environments, the model incorporates the evaluation of a number of Corrosion Resistant Alloys and incorporates a life-cycle cost evaluation.

KEYWORDS

Corrosion rate prediction, flowlines, tubing, materials selection, hydrogen sulphide, life cycle costing

INTRODUCTION

The estimation of corrosion rates and selection of materials for oil and gas producing systems involves a number of logical steps that have to be taken in a consistent way. Preparing a computer model forces the program writer to establish fixed calculations and decision-making steps based upon a broad knowledge of the subject. The benefit to the user of the program is that there is no ambiguity about the reason for a particular corrosion allowance requirement or material selection decision. There is also an auditable decision making trail which can be interrogated at any stage, because it is clearly fixed, and there is consistency of approach when two or more persons are asked to make the same evaluation. The computerised approach also lends itself to extra analysis of the available data, such as comparison of options on a life-cycle cost basis.

In preparing the “Electronic Corrosion Engineer”¹ (ECE) computer program for corrosion rate estimation and materials selection, a deliberate attempt was made to mimic very closely the decision-making steps which corrosion engineers have to make for oil and gas production systems. A benefit of this approach is that the software is very user-friendly and intuitive, for a relative new-comer to the subject, whilst it also offers some quite sophisticated analytical capability for the more experienced corrosion engineer. It has not been our policy to create a “black box” which provides an answer without explanation; the model has extensive technical support in the “help” files with various references provided for further study.

¹ Trade name

This paper concentrates on outlining in more detail that technical background, particularly highlighting some of the more advanced features that have been incorporated into the latest version.

BASIC OVERVIEW

The software is aimed to provide guidance on two key capital items in any project – the production tubing and flowlines. This concentration on these items, essential pipe in either vertical or horizontal flow regimes, allowed the corrosion model to be optimised for these components and simplified the range of materials to be covered in the material selection area to about 8 key alloys for each component. It is also possible to use the program for the estimation of corrosion rates and the evaluation of materials for use in facilities.

The essential part of the software is the “Corrosion Predictor” which calculates the corrosion rate of carbon steel from the given input data. Data is limited to the critical inputs necessary, with some default values controllable “for expert use”. For ease, and compatibility with the diverse units that are common in the industry, all inputs can be made in either metric or imperial units.

The output of the Corrosion Predictor is a graph of corrosion rate with depth or distance along the line; a value reported for the maximum corrosion rate; a graph of the risk of failure as a function of time; a graph of pH variation with distance and a series of output data related to fluid velocities, estimated flow rates, partial pressures of acid gases, an indication of the likely flow regime at the output of the system and evaluation of sour service requirements and API RP 14E erosional velocity limit (when appropriate).

Additional “Tools” which can be drawn down are a “CRA Evaluator” for alloy selection, a “CRA Manufacturers” database and a “Life-cycle cost calculation”. These automatically input data from the Corrosion Predictor, but data can be overwritten so that these tools can be used independently. A consistent approach is used for data input on the left side of the screen and data output on the right to facilitate ease of use.

DESCRIPTION OF THE CORROSION MODEL

Model Calculation Steps

The model calculates the CO₂ corrosion rate for 100 equidistant points along the length of the pipe. In order to reduce the number of less readily available input parameters, pressure and temperature are assumed to vary linearly with distance. This has the advantage that the influence of pressure or temperature can be shown in the resulting graphs of corrosion rate vs distance. When needed for situations with non-linear temperature profile, the full pipe length may be split in smaller lengths, in such a way that deviations from linearity are smaller.

Each point corresponds with a small section of the pipe for which the corrosion rate is determined from the CO₂ content, pressure, temperature, flow velocity and pH, taking as input parameter the dissolved iron concentration developed in the previous section, and the concentration of other bicarbonate salts. The corrosion rate is corrected for the effects of the presence of crude oil or gas condensate, H₂S, acetic acid and inhibition.

The pH in each section is calculated from the concentration of the following species:



The effective carbonic acid concentration is calculated from the CO₂ fugacity and Henry's constant, which is temperature dependent. The concentrations of the carbonate and bicarbonate species then follow from the dissociation constants, which are also temperature dependent. For the H₂S derived species the calculations are analogous.

In each section the pH is calculated by determining the point where the following two conditions are obeyed:

1. charge balance of all dissolved ions
2. the flux of dissolved iron, with contributions from corrosion and precipitation, and in- and out-flow, balances.

Depending on temperature, the resulting dissolved iron concentration can lead to super saturation in terms of FeCO_3 , with concurrent increase in pH.

In the presence of H_2S , however, when the FeS solubility product is exceeded, the pH follows from a charge balance only, of the dissolved ions HCO_3^- , CO_3^{2-} , HS^- , S^{2-} , H^+ , OH^- , Fe^{++} , and CH_3COO^- (acetate).

Also, in each section, the volumes of oil, water and gas are calculated, including the amount of water condensing from gas. In the case that glycol is injected for corrosion rate reduction, the extra water absorbed from the gas is taken into account, together with the change in glycol concentration. From these volumes of liquid an estimate is then made of the liquid hold up. The latter is used to obtain the velocity of the liquid, which is used as an input for the corrosion rate calculation. Since a complete profile of the pipe is seldom available, a complete calculation of the hold up is not included in the model. However, hold-ups calculated from independent hydrodynamic models can be used for input.

For horizontal flow, the outlet velocity of the gas is used to estimate the likelihood of stratified flow. When this is the case, a different corrosion rate is calculated for the top of the line. Furthermore, the occurrence of stratified flow has an impact on the effect of the presence of crude oil on the corrosion rate. In the case of annular horizontal flow, the effect of "low spots", where water can still form a separate layer of water, can be evaluated. This is done by repeating the calculation without the protection resulting from the presence of oil, if any.

When the flow is annular, the velocity of the gas is also used to predict the onset of erosion, using API RP 14E as a criterion. When the erosional velocity is exceeded at any point of the line, reduction of corrosion through protective layers, e.g. carbonate and sulfide layers, or inhibitor film, is no longer taken into account.

The accumulated risk of failure (or survival) vs. time is calculated from the maximum corrosion rate along the length of the pipe, using a standard deviation of 25%, and assuming a normal distribution. When H_2S is present, the distribution is changed by adding a contribution to reflect early failure by pitting.

CO₂ corrosion rate calculation

For normalized steels the following base equations¹ are used in the model:

$$\log(V_r) = 4.84 - \frac{1119}{t + 273} + 0.58 \log(f_{\text{CO}_2}) - 0.34 (\text{pH}_{\text{actual}} - \text{pH}_{\text{CO}_2}) \quad \text{Eq. 1}$$

and

$$V_m = 2.8 \frac{U^{0.8}}{d^{0.2}} f_{\text{CO}_2} \quad \text{Eq. 2}$$

where t is the temperature ($^{\circ}\text{C}$), U is the liquid velocity, d the diameter, and f_{CO_2} is the fugacity of the CO_2 :

$$\log(f_{\text{CO}_2}) = \log(\text{pCO}_2) + \left(0.0031 - \frac{1.4}{t + 273}\right) P \quad \text{Eq. 3}$$

pCO_2 being the partial pressure of the CO_2 (=mol% x total pressure P).

$\text{pH}_{\text{actual}}$ and pH_{CO_2} are the actual pH and the pH of pure water and CO_2 , respectively. All units are expressed in the S.I. system (kg, m, sec etc). The actual constants to be used in the above equations are slightly dependent on steel composition i.e. carbon content.

For quenched and tempered steels these equations are slightly different.

The resulting corrosion rate V_{cor} can be expressed by means of the "resistance model":

$$\frac{1}{V_{cor}} = \frac{1}{V_r} + \frac{1}{V_m} \quad \text{Eq. 4}$$

V_r is mainly determined by reaction kinetics of the reduction processes, but V_m is determined by the rate of mass transport of carbonic acid to the steel's surface, i.e. liquid velocity.

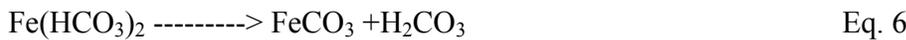
These equations represent a best fit to a large number of flow-loop data measured at IFE (Institutt For Energiteknikk) in Norway, where test conditions and environments were strictly controlled¹. From a regression analysis with these data, the error distribution obtained with this fit had a standard deviation of 25%.

The result has to be corrected for the presence of protective scale, H₂S, crude oil or condensate, glycol and inhibitor by means of multipliers described below:

$$\text{Corrosion Rate} = V_{cor} \times F_{scale} \times F_{h2s} \times F_{cond} \times F_{oil} \times F_{inhib} \times F_{glyc} \quad \text{Eq. 5}$$

Effect of dissolved iron (bi)carbonate

An important role is played by dissolved iron bicarbonate, which is the initial corrosion product formed. Although the solubility of FeCO₃ is quite low, the slow reaction:



This causes a supersaturation, with dissolved Fe⁺⁺, i.e. the concentration is higher than the solubility product of FeCO₃ would allow. This increases the pH (=pH_{actual} in Eq. 1) and reduces the corrosion rate. The extent of supersaturation with Fe(HCO₃)₂ is temperature dependent (more supersaturation, because of the lower solubility, at lower temperatures), and plays a major role at temperatures < 60 °C.

The precipitation flux of FeCO₃ (=removal rate of Fe(HCO₃)₂ from solution) can be calculated from²:

$$\left[\dot{Fe}^{2+} \right]_{precip} = k_r \frac{A}{V} K_{sp} (S - 1)(1 - S^{-1}) \quad \text{Eq. 7}$$

$$S = \frac{[Fe^{2+}][CO_3^{2-}]}{K_{sp}} \quad \text{Eq. 8}$$

where k_r and K_{sp} are the precipitation rate constant and the solubility product of FeCO₃, A/V is the surface area/volume ratio and S is the supersaturation level.

For a once-through system like a pipeline, the concentration profile of the dissolved Fe⁺⁺ ions will be unchanging.

In general, it is not exactly known what the concentration of dissolved iron is in the water entering the inlet of a pipe. Although in practice this concentration will be often saturated or supersaturated with respect to iron carbonate, much of the data generated with laboratory tests refer to water without dissolved iron. For this reason the model offers the opportunity to choose between no dissolved iron, saturated or supersaturated with iron for water entering the pipe.

The difference can be appreciable, depending on conditions. An example is given in Fig. 1, together, with a graph demonstrating, the influence of pH and dissolved Fe, on corrosion and precipitation of iron carbonate.

Effect of High temperature Carbonate Scaling

The scaling factor F_{scale} , which accounts for the lower corrosion rate when a protective FeCO_3 scale is formed, above about 60 °C, depends on CO_2 pressure:

$$\log(F_{\text{scale}}) = \frac{2400}{t + 273} - 0.6 \log(f\text{CO}_2) - 6.7 \quad \text{Eq. 9}$$

Influence of H_2S

The H_2S factor accounts for the reduction in corrosion rate when protective FeS is formed:

$$F_{\text{H}_2\text{S}} = \frac{1}{1 + 1800 \frac{p\text{H}_2\text{S}}{p\text{CO}_2}} \quad \text{Eq. 10}$$

the ratio of partial pressures $p\text{CO}_2/p\text{H}_2\text{S}$ is often used³ to distinguish CO_2 dominated corrosion from protection by FeS .

The above equation gives a 90% reduction of corrosion at $p\text{CO}_2/p\text{H}_2\text{S}$ of 200. This factor is highly speculative, since the FeS layer can sometimes suffer from isolated spots of breakdown and the subsequent occurrence of pitting corrosion.

These pitting corrosion rates are often found to be in the order of magnitude of the worst case pitting corrosion rate with CO_2 only, without the protection from a sulfide layer. For this reason the model does an additional calculation of the corrosion rate without the filming effect. It should be appreciated that the presence of H_2S also has a profound impact on the chemistry –and corrosivity- of the corrosive fluid, because the concentration of dissolved iron ions can be reduced through the precipitation of FeS , which decreases the pH. For this reason this pitting rate is often somewhat higher than with CO_2 only. An example is shown in Fig. 2 where the corrosion rate quickly reduces as the concentration of H_2S increases and an adherent iron sulfide film is formed. The difference between the pitting rate (the corrosion rate calculated without the H_2S effect) and the filmed surface corrosion rate may be indicative of the likelihood that pitting occurs: a larger difference may indicate a tendency to form fewer pits. Such behaviour has been observed in many field situations where failure in H_2S -containing environments can be highly localised at very few pits.

The standard deviation for the prediction of corrosion rate *including* H_2S is larger (=less certain) and has been set to 30% (compared to 25% for CO_2 -only situations). Furthermore, to reflect the uncertainty that failure may arise from pitting, the accumulated risk of failure has been increased arbitrarily by adding a 25% chance that the corrosion rate is based entirely upon the rate without H_2S present. Taking this approach, when H_2S is present, the risk of loss of protection of the sulfide layer shows up as a plateau early in the life in the accumulated risk curve (Fig.3), when, of course, the “ CO_2 -only” corrosion rate is high enough to show the effect. In this way the model warns the user not to take the low corrosion rates often experienced in the presence of H_2S at face-value.

Top-of-line corrosion

When the flow pattern is stratified, the corrosion at the top of the line is controlled by the rate of condensation of water from gas and the composition of the condensing phase at the top quadrant of the pipe. This is accounted for by the condensation factor F_{cond} . This rate is often lower than at the bottom; for practical situations when the condensation rate is below 0.25 $\text{g/m}^2/\text{s}$, this factor is conservatively set at 0.1. This reflects the effects of the lower velocity and of the saturation with corrosion products for the condensing phase.

However, the composition of the condensing phase also has to be taken into account: dissolved bicarbonate salts (other than iron bicarbonate) and crude oil are only present in the liquid at the bottom. The same holds for most chemical corrosion inhibitors, and this can result in a top-of-line corrosion rate, which can exceed the one at the bottom.

Influence of crude oil or condensate

The oil factor, which describes the effect of the presence of crude oil (or gas condensate) on the corrosion rate:

$$F_{oil} = 0.059 \frac{W}{W_{break}} U_{liq} + \frac{1.1 \cdot 10^{-4}}{W_{break}^2} \frac{\alpha}{90} + 0.059 \frac{W}{W_{break}} U_{liq} \frac{\alpha}{90}, \quad \text{Eq. 11}$$

where W is the water fraction in the oil, U_{liq} =liquid velocity, α = angle with vertical, and

$$W_{break} = -0.0166 \times API + 0.83 \quad \text{Eq. 12}$$

is the watercut where the water/oil emulsion just breaks, as function of the API density ($50 > \text{API} > 20$). High API density oils or gas condensates give little or no protection against corrosion since their water/oil emulsion breaks too easily, leaving the water free to contact the steel's surface. In these cases the amount of water is not relevant: at any watercut the fluid can corrode the steel.

For low horizontal flow velocities < 1 m/s, the oil factor is =1, since the water separates from the oil phase on the bottom of the line⁴. For diameters > 0.2 m this critical velocity is probably somewhat higher. At low velocities, the watercut itself is not important in influencing the actual corrosion rate: only the size of the corroding area will depend on watercut, but not the depth of the corrosion pits. Higher watercuts will wet a larger area and therefore allow a larger corroded area.

The oil factor was derived from a field study of tubing corrosion of two oil fields⁵.

Depending on watercut and type of oil, the angle of deviation of a well tubing, can have a profound influence on the corrosion rate. This angle is, in general, not constant throughout the depth of a well, and the model provides the possibility to input the well profile, when known, or to study the effect of various angles at various depths (Fig. 5).

Apart from changing the flowrate of oil or water separately, the model also provides the possibility to change the watercut of the liquid, while keeping the total liquid velocity constant. This allows, for example, the prediction of the changing corrosion rate of a well with anticipated future increase in water-cut, while keeping the production rate constant.

Influence of Acetic Acid

The mass transfer term, V_m , includes only the transport of one corrosive species, carbonic acid, as shown by Eq. 4. This equation can be modified to account for the additional transport of acetic acid, HAc, to the metal surface using:

$$V_{m(HAc)} = k_m [HAc]_{undiss} \quad \text{Eq. 13}$$

where $V_{m(HAc)}$ is the mass transfer term for HAc in mm/yr, k_m is the mass transfer rate of HAc and $[HAc]_{undiss}$ is the bulk concentration of undissociated HAc:

$$[HAc]_{undiss} = \frac{[HAc]_{total}}{1 + \frac{K_{diss}}{[H^+]}} \quad \text{Eq. 14}$$

Similar to CO_2 (Eq. 2), the mass transfer rate of HAc can be found from

$$k_m = \frac{D_{HAc}^{0.7} U^{0.8}}{\nu^{0.5} d^{0.2}} \quad \text{Eq. 15}$$

where D_{HAc} is the diffusion coefficient of HAc in water, and ν is the kinematic viscosity of water.

The resulting corrosion rate then becomes:

$$\frac{1}{V_{corr}} = \frac{1}{V_r} + \frac{1}{V_{m(H_2CO_3)} + V_{m(HAc)}} \quad \text{Eq. 16}$$

where $V_{m(H_2CO_3)}$ is the mass transfer rate of carbonic acid found from Eq. 2. This is called the George-DeWaard-Nesic model, which has been shown recently to explain satisfactorily results obtained from rotating cylinder experiments⁶.

Figure 6, is an example is given of the effect of the HAc concentration on the corrosion rate at top and bottom. As a consequence of the way the top-of-line corrosion is calculated -corrosion rate $\times F_{cond}$ for liquid without dissolved iron- there is more undissociated HAc in the top than at the bottom, leading to a relatively high effect of the presence of HAc at the top of the line. Although the approach is probably very conservative, this would be in line with recent findings of Gunaltun⁷, where acetic acid concentrations > 2000 ppm led to excessive top of line corrosion.

Inhibition

When a corrosion inhibitor is added to the system, its effect is calculated from its availability –using the multiplier F_{inhib} – at all times. With this approach, for example, operation of a pipe for one week per year without inhibitor corresponds to a factor of 0.96. This implies the assumption that the inhibitor is 100% efficient when present in the correct concentration.

It is important to note that the commonly applied inhibitors are not effective against top of line corrosion: they remain in the bottom phase.

For tubing there is also the option, for oil wells, of using squeeze inhibition. Here the inhibitor solution is injected into the producing reservoir formation, and then slowly released into the produced oil. On average, the inhibition effect inside the tubing is then built up in about one month, and then reduces slowly in about 2 months. The corrosion model can add the effect of consecutive squeezes for the calculation of the average corrosion rate per year.

Inhibition may prevent, or postpone, the onset of pitting on a sulfided surface. However, once pitting starts, the inhibitors may be unable to repair the degraded protective films, particularly if they have not been selected for the exact operating conditions⁸. Thus a conservative approach is taken that under sour conditions, the effect of (continuous) inhibition on the predicted *pitting* rate is set to zero ($F_{inhib} = 1$). Inhibitors do contribute to a reduced iron sulphide filming corrosion rate, and practical experience shows that they can reduce the *probability* of pitting taking place.

Effect of Glycol

Glycol addition may be regarded as a special case of inhibition. In wet gas pipelines glycol (DEG) is sometimes injected to control gas hydrate formation in wet gas lines and, at higher dosing rates, control corrosion. In the case where high injection rates of glycol are used for corrosion control, the glycol can be recovered at the outlet, dried and re-injected via a "piggy-back" small diameter line at the inlet. The reduction in corrosion rate is given by the glycol factor F_{glyc} :

$$\log F_{glyc} = 1.6 [\log (W\%) - 2] \quad \text{Eq. 17}$$

where $W\%$ is the water concentration in the water/glycol mixture. The glycol also acts as a drying agent, and also reduces the water condensation rate at the top of the line, while the condensing phase also contains glycol, and also reduces top-of-the-line corrosion. For such an approach to succeed, it is essential that all liquid water is removed before the gas enters the pipeline. From the knocked-out liquids, gas condensate is normally reinjected, but, as shown

above, this does not add much to corrosion rate reduction. The glycol concentration becomes smaller with distance into the pipeline. The model calculates how much water is absorbed from the gas into the water/glycol liquid at each point in the line, as a function of temperature and pressure, and then applies the glycol factor. Details about the calculation procedure have been published elsewhere^{9,10}. An example of the output of the model is given in Fig.7

DESCRIPTION OF THE MATERIAL SELECTION MODEL

Alloy Selection

The input data for the material selection model are temperature, pressure, CO₂ and H₂S content of the gas phase (or equilibrium gas phase), dissolved chloride and bicarbonate content of the water. From this data the room temperature pH is calculated. All those input data are then used in a series of mathematical and logical relationships to consider the suitability of about 8 corrosion resistant alloys (CRAs) which are typically applied for downhole tubing or flowline applications. The criteria for the acceptability of an alloy for a given set of conditions are that it should not show either generalised corrosion, or localised attack such as pitting or stress corrosion cracking. The basis of the environmental evaluation is from a combination of reviews of published data of laboratory test results and actual failure experiences.

The performance of CRAs is often a complex combination of environmental parameters. Typically the behaviour with respect to localised pitting or cracking may change significantly as a function of temperature, chloride content, H₂S partial pressure and pH (a function of CO₂ content and bicarbonate concentration as well as H₂S content). Thus, in many cases, it is necessary to think of the performance of materials using 3-dimensional graphs derived at a range of conditions. The material selection tool has incorporated, mathematically, all the 3-D graphs published previously by Craig as part of the evaluation of alloys performance¹¹, and added further data so that all alloys are considered against a good basis of laboratory and field experience in both CO₂- and H₂S-containing service conditions. For any set of parameters, –temperature, pressure, environment- the model interpolates linearly between actual data points on these 3-D surfaces, to establish whether an alloy is acceptable with respect to corrosion resistance. Cracking resistance is mostly evaluated from the same parameters on the basis of algebraic equations fitted to boundaries between "pass" and "fail" data.

Individual alloys are considered to be indicative of a group of alloys of similar chemical composition, so small variations in performance of individual manufacturers' specific grades of materials are not intended to be reflected in the evaluation.

Fig. 8 illustrates the typical output format using traffic lights to identify materials which are safe or unsafe in the stated conditions. An amber light is used in some cases to show when a material is getting close to a marginal set of conditions.

Life-Cycle Cost Analysis and Suppliers Database

For both downhole tubing and for flowlines there is a life-cycle cost evaluation tool which can be used to give a rough indication of the relative cost difference between the CRA option and carbon steel with or without corrosion inhibitor injection. The LCC approach has been well documented both for downhole tubing¹² and for flowlines¹³. The LCC tool incorporates the future operating costs of injecting inhibitor, either continuously or, for downhole tubing, on a batch basis. For flowlines a simplified approach for evaluating the cost of welding is incorporated, as this can be a significant factor in the capital cost of a CRA flowline. For all costs components the programme provides typical reasonable values for materials, welding, inhibitors, oil and gas but in all cases the user is warned that current figures should be used if an accurate comparison is to be made.

The LCC tool is useful for getting a feel for the relative costs of the CRA option and carbon steel, and particularly useful for highlighting how overall life-cycle costs can be high when inhibition is required for corrosion control. It also shows, by steps in the curve, where replacement of carbon steel may be required due to corrosion. Whilst this may be the cheapest material in the short term, for specific conditions and project circumstances, it may be the more expensive option for a longer service life project (Fig. 9).

Finally, if the decision is reached to select a corrosion resistant alloy for the tubing or flowline, there is a database of manufacturers which can be addressed to establish how many suppliers are available in the required dimensions and alloy type. Contact details for those manufacturers are given. Because such information may change in the short term, this part of the program is the only part where the users can over-write and save changes so that they can keep the database up to date, or replace global contact details with their local supplier information if preferred.

CONCLUSIONS

The use of a computerised model for corrosion analysis and materials selection for tubing, flowlines and facilities is beneficial in forcing a rigorous and consistent approach to be taken. A well-constructed model, closely mimicking the logical steps taken by materials engineers in selecting materials provides a very useful and user-friendly tool, making the job easy and providing a well-founded basis for conclusions reached. The model described in this paper is based upon good use of the best research data and field experience gathered regarding the performance of carbon steels and corrosion resistant alloys commonly used in the oil and gas production environment.

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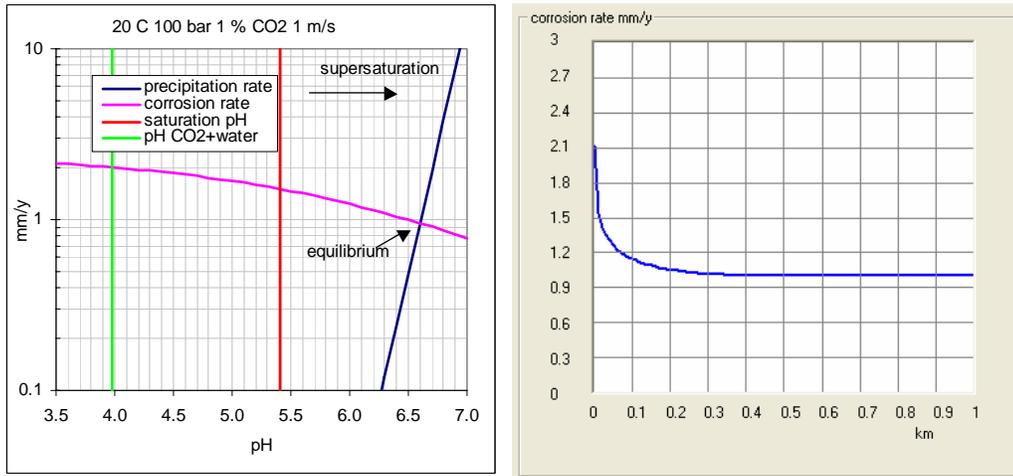


Figure 1. Example of effect of build-up of dissolved iron concentration on pH and corrosion rate. Left: calculation scheme, right: corresponding model output. Waterfilled pipe, $d=0.3$ m, conditions as shown.

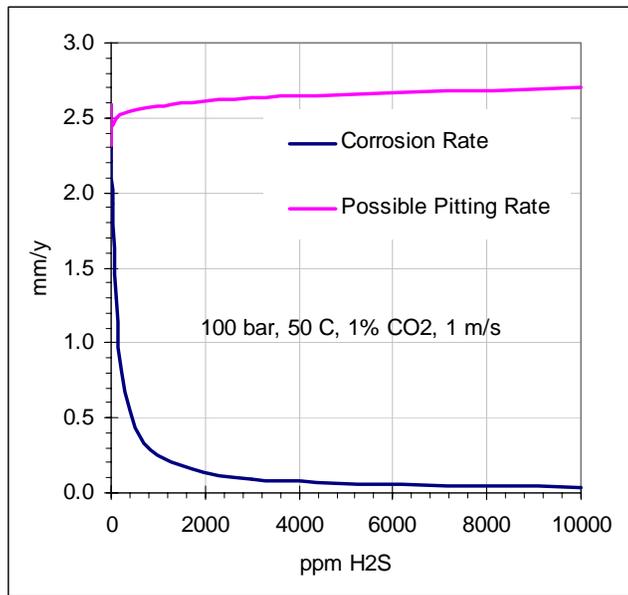


Figure 2. Example of effect of H₂S concentration from Eq. 10. Iron concentration was supersaturated in FeCO₃.

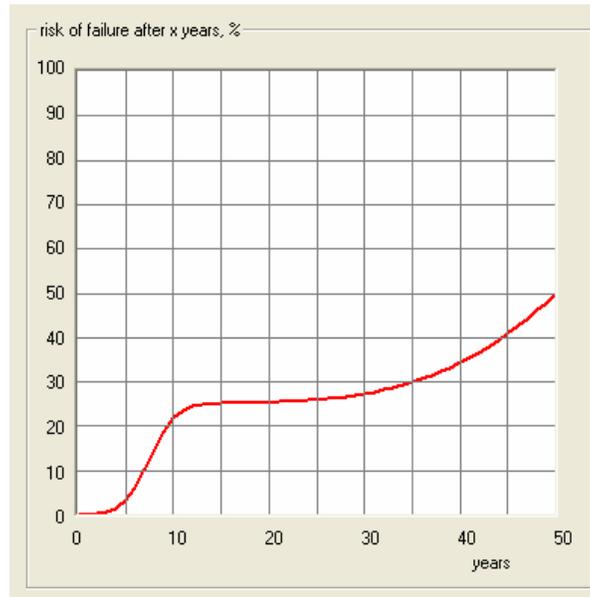


Figure 3. Example of effect of H₂S on risk of failure. Conditions: 0.5 m/s, 20 °C, 1 % CO₂ at 100 bar, 0.15% H₂S.

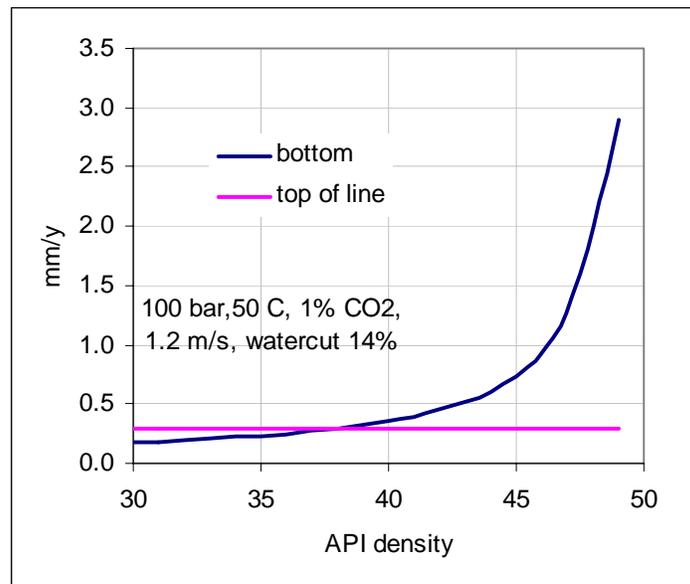


Figure 4. Example of top and bottom of line corrosion rates as a function of oil density.

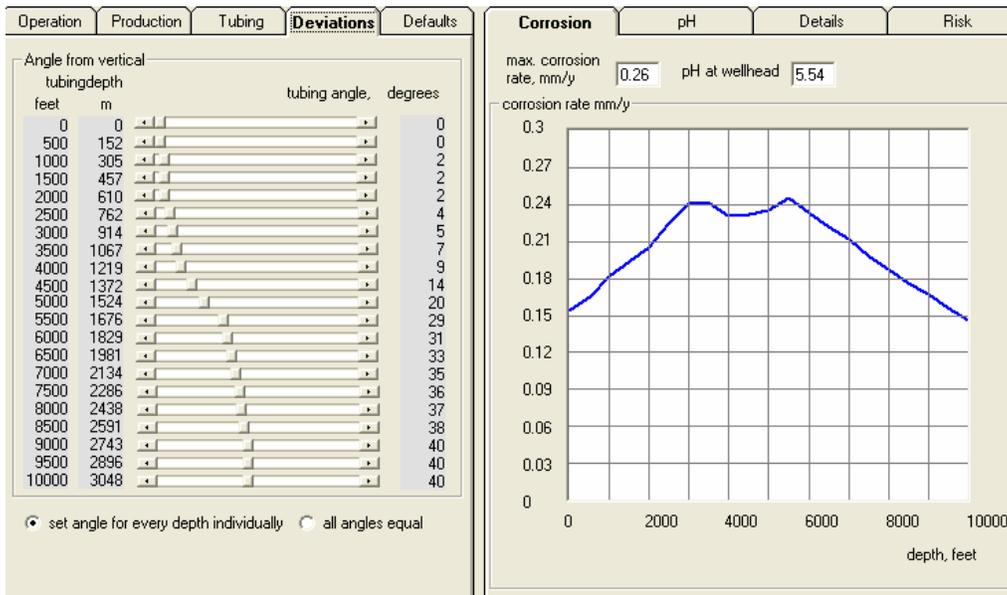


Figure 5. Example of in- and output of the model, for production tubing, with input for angle of deviation.

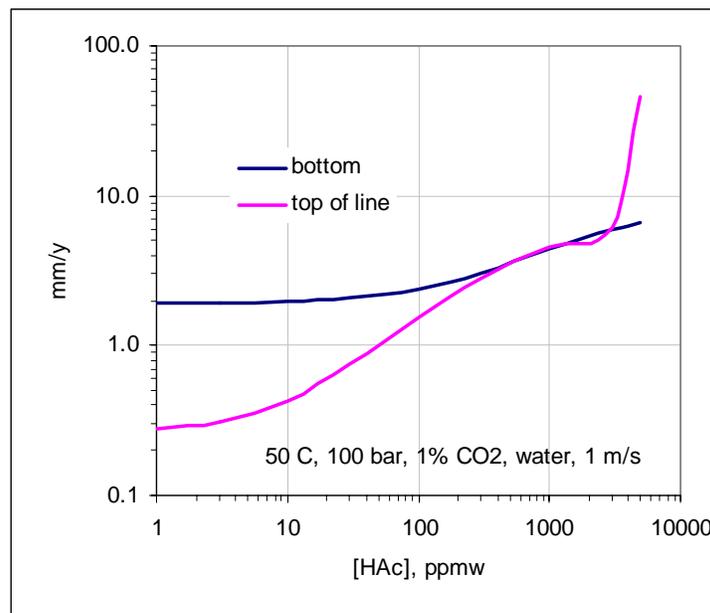


Figure 6. Example of effect of HAc concentration. Conditions as indicated on the figure.

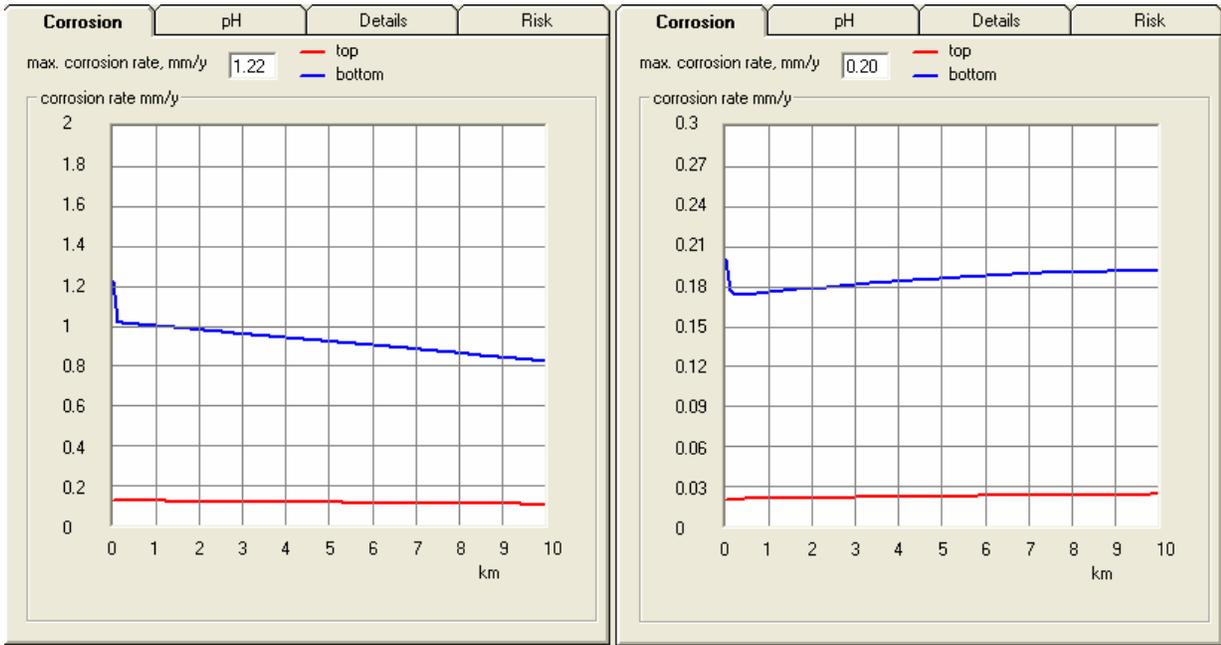


Figure 7. Example of effect of glycol injection. Conditions: pressure 70-60 bar, temperature 50-40 C, 0.8% CO₂, 12" pipe, 100 m³/d condensate, 1.3 MMsm³/d gas, left: without glycol, right: with 2000 kg/MMsm³ of glycol.

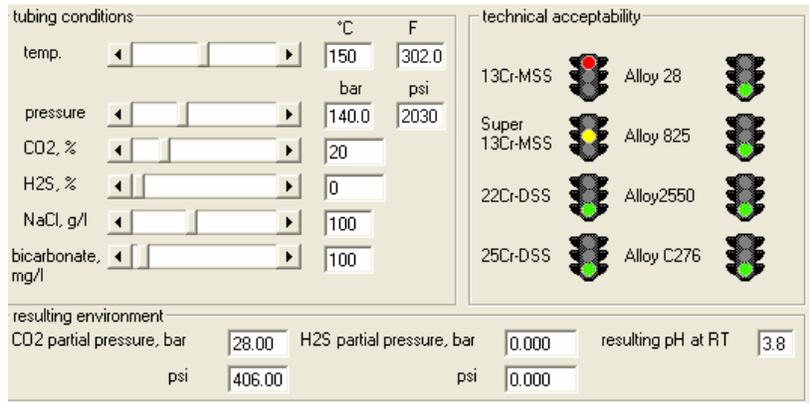


Fig.8. Example CRA evaluation for downhole tubing. The traffic lights indicate standard 13Cr martensitic stainless steel is unsuitable, Super 13Cr is only marginal and all other alloys are acceptable for the stated conditions

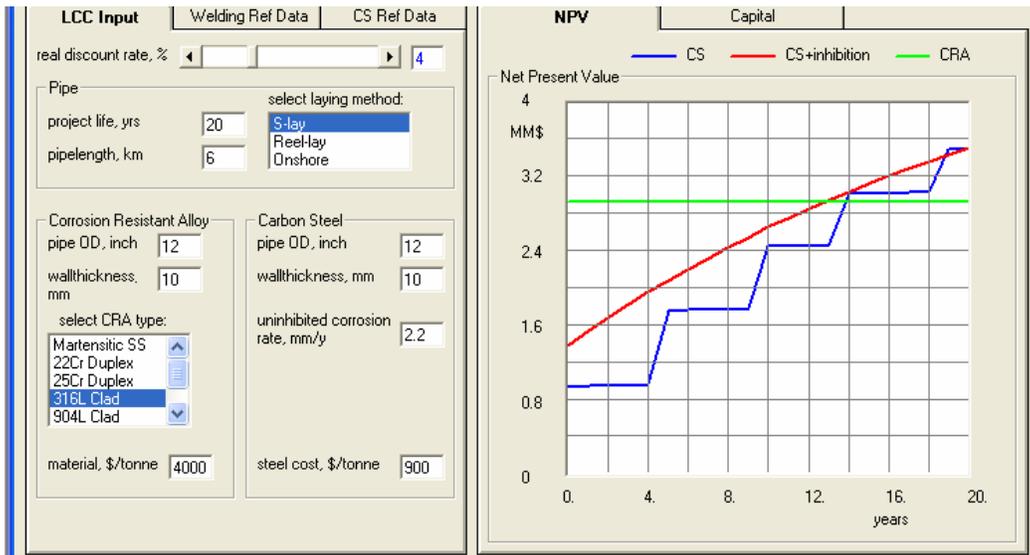


Fig.9. Life-cycle cost example for a 12 inch flowline of 6km length. Carbon steel would require several replacements during the project life unless protected by inhibition. For lifetimes beyond approximately 13 years a 316L clad pipe is the cheapest solution for the specific input details.